



CHALMERS
UNIVERSITY OF TECHNOLOGY



An Evaluation of the Performance of the GoBiGas Gasification Process

Master's Thesis within the *Sustainable Energy Systems* programme

MALIN KARLBRINK

MASTER'S THESIS

An Evaluation of the Performance of the
GoBiGas Gasification Process

Master's Thesis within the *Sustainable Energy Systems* programme

MALIN KARLBRINK

SUPERVISORS

Dr. Claes Breitholtz

Dr. Anton Larsson

EXAMINER

Professor Henrik Thunman

Department of Energy and Environment

Division of Energy Technology

CHALMERS UNIVERSITY OF TECHNOLOGY

Göteborg, Sweden 2015

An Evaluation of the Performance of the GoBiGas Gasification Process
Master's Thesis within the *Sustainable Energy Systems* programme
MALIN KARLBRINK

© MALIN KARLBRINK, 2015

Department of Energy and Environment
Division of Energy Technology
Chalmers University of Technology
SE-412 96 Göteborg
Sweden
Telephone: + 46 (0)31-772 1000

Cover:
GoBiGas, photograph, viewed 31 August 2015
http://www.goteborgenergi.se/Files/Bilder/Biogas/gobigas%20bygg_h8q6586.jpg?TS=635296973031793343

Chalmers Reproservice
Göteborg, Sweden 2015

An Evaluation of the Performance of the GoBiGas Gasification Process

Master's Thesis within the *Sustainable Energy Systems* programme

MALIN KARLBRINK

Department of Energy and Environment

Division of Energy Technology

Chalmers University of Technology

ABSTRACT

The GoBiGas (Gothenburg Biomass Gasification) project was initiated by Göteborg Energi to enable biogas/bio-SNG production through gasification of biomass from forest residuals. The purpose of GoBiGas is to gain knowledge, at a pre-commercial scale, within gasification of forest residues as well as methanation for the purpose of producing biomethane of sufficient quality to be distributed in the conventional gas grid. The performance goal of GoBiGas is that the cold gas efficiency should be at least 65 % and the overall thermal efficiency should be at least 90 %. The aim of this thesis is to establish methods to verify process data and quantify the performance of the gasification process of GoBiGas. Losses in different parts of the process and potential improvements are discussed and proposed.

The process data was verified through energy and carbon balances over the gasification process and section. To include variations in the measurements, the process data was randomized with a normal distribution. Using normally distributed data implies that the results from the calculations, such as the efficiencies of the process, were normally distributed with a mean and standard deviation.

The method of comparing the normal distributions from the energy and carbon balances did not show any large errors in the measurements of the process data. The cold gas efficiency of the gasification process was within 73 – 80 % and reaching a cold gas efficiency of 65 % over the entire process is possible if less product gas is combusted in the combustion chamber. The overall thermal efficiency of the gasification process was within 80 – 86 %, and to reach an overall thermal efficiency of 90 % over the entire process, the heat losses need to be decreased. The losses in the gasification process could be decreased by; increasing the isolation of the gasification section, decreasing the temperatures in the gasification section, increasing the preheating of steam and air, and combusting less product gas.

Key words:

GoBiGas, indirect gasification, gasification, DFB, biomass, bio-SNG, biogas

TABLE OF CONTENTS

ABSTRACT.....	i
ACKNOWLEDGMENTS	v
NOMENCLATURE.....	vii
GREEK LETTERS.....	viii
SUBSCRIPTS AND SUPERSCRIPTS.....	viii
ABBREVIATIONS.....	ix
1 INTRODUCTION.....	1
1.1 THE GOBiGAS PROJECT	2
1.2 AIM	3
1.3 PROBLEM FORMULATION	3
1.4 DELIMITATIONS	3
2 THEORY	5
2.1 GASIFICATION.....	5
2.2 CARBON BALANCE	7
2.3 DEFINITION OF EFFICIENCIES.....	7
2.4 ENERGY BALANCE.....	10
2.5 CONVECTIVE AND RADIATIVE HEAT LOSSES.....	11
2.6 THE NORMAL DISTRIBUTION.....	13
2.7 ONE-AT-A-TIME SENSITIVITY ANALYSIS.....	14
3 THE GOBiGAS GASIFICATION PROCESS.....	15
4 METHODOLOGY.....	19
4.1 CALCULATION PROCEDURE	19
4.2 CALCULATION OF $\eta_{CH_4,max}$	19
4.3 GASIFICATION PROCESS ENERGY AND CARBON BALANCE	20
4.4 GASIFICATION SECTION ENERGY AND CARBON BALANCE.....	21
4.5 GASIFIER ENERGY AND CARBON BALANCE.....	21
4.6 SENSITIVITY ANALYSIS	22
4.7 HEAT LOSSES	23
4.8 BIOMASS FLOW	24
5 EXPERIMENTS	25
6 RESULTS	27
6.1 ENERGY AND CARBON BALANCE OVER THE GASIFICATION PROCESS	27
6.2 ENERGY AND CARBON BALANCE OVER THE GASIFICATION SECTION.....	31
6.3 ENERGY AND CARBON BALANCE OVER THE GASIFIER.....	33
7 DISCUSSION	35
8 CONCLUSIONS.....	37
9 REFERENCES.....	39
APPENDIX.....	I
A CALCULATION OF ENERGY FLOWS	I
B INSTALLED EFFECT OF LARGEST ELECTRICITY CONSUMERS.....	II
C CONVECTIVE HEAT TRANSFER COEFFICIENT CALCULATION.....	III
D RADIATIVE AND CONVECTIVE HEAT LOSSES	IV

E	ONE-AT-A-TIME SENSITIVITY ANALYSIS RESULTS FOR C2.....	VI
---	--	----

ACKNOWLEDGMENTS

First of all I would like to thank Claes Breitholtz, at Valmet, for giving me the opportunity of doing this thesis and for all of your support throughout the process. Thank you for not answering all of my questions and instead taking the time to give me the tools to answer them myself.

Anton Larsson, thank you for all your advice and help in finishing this thesis. Thank you for all the times you helped me export data when I sent you the wrong tag names.

I would like to thank my examiner, Henrik Thunman, for the opportunity of continuing working on the thesis during the summer at Chalmers. Thank you for sharing your ideas and for helping me to improve this thesis.

I would also like to thank the gasification group at Chalmers, the staff of GoBiGas, the Swedish Gasification Center (SFC), and the Center for Indirect Gasification of Biomass (CIGB).

NOMENCLATURE

Δh_R	Enthalpy of reaction [kJ/kg]
A	Area [m ²]
C_p	Specific heat capacity [kJ/kgK]
D	Diameter [m]
f	Probability density function
Gr	Grashof number [-]
g	Gravitational acceleration [m/s ²]
h	Convective heat transfer coefficient [W/m ² K]
h	Enthalpy [kJ/kg]
$h_{calcination}$	Heat of calcination (-3025 kJ/kg)
h_f^o	Enthalpy of formation [kJ/kg]
k	Thermal conductivity [W/mK]
L	Length [m]
LHV	Lower heating value [MJ/kg]
M	Molar weight [kg/mol]
$\dot{m}$	Mass flow [kg/s]
Nu	Nusselt number [-]
$\dot{n}$	Molar flow [mol/s]
P^o	Standard Pressure (1 bar)
Pr	Prandtl number [-]
$\dot{Q}$	Energy flow [W]
q_{conv}	Rate of convective heat transfer [W]
q_{rad}	Net radiation heat loss [W]
T	Temperature [°C] or [K]
T_C	Temperature of cold surrounding [K]
T_H	Temperature of hot body [K]
T_{ref}	Reference temperature (25 °C)
$\dot{V}$	Volumetric flow [m ³ /h]
Y	Mass fraction [-]
X	Random variable

X_{comb}	Fraction of combusted char [-]
X_{gas}	Fraction of gasified char [-]

GREEK LETTERS

β	Coefficient of thermal expansion [m^{-1}]
γ	Fraction of the percentile change of the output and of the variable in the OAT sensitivity analysis [-]
ϵ	Emissivity coefficient [-]
η_{cg}	Cold gas efficiency [-]
$\eta_{cg,G}$	Gasifier cold gas efficiency [-]
η_i	Normalized energy flow [-]
$\eta_{CH_4,max}$	Maximum conversion of product gas to methane [-]
η_o	Overall thermal efficiency [-]
μ	Viscosity [kg/ms]
μ	Mean value
ν	Normalized carbon flow [-]
ρ	Density [kg/m^3]
σ	Stefan-Boltzman constant ($5.676 \cdot 10^{-8} W/m^2K^4$)
σ	Standard deviation

SUBSCRIPTS AND SUPERSSCRIPTS

bio	Biomass
BDW	Blow down water
BFW	Boiler feed water
CW	Cooling water
DHW	District heating water
el	Electricity
FG	Flue gas
HL	Convective and radiative heat losses from the gasification process
HL,G	Convective and radiative heat losses from the gasifier
HW	Hot water

i	Flow i
j	Species j
off	Gases from the methanation process
PG	Product gas from the gasification process
PG,G	Product gas from the gasification process added to the recirculated product gas
PG,rec	Recirculated product gas
prod	Products
reac	Reactants

ABBREVIATIONS

BFB	Bubbling fluidized bed
BFW	Boiler feed water
C	Elemental carbon
CaCO ₃	Limestone molecule
CFB	Circulating fluidized bed
CH ₄	Methane molecule
C ₂ H ₄	Ethylene molecule
C _x H _y	Hydrocarbon
CO	Carbon monoxide molecule
CO ₂	Carbon dioxide molecule
DFB	Dual fluidized bed
EU	European Union
FG	Flue gas
GHG	Greenhouse gas
H ₂	Hydrogen molecule
H ₂ O	Water molecule
HHV	Higher heating value
LHV	Lower heating value
N ₂	Nitrogen molecule
NH ₃	Ammonia molecule
NO _x	Mono-nitrogen oxides
N ₂ O	Nitrous oxide molecule

O ₂	Oxygen molecule
OAT	One-at-a-time
PG	Product gas
RME	Rapeseed methyl ester
SNG	Substitute natural gas

1 INTRODUCTION

In 2012, the transport sector was responsible for around 25% of the European Union's (EU) total greenhouse gas (GHG) emissions and 70% of the EU's GHG emissions from the transport sector were from road transport. Out of EU's total carbon dioxide emissions, the road transport was responsible for around 20%. (European Commission, 2015) The EU have set up several policies aiming to lower GHG emissions from the transport sector, which is one of the most important drivers to promote biofuels. (Biofuels Research Advisory Council, 2006)

Figure 1.1 shows the usage of biofuels in the Swedish transport sector between the years 2000-2012. Between 2011 and 2012, the usage of biofuels increased with 17% and biogas stood for 12% of the biofuel usage. There are two clear trends for biogas in the Sweden: the biogas production increases and the fraction of the biogas that is upgraded increases (50% in 2011). During the gas upgrade the methane content in the biogas increases from about 60% to above 90%, which is necessary to be able to use the biogas as a vehicle fuel. (Energimyndigheten, 2013)

One of several biofuels available is substitute natural gas (SNG), which is produced through gasification of biomass. Biomass is regarded as a renewable energy carrier, since it is carbon neutral, meaning that the carbon dioxide released during thermal utilization is absorbed via photosynthesis during the growth of the biomass. (Gröbl T, Walter H, & Haider M, 2012) Bio-SNG is categorized as a 2nd generation biofuel and the biomass feedstock used in producing 2nd generation biofuels is dominated by lignocellulosic residues from forest and agriculture, such as tree barks, tops and branches, demolition wood, tall grasses, and crop residues. The main difference between 2nd generation fuels from 1st generation, is that the biomass can be harvested with much less interference to the food economy. The technologies in converting biomass into 2nd generation biofuels are still under development (Zhang W, 2010), and one project with the purpose of gaining knowledge within production of bio-SNG through gasification is the GoBiGas project.

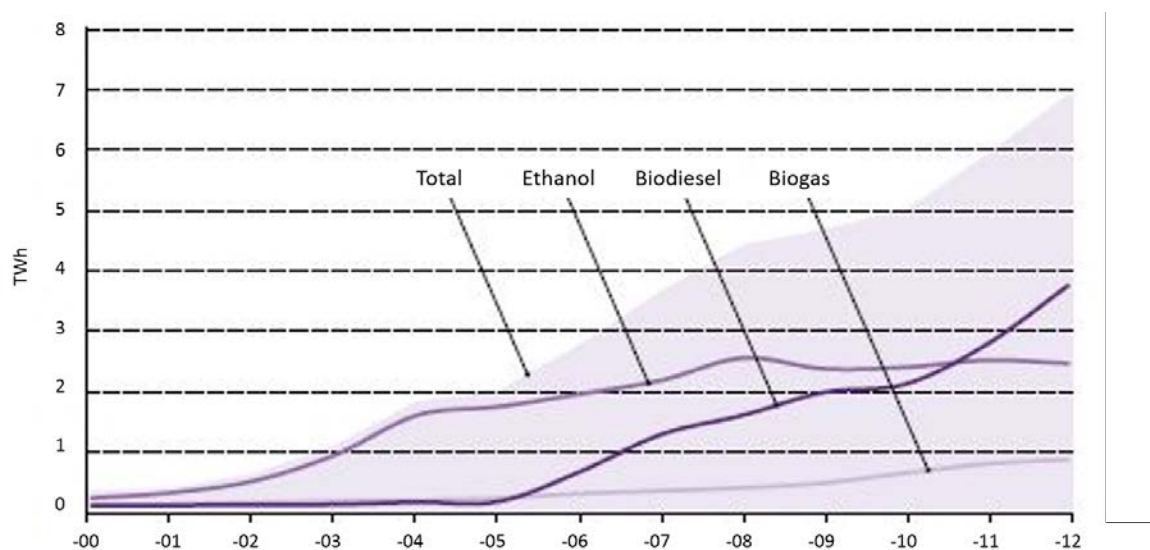


Figure 1.1: Usage of biofuels in the Swedish transport sector between the years 2000-2012 (Energimyndigheten, 2013)

1.1 THE GoBiGAS PROJECT

The GoBiGas (Gothenburg Biomass Gasification) project was initiated by Göteborg Energi to enable biogas/bio-SNG production through gasification of biomass from forest residuals. The process consists of a gasification and methanation process and shared help systems. Valmet has, together with Repotec, supplied the gasification process, which will be more thoroughly described in Chapter 3. In the gasification process the biomass is gasified and converted into syngas that is cleaned and upgraded into bio-SNG (95% methane) in the methanation process. After the methanation the bio-SNG is injected into the existing natural gas network. (Göteborg Energi, 2015)

Gasification and methanation of forest raw material has never been tested in practice in a full scale operating plant before. The purpose of GoBiGas is to gain knowledge, at a pre-commercial scale, within gasification of forest residues as well as methanation for the purpose of producing biomethane of sufficient quality to be distributed in the conventional gas grid. (Andersson L, 2015)

The project consists of two phases, where the first phase is a demonstration plant that supplies 20 MW of gas. The second phase will be a commercial plant with a planned gas production of 80-100 MW which will be able to supply 80 000 – 100 000 cars with fuel. (Göteborg Energi, 2015) The investment decision for the second phase is put on hold until the first phase has proven successful (Andersson L, 2015).

The performance goal of GoBiGas is that at least 65 % of the energy content in the biomass is converted into bio-SNG ($\eta_{cg,tot}$) and that the total energy efficiency of the process ($\eta_{o,tot}$) shall be at least 90 %. (Göteborg Energi, 2015)

The overall objective of the first phase of the GoBiGas project is to verify:

- The efficiency of conversion from biomass to biomethane and the total energy-efficiency of the process
- Product quality
- Environmental performance
- Fuel flexibility

The purpose of this thesis falls into the first objective above. It will only include the drying, gasification, and gas cleaning part of the general bio-SNG production by thermo-chemical gasification process illustrated in Figure 1.2.

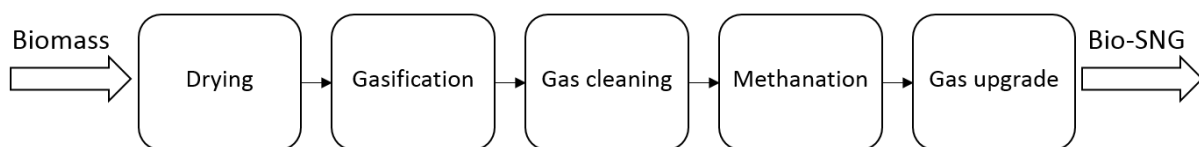


Figure 1.2: General process for bio-SNG production by thermo-chemical gasification.

1.2 AIM

The aim of this Master's Thesis is to establish methods to verify process data and quantify the performance of the GoBiGas gasification process. Losses in different parts of the process and potential improvements will be discussed and proposed.

1.3 PROBLEM FORMULATION

The gasification and gas cleaning technology used in GoBiGas is rather new as industrial process. To be able to improve the process, the performance and the losses in the process should be evaluated and quantified. The evaluation of the process performance will enable an investigation of where the losses in the process occur and how improvements to the process can be made.

1.4 DELIMITATIONS

This Master's Thesis will focus on the evaluation of the gasification section including the gas cleaning system and not the methanation section of the GoBiGas process.

The carbon and energy balances will only be based on available process data, assumptions and temporary measurement points.

The bed material and ashes were excluded from the carbon balances, since only reacting material flows are included.

Hydrogen and oxygen balances were not set up because of uncertainties in the measurements of e.g. water in the product and flue gas.

2 THEORY

Energy and carbon balances are set up to verify the measured process data used to calculate the efficiencies of the gasification process and the gasifier. In this chapter the definitions of the different efficiencies studied in this thesis are presented and the theory behind the energy and carbon balances are explained. Indirect gasification and biomass conversion is generally described as an introduction to Chapter 3, where the GoBiGas gasification process is described more thoroughly.

2.1 GASIFICATION

Gasification and combustion are two closely related thermochemical processes. Gasification packs energy into chemical bonds in the product gas, while combustion breaks the bonds to release the energy. The gasification process strips carbon away and adds hydrogen to the feedstock to produce gases with a higher hydrogen-to-carbon (H/C) ratio. Combustion on the other hand oxidizes the hydrogen and carbon into water and carbon dioxide, respectively. (Basu P, 2010)

There are two types of gasification: direct or indirect gasification, depending on the way the heat for the gasification is provided to the gasifier. In the direct gasification the heat is released by partial oxidation of the fuel in the gasifier itself, whereas indirect gasification obtains the heat necessary for gasification from a source outside of the gasifier. (Gómez-Barea A & Leckner B, 2010) In indirect gasification the product gas has a very low nitrogen content, since the gasification reactor is fluidized with steam and not air. This makes it suitable for synthesis of biofuels (Heyne S, 2013). An example of an indirect gasifier is the dual fluidized bed (DFB) gasifier. Figure 2.1 illustrates a simplification of the DFB process,

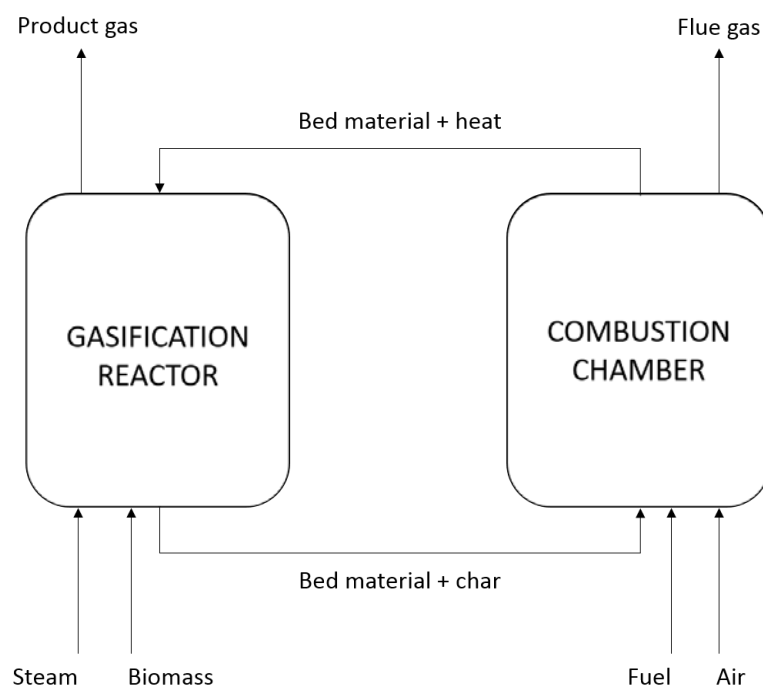


Figure 2.1: Principle of the Dual Fluidized Bed process

which is basically composed of two interconnected reactors. The heat necessary in the endothermic bubbling fluidized bed (BFB) gasifier is generated in the exothermic circulating fluidized bed (CFB) combustor and transferred with the bed material.

In general, biomass conversion includes three steps:

- Drying
- Pyrolysis/devolatilization
- Char conversion (gasification or combustion)

The biomass is fed into the gasifier and after it is dried, the temperature rises and the biomass begins to decompose releasing volatiles, which are defined as the part of the fuel that can be converted to gas with heat as the only driving force. Devolatilization is the oxidative decomposition of biomass and pyrolysis is the non-oxidative decomposition. Pyrolysis can be divided into two steps; in the first step biomass is thermally decomposed into gases (CO , CO_2 , CH_4 , H_2 and H_2O), tar and char. In the second step, tar cracking takes place and the products are composed of CO , CO_2 , H_2 , heavier hydrocarbons and inert tar. (Nguyen T B.D, Ngo S I, Lim Y, Lee J W, Lee U, & Song B, 2012) After the pyrolysis a fraction of the char is gasified through steam gasification.

The unconverted fraction of the char is transported with the bed material to the combustion chamber, where it is combusted in the presence of air. Combustion in the combustion chamber increases the flexibility of the system, since it allows combustion of by-products and/or auxiliary fuel, if additional heat is required in the gasifier or other parts of the process. (Larsson, 2014)

The char, which mainly consists of carbon, is converted in the presence of a reactant (e.g. H_2O , CO_2 or O_2). If the reactant is H_2O or CO_2 the char is gasified and if the reactant is O_2 the char is combusted. (Larsson A, Seeman M, Neves D, & Thunman H, 2013)

The char is converted through oxidation by O_2 through:



The char is converted through gasification by H_2O and CO_2 through:



Within the gasifier, the syngas ($\text{CO} + \text{H}_2$) can react and produce methane and hydrocarbons in the product gas can react and produce syngas:



In an ideal case without supplementary fuels, the gasification will take place according to reaction R3 and R5 and the combustion according to reaction R1. The char combustion will generate exactly the heat necessary for the drying of the biomass, devolatilization and the char gasification. This results in the following equation,

$$0 = \Delta h_{drying} + \Delta h_{devolutilization} + X_{gas}^* \Delta h_{char\ gasification} + X_{comb}^* \Delta h_{char\ combustion} \quad (2.1)$$

where X_{comb}^* is the minimum mass fraction of the char that is combusted to generate the heat demand and X_{gas}^* is the maximum mass fraction of the char that is gasified ($X_{gas}^* = 1 - X_{comb}^*$). A similar calculation was done by Larsson, where X_{gas}^* was calculated to $\approx 75\%$ for any fuel. (Larsson A, 2014)

2.2 CARBON BALANCE

A mass balance is an application of conservation of mass, which states that matter within an isolated system is conserved and cannot be created or destroyed. This implies that the molar flow of an element entering the system is equal to the molar flow leaving the system.

$$\sum \dot{n}_{j,in} = \sum \dot{n}_{j,out} \quad (2.2)$$

$\dot{n}$ is the molar flow, in mol/s , and j is an element within the system, e.g. C, H, or O.

An example of a carbon balance for a fuel (containing C, H and O) where the carbon is converted into CO, CO₂, and CH₄, is presented below.

$$\sum \dot{n}_{C,in} = \sum \dot{n}_{C,out} \rightarrow \dot{n}_C = \dot{n}_{CO} + \dot{n}_{CO_2} + \dot{n}_{CH_4}$$

The normalized carbon flows ($v_{C,i}$) are defined as the fraction between the carbon flow, i , and the total carbon flow entering the gasification process,

$$v_{C,i} = \frac{\dot{n}_{C,i}}{\dot{n}_{C,in}} \quad (2.3)$$

and since $\dot{n}_{C,in} = \dot{n}_{C,out}$,

$$\sum_i \frac{\dot{n}_{C,in,i}}{\dot{n}_{C,in}} = \sum_i \frac{\dot{n}_{C,out,i}}{\dot{n}_{C,in}} = 1$$

2.3 DEFINITION OF EFFICIENCIES

The two efficiencies used to evaluate the gasification process are the cold gas efficiency (η_{cg}) and the overall thermal efficiency (η_o). In Figure 2.2 the energy flows entering and leaving the gasification process are presented. These flows are needed to calculate the efficiencies of the gasification process.

The total energy flow entering and leaving the gasification process is calculated as follows.

$$\dot{Q}_{in} = \sum_j \dot{Q}_{in,i} = \dot{Q}_{bio} + \dot{Q}_{RME} + \dot{Q}_{off} + \dot{Q}_{CaCO_3} + \dot{Q}_{BFW} + \dot{Q}_{el} \quad (2.4)$$

$$\dot{Q}_{out} = \sum_j \dot{Q}_{out,i} = \dot{Q}_{PG} + \dot{Q}_{FG} + \dot{Q}_{CW} + \dot{Q}_{HW} + \dot{Q}_{DHW} + \dot{Q}_{BDW} + \dot{Q}_{HL} \quad (2.5)$$

The description of the different flows are presented in Table 2.1 and the calculations of the energy flows are presented in Table A.1 in Appendix.

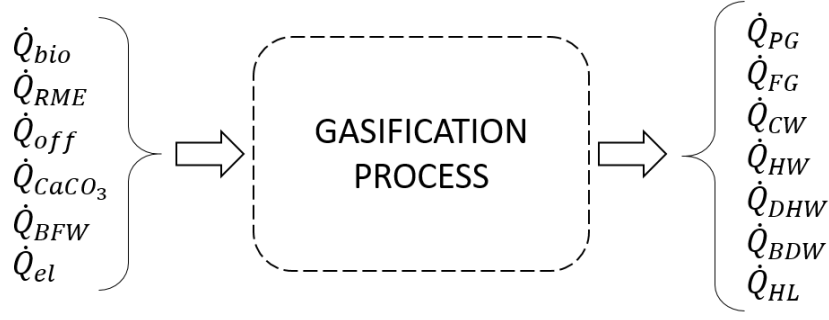


Figure 2.2: Energy flows included in the energy balance over the gasification process.

The cold gas efficiency is defined as the fraction of the chemically bound energy in the biomass that is converted into chemically bound energy in the product gas, from the gasification process, and is calculated as follows,

$$\eta_{cg} = \frac{\dot{m}_{PG} LHV_{PG}}{\dot{m}_{bio} LHV_{bio}} \quad (2.6)$$

where $\dot{m}_{PG}$ and $\dot{m}_{bio}$ is the mass flow of product gas and biomass respectively, in kg/s , and LHV_{PG} and LHV_{bio} is the lower heating value of the product gas and the as received biomass respectively, in MJ/kg .

To be able to compare η_{cg} with the performance goal of GoBiGas, stating that at least 65 % of the energy content in the biomass is converted into bio-SNG, $\eta_{cg,tot}$ needs to be approximated. This can be done by calculating the maximum conversion of the product gas to methane, $\eta_{CH_4,max}$.

$$\eta_{CH_4,max} = \frac{n_{CH_4,max} M_{CH_4} LHV_{CH_4}}{\sum n_j M_j LHV_j} \quad (2.7)$$

Where $n_{CH_4,max}$ is the maximum molar amount of methane that can be produced from the product gas, including the methane already in the product gas. n_j is the molar amount of species j in the product gas, and M is the molar mass, in kg/mol .

The maximum $\eta_{cg,tot}$ is calculated as follows:

$$\eta_{cg,tot} = \eta_{CH_4,max} \eta_{cg} \quad (2.8)$$

The normalized energy flows (η_i) are defined as the fraction between the energy flow, i , and the energy entering the gasification process,

$$\eta_i = \frac{\dot{Q}_i}{\dot{Q}_{in}} \quad (2.9)$$

and since $\dot{Q}_{in} = \dot{Q}_{out}$,

$$\sum_i \frac{\dot{Q}_{in,i}}{\dot{Q}_{in}} = \sum_i \frac{\dot{Q}_{out,i}}{\dot{Q}_{in}} = 1$$

The overall thermal efficiency is defined as the fraction of the energy entering the gasification process that is converted into useful products and is calculated as follows,

$$\eta_o = \frac{\dot{Q}_{PG} + \dot{Q}_{HW} + \dot{Q}_{DHW}}{\dot{Q}_{in}} = \eta_{PG} + \eta_{HW} + \eta_{DHW} \quad (2.10)$$

where the useful products are the product gas, hot water used in the methanation process and district heating system, and the district heating water.

The fraction of the energy content entering the gasification process that isn't converted into useful products, and are regarded as losses, can be calculated as follows,

$$\eta_{losses} = 1 - \eta_o = \eta_{FG} + \eta_{CW} + \eta_{BDW} + \eta_{HL} \quad (2.11)$$

In Figure 2.3 the energy flows entering and leaving the gasifier are presented. These flows are used to calculate the efficiencies of the gasifier.

The total energy flow entering and leaving the gasifier is calculated as following,

$$\dot{Q}_{in,G} = \sum_i \dot{Q}_{in,G,i} = \dot{Q}_{bio} + \dot{Q}_{heat} + \dot{Q}_{steam} \quad (2.12)$$

$$\dot{Q}_{out,G} = \sum_i \dot{Q}_{out,G,i} = \dot{Q}_{PG,G} + \dot{Q}_{char} + \dot{Q}_{HL,G} \quad (2.13)$$

and the description of the different flows are presented in Table 2.1.

The definition of the cold gas efficiency over the gasifier is the same as over the gasification process, however, the chemically bound energy flow in the product gas is that leaving the gasification process and the product gas that is recirculated to the combustion chamber.

$$\eta_{cg,G} = \frac{\dot{m}_{PG,G} LHV_{PG,G}}{\dot{m}_{bio} LHV_{bio}} \quad (2.14)$$

The performance of the gasifier is also evaluated through the mass fraction of the char that is combusted (X_{comb}).

$$X_{comb} = \frac{\dot{n}_{char}}{\dot{n}_{char,biomass}} \quad (2.15)$$

Where $\dot{n}_{char}$ and $\dot{n}_{char,biomass}$ is the molar flow of char combusted in the combustion chamber and the total molar flow of char in the biomass respectively, in mol/s .

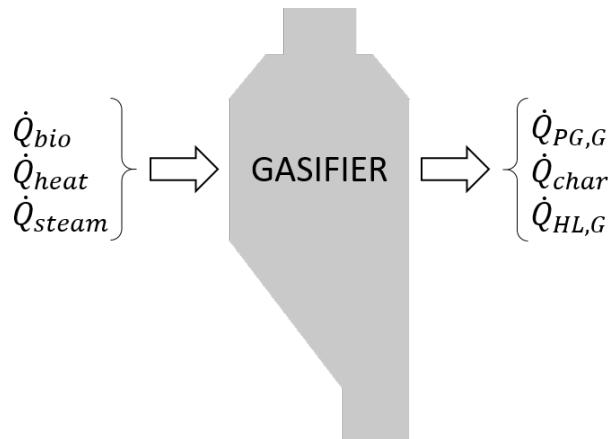


Figure 2.3: Energy flows included in the energy balance over the gasifier.

Table 2.1: Description of energy flows.

Description of energy flows	
$\dot{Q}_{BDW}$	Blow down water
$\dot{Q}_{BFW}$	Boiler feed water
$\dot{Q}_{bio}$	Biomass
$\dot{Q}_{CaCO_3}$	CaCO ₃
$\dot{Q}_{char}$	Ungasified char
$\dot{Q}_{CW}$	Cooling water
$\dot{Q}_{DHW}$	District heating water
$\dot{Q}_{el}$	Electricity
$\dot{Q}_{FG}$	Flue gas
$\dot{Q}_{heat}$	Heat from the combustion chamber
$\dot{Q}_{HL}$	Convective and radiative heat losses in the gasification process
$\dot{Q}_{HL,G}$	Convective and radiative heat losses in the gasifier
$\dot{Q}_{HW}$	Hot water
$\dot{Q}_{off}$	Gases from the methanation process
$\dot{Q}_{PG}$	Product gas leaving the gasification process
$\dot{Q}_{PG,G}$	Product gas leaving the gasifier
$\dot{Q}_{RME}$	RME
$\dot{Q}_{steam}$	Steam

2.4 ENERGY BALANCE

The energy balance is based on the postulate of conservation of energy in the universe, also known as the “First Law of Thermodynamics”. As explained by Elliott & Lira, it states that the total energy of an isolated system remains constant and, therefore, energy cannot be created or destroyed, only change form. This implies that the energy entering a system is equal to the energy leaving the system.

$$\dot{Q}_{in} = \sum_i \dot{Q}_{in,i} = \sum_i \dot{Q}_{out,i} = \dot{Q}_{out} \quad (2.16)$$

The energy flow, $\dot{Q}_i$, can be calculated by multiplying the enthalpy (h) with the mass flow ($\dot{m}$).

$$\dot{Q}_i = \dot{m}_i h_i = \dot{m}_i (h_{f,i}^o + \int_{T_{ref}}^T C_{p,i} dT) \quad (2.17)$$

The enthalpy of formation (h_f^o) is the energy associated with chemical bonds, in kJ/kg , the specific heat capacity (C_p) is the energy required to raise the temperature of a substance by 1°C, in kJ/kgK , and T_{ref} is a reference temperature (25 °C).

In this thesis, the fuel properties are calculated on as received bases (including moisture and ash). When calculating the energy content of fuels it is convenient to use the fuels lower heating value, which is derived from the enthalpy of reaction. The enthalpy of reaction (Δh_R), at a reference temperature, is the change in energy or heat content (per unit mass)

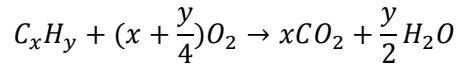
associated with the reaction. It is defined as the enthalpy difference of the products and the reactants. If $\Delta h_R < 0$ energy is released from the reaction and if $\Delta h_R > 0$ energy is required.

$$\Delta h_R = h_{prod} - h_{reac} \quad (2.18)$$

The heat of combustion (or heating value) is the same as the enthalpy of reaction, but with opposite sign. The heating value is the heat release per unit mass when the fuel, initially at a reference temperature (T_{ref}), reacts completely with oxygen and the products are returned to the reference temperature. The higher heating value (HHV) is the heat of combustion assuming that the water in the products is condensed and the lower heating value (LHV) is when the water is not condensed. (Turns S R, 1996)

$$LHV = -\Delta h_R = h_{reac} - h_{prod} \quad (2.19)$$

The products from a complete stoichiometric combustion of a hydrocarbon fuel (C_xH_y) is carbon dioxide and water, since all the fuel carbon is converted to carbon dioxide and all the fuel hydrogen is converted into water.



The enthalpy of the reactants is the sum of the enthalpy of formation at standard reference state (T_{ref}, P^o) of the reactants. Since the enthalpy of formation for elements in their naturally occurring state is equal to zero ($h_{f,O_2}^o = 0$), the enthalpy of the reactants in this example is equal to the enthalpy of formation of the fuel.

$$h_{reac} = h_{f,fuel}^o(T_{ref})$$

The enthalpy of the products is the sum of the enthalpy of formation of the reactants multiplied with the mass fraction of the different species (Y_j).

$$h_{prod} = Y_c h_{f,CO_2}^o(T_{ref}) + Y_{H_2} h_{f,H_2O(g)}^o(T_{ref})$$

By combining these two expressions the LHV of the fuel becomes,

$$LHV_{fuel} = h_{reac} - h_{prod} = h_{f,fuel}^o(T_{ref}) - (Y_c h_{f,CO_2}^o(T_{ref}) + Y_{H_2} h_{f,H_2O(g)}^o(T_{ref}))$$

and the energy flow of the fuel can be calculated as follows,

$$\dot{Q}_{fuel} = \dot{m}_{fuel} \left(LHV_{fuel} + \int_{T_{ref}}^{T_{fuel}} C_{p_{fuel}} dT \right)$$

2.5 CONVECTIVE AND RADIATIVE HEAT LOSSES

The heat losses in the system are mainly losses due to radiation, natural convection and heat losses associated with cooling, which isn't converted into district heating or utilized within the process. Equations used to calculate convective and radiative heat losses are presented in this section.

Convection is the heat transfer between a surface and an adjacent fluid. If the motion of the adjacent fluid past a solid surface is generated by an external source, such as a fan or pump, forced convection occurs. In natural convection, the fluid motion adjacent to the solid surface

is caused by the density difference due to temperature gradients in the fluid. The rate of convective heat transfer can be expressed as

$$q_{conv} = hA\Delta T \quad (2.20)$$

where h is the convective heat transfer coefficient, in W/m^2K , A is the area normal to the direction of heat flow, in m^2 , and ΔT is the temperature difference between surface and fluid, in K .

The convective heat transfer coefficient varies with the geometry of every system and for natural convection in air it takes on values of approximately 5 – 50 W/m^2K . The convective heat transfer coefficient can be calculated with the help of correlations of experimental heat-transfer data and the definition of the Nusselt number.

$$Nu \equiv \frac{hL}{k} \quad (2.21)$$

L is the length of the system, in m , and k is the thermal conductivity, in W/mK .

Natural convection adjacent to vertical cylinders can be evaluated using correlations for vertical plane if the following criterion is fulfilled.

$$\frac{D}{L} \geq \frac{35}{Gr^{1/4}} \quad (2.22)$$

D is the diameter of the cylinder, in m , and Gr is the Grashof number.

$$Gr = \frac{\beta g \rho^2 L^3 \Delta T}{\mu^2} \quad (2.23)$$

β is the fluid coefficient of thermal expansion, in m^{-1} , g is the gravitational acceleration, in m/s^2 , ρ is the fluid density, in kg/m^3 , L is the length, in m , ΔT is the temperature difference between surface and fluid, in K , and μ is the fluid viscosity, in kg/ms .

Churchill and Chu's equation for Nu for natural convection adjacent to vertical planes, which applies to all fluids, is

$$Nu = \left(0.825 + \frac{0.387(Pr * Gr)^{1/6}}{(1 + (0.492/Pr)^{9/16})^{8/27}} \right)^2 \quad (2.24)$$

where Pr is the Prandtl number.

$$Pr = \frac{\mu C_p}{k} \quad (2.25)$$

μ is the fluid viscosity, in kg/ms , C_p is the fluid heat capacity, in J/kgK , and k is the fluid thermal conductivity, in W/mK .

By combining eq. 2.21 and 2.24 the convective heat transfer coefficient (h) can be calculated, which is used in eq. 2.20 to calculate the convective heat losses.

Thermal radiation depends on the temperature difference between the hot body and the cold surroundings and the absolute temperatures. The net radiation heat loss rate when a hot object is radiating energy to its cooler surroundings is expressed as

$$q_{rad} = \epsilon \sigma A (T_H^4 - T_C^4) \quad (2.26)$$

where ϵ is the emissivity coefficient, σ is the Stefan-Boltzmann constant ($5.676 * 10^{-8} \text{ W} / \text{m}^2 \text{K}^4$), A is the area of the emitting surface, in m^2 , T_H and T_C is the temperature of the hot body and the cold surroundings respectively, in K .

The emissivity coefficient is dependent of the material and the temperature of the radiating surface. It can take on values in the range $0 < \epsilon < 1$, and is defined as the ratio of the total emissive power of a surface to the total emissive power of an ideally radiating surface at the same temperature. (Welty J R, Wicks C E, Wilson R E, & Rorrer G L, 2008)

The heat losses, in MW , due to natural convection and radiation is calculated according to,

$$\dot{Q}_{HL} = q_{conv} + q_{rad} \quad (2.27)$$

and the normalized convective and radiative heat losses according to,

$$\eta_{HL} = \frac{\dot{Q}_{HL}}{\dot{Q}_{in}} \quad (2.28)$$

2.6 THE NORMAL DISTRIBUTION

In the calculations in this thesis, the input data was randomized with a normal distribution to take uncertainties in the measurements into account. This implies that the results from the calculations, such as the efficiencies of the process, were normally distributed. The density function of the normal distribution is presented and explained in this section.

If X is a random variable with a density function f , the probability that X falls in the interval (a, b) , for any $a < b$, is the area under the density function between a and b :

$$P(a < X < b) = \int_b^a f(x)dx \quad (2.29)$$

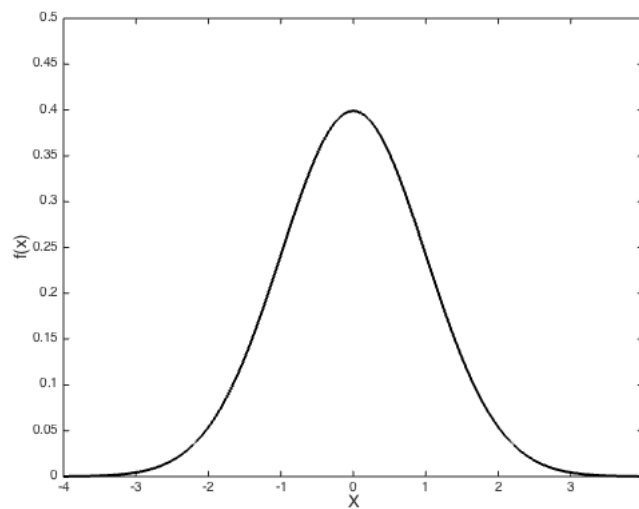


Figure 2.4: Normal density function with $\mu = 0$ and $\sigma = 1$.

The density function of the normal distribution ($f(x)$) depends on the mean (μ) and standard deviation (σ) of the normal density. (Rice, 2007) In Figure 2.4, a normal density function with $\mu = 0$ and $\sigma = 1$ is illustrated.

$$f(x) = \frac{1}{\sigma\sqrt{2\pi}} e^{-(x-\mu)^2/2\sigma^2} \quad (2.30)$$

$$-\infty < x < \infty, \quad -\infty < \mu < \infty, \quad \sigma > 0$$

2.7 ONE-AT-A-TIME SENSITIVITY ANALYSIS

Sensitivity analysis can be defined as “the varying of parameters in a simulation to discover the parameters that have the greatest influence on the features of interest” (Nicholson, 2014). One type of sensitivity analysis is the one-at-a-time (OAT) sensitivity analysis, where the impact of changing the values of each parameter is evaluated in turn. A baseline value, which is not changed throughout the analysis, is chosen for each parameter and the parameters are moved away from their baseline value one at a time, while the other parameters are kept constant at their baseline value. (Saltelli, Chan, & Scott, 2000) This approach makes it possible to evaluate the influence of every parameter on the output of the model.

3 THE GOBiGAS GASIFICATION PROCESS

The GoBiGas process consists of the gasification process, where the biomass is converted into syngas (the product gas), and the methanation process, where the syngas is cleaned and upgraded into bio-SNG. The load of the process is defined by the product gas volumetric flow from the gasification process to the methanation process (excluding the product gas combusted in the combustion chamber), where a 100% load corresponds to a product gas volumetric flow of 6900 Nm³/h.

During start-up of the entire process, the gasification process has to reach stable operation before starting up the methanation process. When the methanation process is in operation, off-gases from the methanation process are combusted in the post combustion chamber and CO₂, which is removed from the product gas in the methanation process, is used as an inert gas. Inert gas has several purposes in the process, e.g. inertizing the biomass silos, cleaning the product gas filter, and as sealing gas. During start-up, before the CO₂-removal process in the methanation process is in operation, N₂ is used as inert gas.

The GoBiGas gasification process consists of the gasification section (gasifier, combustion chamber, cyclone, and post combustion chamber) and the product and flue gas cleaning systems. A simplified scheme of the gasification process and the first step of the methanation process (tar adsorbers) can be seen in Figure 3.4. The gasification section consists of a dual fluidized bed (DFB) biomass gasifier, which is composed of the endothermic bubbling fluidized bed (BFB) gasifier and the exothermic circulating fluidized bed (CFB) combustor. A general scheme of the gasification section is illustrated in Figure 3.1.

The bed material (olivine sand) is heated up in the combustion chamber and separated from the combustion gases in the cyclone. After the cyclone the bed material is fed to the gasifier,

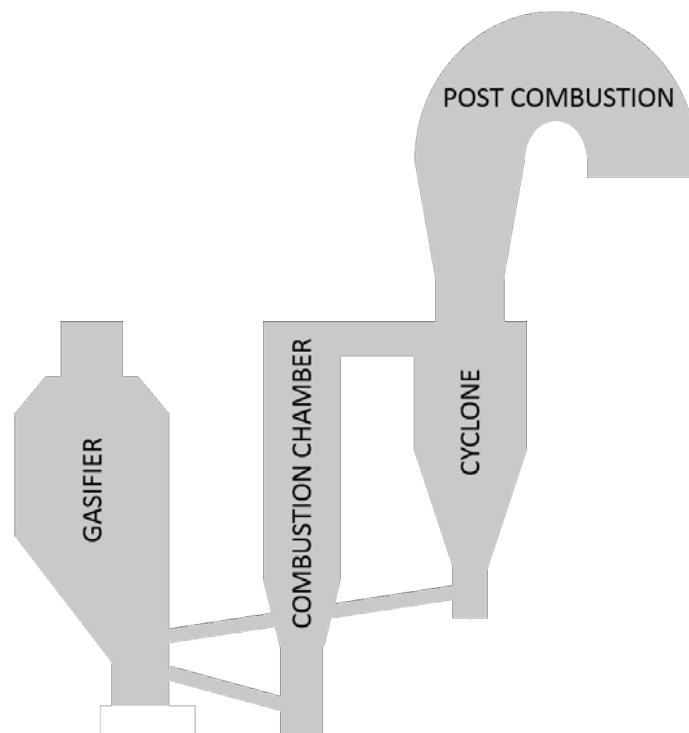


Figure 3.1: General scheme of the GoBiGas gasification section.

Table 3.1: Composition of the biomass.

Mass fraction as received fuel	
Moisture	8.1 % *
Volatiles	74.6 %
Char	17.0 % **
Ash	0.28 % *
Mass fraction dry fuel	
C	50.7 % *
H	6.1 % *
O	43.2 % *
N	< 0.1 % *

* from analysis by Belab, ** (Lundberg , Tchoffor, Pallarès , Johansson , Thunman, & Davidsson, 2015)

where the heat is released. The bed material is then transported from the gasifier back to the combustion chamber through a chute, fluidized with steam, at the bottom of the gasifier.

The wood pellets (with a composition presented in Table 3.1) are transported on a belt conveyor to the two biomass silos. While one silo is in operation and feeding pellets into the biomass dosing bin, the other silo is filled with pellets. From the dosing bin the pellets are fed, through the biomass feeding screw, into the steam fluidized gasifier where the pellets are pyrolyzed and gasified. The product gases, which mainly consist of CO, CO₂, H₂, CH₄ and water vapor, leave at the top of the gasifier.

A general scheme of the product gas cleaning and cooling system is illustrated in Figure 3.2. The product gas leaving the gasifier is cooled in the product gas cooler, which is a tube and shell heat exchanger with gas on the tube side and water on the shell side. A pre-coat material (CaCO₃) is added to the product gas after the cooler so that any condensing tar will be adsorbed on the pre-coat material and not stick to the product gas filter fabric.

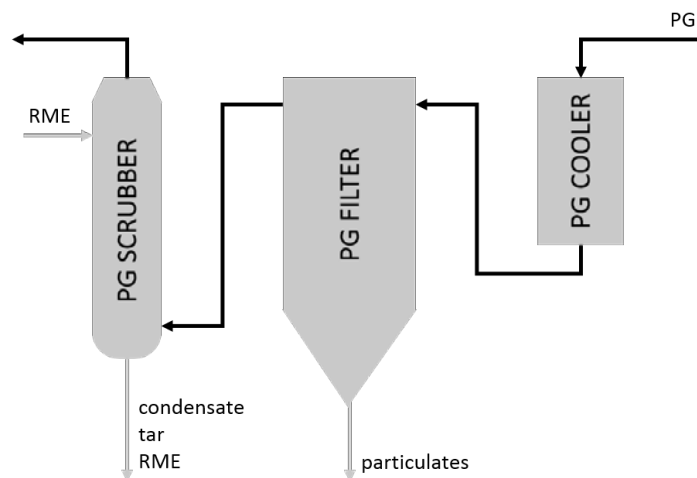


Figure 3.2: General scheme of the product gas cleaning and cooling system.

Particulates, which contain char, pre-coat material, and tar, in the product gas are removed in the product gas filter and to utilize the heating value of the ungasified carbon residues, the dust separated in the filter is transported to the combustion chamber and combusted.

In the product gas scrubber the gas is cooled, most of the water is condensed and the tar is dissolved in RME (the scrubber liquid). The product gas leaves the scrubber at the top and at the bottom the liquids are sent to the scrubber reservoir, where RME and water are separated by gravity. The pure RME stream is recycled back to the scrubber and the condensate is pumped to the condensate evaporator, where dissolved gases are driven off and sent to the post combustion chamber. A small amount of RME is combusted in the combustion chamber and replaced by new RME.

The combustion chamber is fluidized with combustion air and the main fuel used in the combustion chamber is char-coal from the gasification and recycled product gas. The combustion is operated as close to stoichiometric conditions as possible and the flue gases have high concentrations of combustible compounds (mainly CO). The flue gases are separated from the bed material in the cyclone and are combusted in the post combustion chamber together with gases from the methanation process.

A general scheme of the flue gas cleaning and cooling system is illustrated in Figure 3.3. The flue gases are cooled in the flue gas channels in flue gas coolers by heat exchange with the hot water system. In the steam superheaters superheated fluidization steam for the gasifier is produced, and in the air heaters primary and secondary combustion air is produced. In the bottom of the flue gas channels coarse ash is collected, which is recycled to the gasifier to recover catalytically active material. In the flue gas filter fine dust is separated that is sent to the flue gas ash silo for storage until transported off-site. After the flue gas fan and before the flue gas goes to the stack there are online analyzers monitoring the concentrations of O₂, CO, CO₂, H₂O, NO_x, NH₃ and N₂O. (Andersson L, 2015)

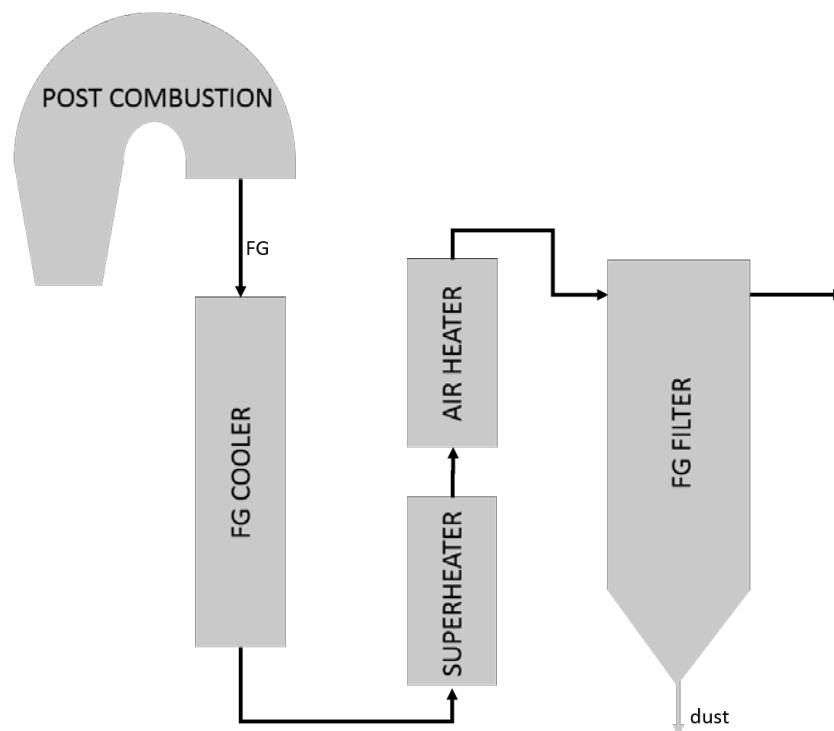


Figure 3.3: General scheme of the flue gas cleaning and cooling system.

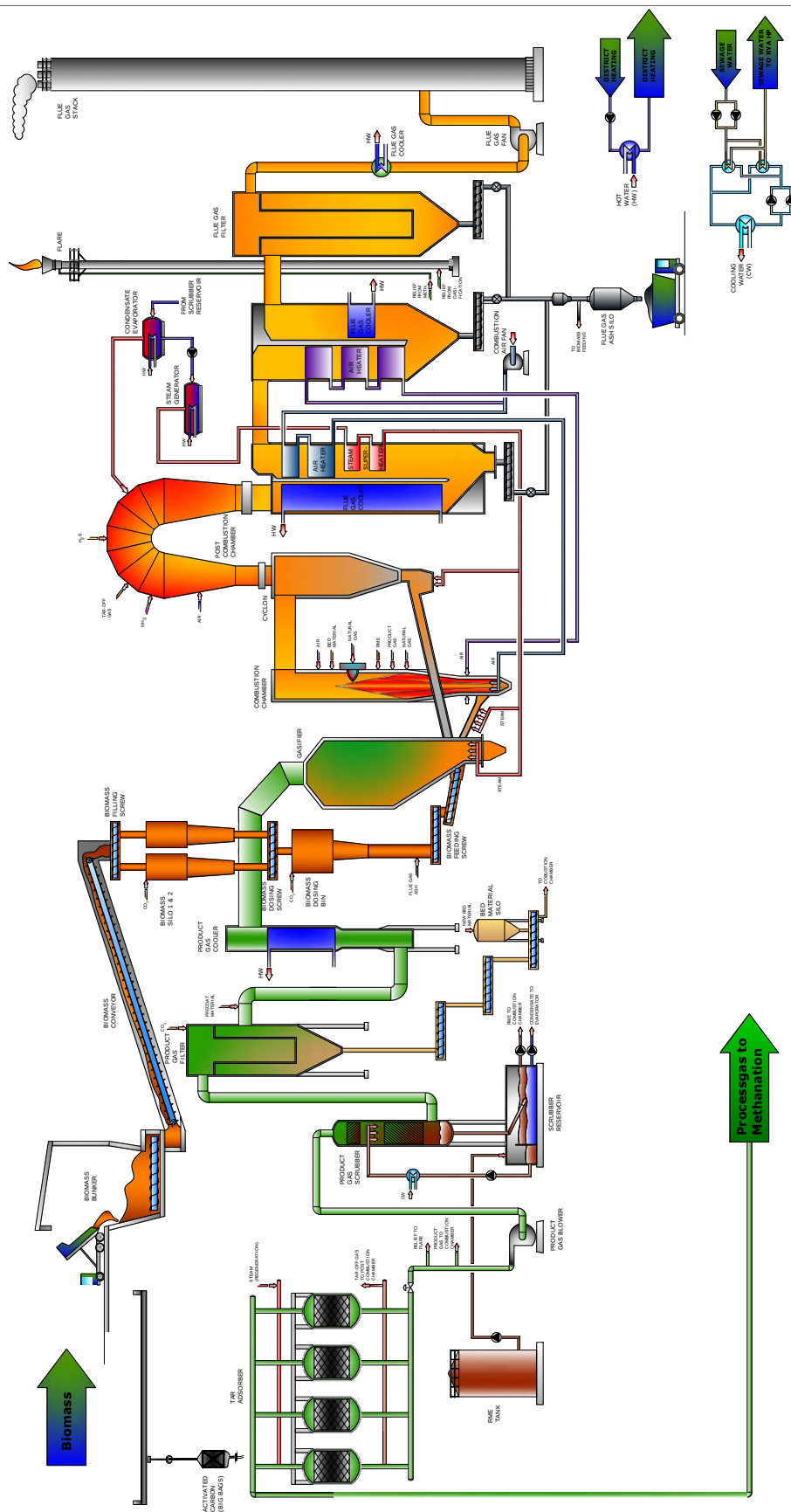


Figure 3.4: The GoBiGas gasification process and the tar adsorbers.

4 METHODOLOGY

Energy and carbon balances were set up to verify the measured and approximated data that were used to calculate the efficiencies of the process. The data was verified by checking if the energy and carbon balances were fulfilled. To be able to identify possible uncertainties and to determine over what parts of the process to set up balances, one-at-a-time sensitivity analysis was done. By setting up energy and carbon balances over different parts of the process, the uncertainties and errors in the input data was located.

The energy and carbon balances were based on available process data. A majority of the data from the process was logged continuously during operation and exported with 5-minute averages over the chosen time interval. Some data points were manually read from instruments on site during operation and others were received from analysis of test samples of different streams in the process, e.g. the product gas.

4.1 CALCULATION PROCEDURE

The process is, in some sense, over determined, since a large majority of the flows are measured in the process. To include the variations in the measured data points, the data used in the calculations was randomized with a normal distribution, using the `normrnd` function in MATLAB. The normal distribution, as explain in Section 2.6, is a function of a mean value and a standard deviation. The mean and standard deviation used in randomizing the data was calculated from the exported data. Using normally distributed data implies that the results from the calculations, such as the efficiencies of the process, were normally distributed with a mean and standard deviation.

By comparing the normal distribution of $\dot{Q}_{in}$ with $\dot{Q}_{out}$ and $\dot{n}_{C,in}$ with $\dot{n}_{C,out}$, it was possible to determine possible errors in the measurements and assumptions. From the sensitivity analysis, the variables with the largest impact on the balances could be determined. Based on these variables, new system boundaries were set up to include or eliminate the variables with the largest impact on the balances. By calculating the energy and carbon flows entering and leaving the new system boundary it is possible to determine which variables that were most likely to be wrongly measured.

4.2 CALCULATION OF $\eta_{CH_4,max}$

As mentioned in Section 2.3, $\eta_{CH_4,max}$ was used to approximate the maximum energy content in the biomass that was converted into bio-SNG ($\eta_{cg,tot}$). The product gas mainly contained; CH₄, C₂H₄, H₂, O₂, H₂O, CO and CO₂, which lead to:

$$\eta_{CH_4,max} = \frac{n_{CH_4,max} M_{CH_4} LHV_{CH_4}}{n_{CH_4} M_{CH_4} LHV_{CH_4} + n_{H_2} M_{H_2} LHV_{H_2} + n_{CO} M_{CO} LHV_{CO} + n_{C_2H_4} M_{C_2H_4} LHV_{C_2H_4}}$$

since O₂, H₂O and CO₂ have a LHV = 0.

4.3 GASIFICATION PROCESS ENERGY AND CARBON BALANCE

The flows entering and leaving the gasification process are illustrated in Figure 4.1. For a majority of the flows values of temperature, pressure, flow, and composition were measured, which were needed for the energy and carbon balance. For the unmeasured flows assumptions were made, which are presented in this section.

Gases from the methanation process include H_2S , CO_2 , and tar off gases, as well as NH_3 enriched stripper gas. These gases were recycled to the post combustion chamber for energy recovery and the composition and flow were not measured. The first case studied in this thesis is a case where only the gasification process was in operation and, therefore, there were no gases from the methanation process available. In the second case, only the tar absorbers in the methanation process were in operation, implying that the only gases from the methanation entering the post combustion chamber are the tar off gases. The tar off gases were assumed to contain the benzene that was analyzed in the product gas after the product gas fan.

As no measurements of the precoat material entering the gasification process were available, it was estimated to have a flow of 40 kg/h, based on the consumption over several days of operation. In the energy balance, the energy content of CaCO_3 was set equal to the heat of calcination of the amount of CaCO_3 , since it chemically decomposes in the combustion chamber

The inert gas in the two studied cases was N_2 , which wasn't included in the carbon balance. There were no measurements of the flow of N_2 , but the energy content was assumed to be negligible.

The electricity consumption of the gasification process was quantified by the installed effect of the largest electricity consumers, which were fans and pumps. In Table A.2 the installed effect of the total air fan, primary air fan, product gas fan, flue gas fan, hot water pumps, and the scrubber pump, is presented. The electricity consumption of the gasification process was approximated to be 0.443 MW and it was assumed that all of the electrical energy was converted into heat that entered the process.

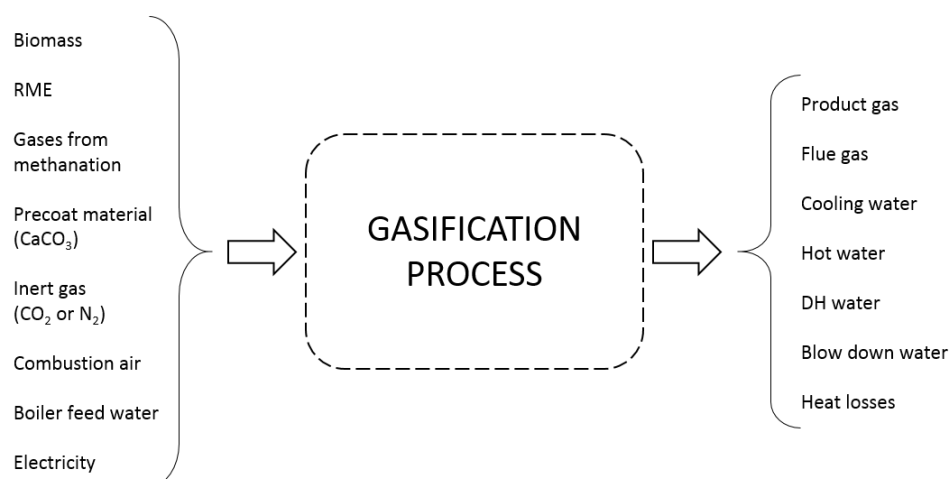


Figure 4.1: Flows entering and leaving the gasification process.

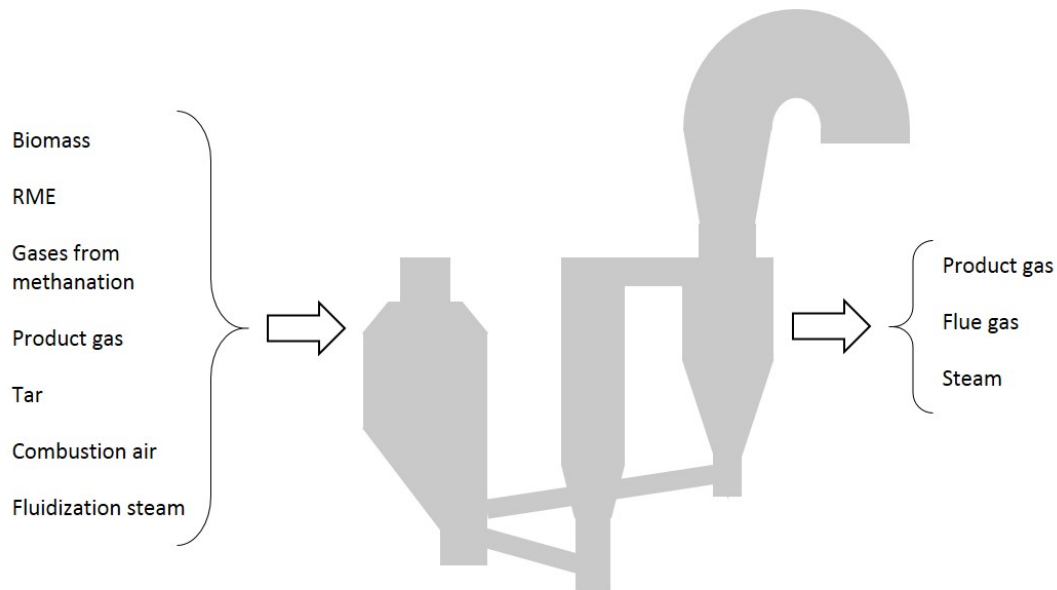


Figure 4.2: Flows entering and leaving the gasification section.

4.4 GASIFICATION SECTION ENERGY AND CARBON BALANCE

The flows entering and leaving the gasification section are illustrated in Figure 4.2. The main difference between the balance over the gasification section and the gasification process is that the product and flue gas cleaning systems are excluded, as well as the hot and cooling water systems.

The composition of the product gas was measured after the product gas fan and a tar analysis was done on the product gas leaving the gasifier. The dry composition of the product gas leaving the gasifier was approximated by adding these species together. The water vapor content in the product gas was calculated by adding the moisture in the biomass with the fluidization steam, which didn't participate in the gasification reactions in the gasifier. The steam was assumed to follow the direction of the bed material, implying that the steam leaving the gasifier, with the product gas, was the steam fluidizing the gasifier and the steam fluidizing the chute between the cyclone and the gasifier.

RME from the scrubber reservoir and particulates from the PG filter were combusted in the combustion chamber. Since, RME was not accumulated in the process, the RME flow was set to be equal to the measured RME flow entering the gasification process. The particulates and the tar in the RME were assumed to be equal to the hydrocarbons heavier than benzene measured in the product gas leaving the gasifier.

4.5 GASIFIER ENERGY AND CARBON BALANCE

The flows entering and leaving the gasifier are illustrated in Figure 4.3. The composition of the product gas was approximated with the same procedure as explained in Section 4.4.

The char flow, assumed to contain pure carbon, was approximated through the carbon balance, which made it possible to approximate the heat flow through the energy balance. To

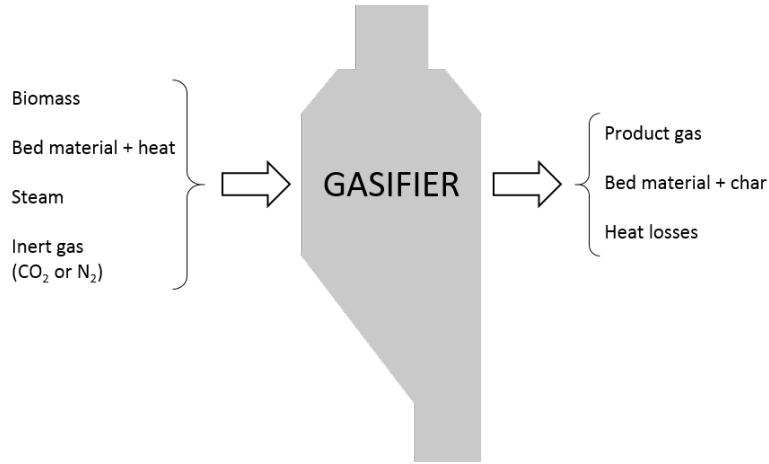


Figure 4.3: Flows entering and leaving the gasifier.

be able to calculate the fraction of char that is combusted (X_{comb}), the total char fraction of the biomass is needed, which is presented in Table 3.1.

$$X_{comb} = \frac{\dot{n}_{C,char} M_{char}}{Y_{char} \dot{m}_{biomass}}$$

4.6 SENSITIVITY ANALYSIS

One-at-a-time sensitivity analysis was used to evaluate the parameters that had the greatest influence on the energy and carbon entering and leaving the system, hereafter called output. This method made it possible to see which of the measured values that were of importance to measure with a small error since a small error in the measured values would affect the output significantly.

Each parameter in the energy and carbon balances was varied with 5% from their measured value one at a time, while the rest of the parameters were held constant. Each parameters influence on the output was evaluated by how many percent the output varied. γ is defined as the fraction of the percentile change of the output and of the variable.

$$\gamma = \frac{\Delta output}{\Delta variable} = \frac{\Delta output}{5\%}$$

Since a parameter can influence the energy and carbon balance differently, γ , for each parameter, was calculated for all of the different outputs ($\dot{Q}_{in}$, $\dot{Q}_{out}$, $\dot{n}_{C,in}$ and $\dot{n}_{C,out}$).

$$\gamma_{\dot{Q}_{in}} = \frac{\Delta \dot{Q}_{in}}{\Delta variable} \quad \gamma_{\dot{n}_{C,in}} = \frac{\Delta \dot{n}_{C,in}}{\Delta variable}$$

$$\gamma_{\dot{Q}_{out}} = \frac{\Delta \dot{Q}_{out}}{\Delta variable} \quad \gamma_{\dot{n}_{C,out}} = \frac{\Delta \dot{n}_{C,out}}{\Delta variable}$$

E.g. if a 5% error (change) in a measured value of an energy flow resulted in a 4% change in the total entering energy flow ($\gamma_{\dot{Q}_{in}} = 0.8$), an error in the measured value would result in a significant error in the balance. On the other hand, if the total entering energy flow only

changed with 0.1% ($\gamma \dot{Q}_{in} = 0.02$) a 5% error in the measured value will not affect the balance significantly.

To be able to decide if a possible error in a measurement of a parameters value influences the balance significantly, it was decided that if $\gamma \geq 0.2$ it was significant.

4.7 HEAT LOSSES

The heat losses in the system were mainly losses due to radiation, natural convection and heat losses associated with cooling that isn't converted into district heating or utilized within the process.

There are several different cooling steps within the process where the heat was utilized in different ways. E.g. cooling of flue gases, which preheated process air, steam, and hot water to the methanation process and district heating. The two process streams that were cooled where the heat wasn't utilized were; RME in the scrubber oil cooler and the product gas in product gas cooler II. Since these process streams were cooled with the cooling water, the heat losses associated with cooling was approximated as the energy content in the cooling water.

The heat losses in the system due to radiation and natural convection were estimated with the equations and correlation presented in Section 2.5. The surface temperature and emissivity coefficient of the gasification section was measure with an IR-camera at surfaces that were not covered with contact protection shields. It was assumed that the entire gasification section had the same surface temperature based on the surface temperature measurements and that the surrounding temperature was equal to 30 °C on all floors in the building. The surface area of the gasification section was estimated from the surface area of the different components.

While estimating the convective heat transfer coefficient, the gasifier, combustion chamber and the cyclone were approximated as vertical cylinders with the same height as the components and with a diameter that would correspond to their surface area. The convective heat transfer coefficient was also calculated when the components were approximated as several different cylinders with different diameters stacked on each other. As the post combustion chamber has a very noncylinder-like shape, only part of the post combustion chamber was used for estimation of the convective heat transfer coefficient.

The convective heat transfer coefficient was calculated for five segments on the gasifier, four on the combustion chamber, two on the post combustion chamber and three on the cyclone. The different segments are illustrated in Figure A.1 in Appendix.

An interval in which the convective heat transfer coefficient would lie within was chosen based on the calculations of the heat transfer coefficient for the different segments and the values that it takes on in air. An interval was also chosen for the emissivity coefficient. This lead to an interval of how large the heat losses in the gasification section could be and an average value over this interval was chosen.

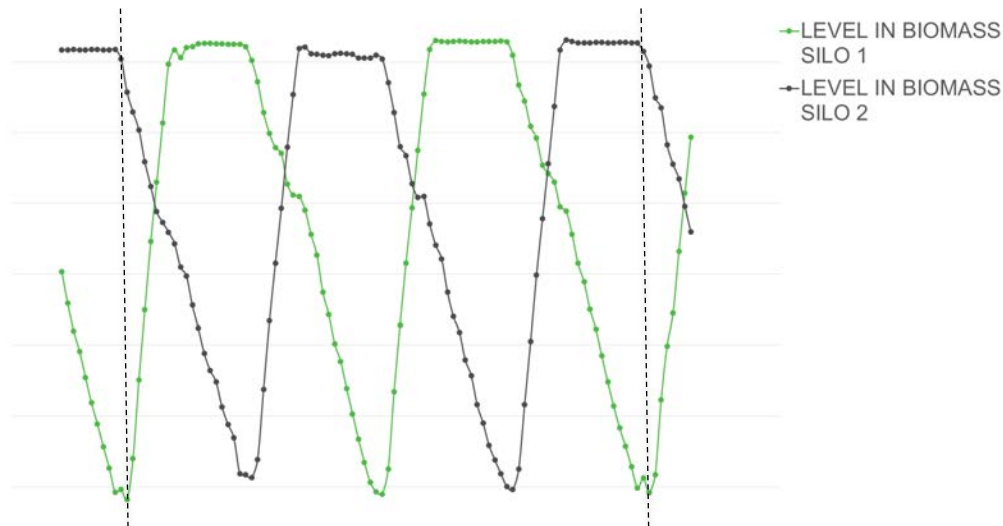


Figure 4.4: The change in level in biomass silo 1 and 2 with time.

4.8 BIOMASS FLOW

To make sure that the biomass flow entering the system boundary was equal to the flow entering the gasifier, the average biomass flow was taken during a time interval that included complete filling and emptying cycles of the biomass silos. In other words, the average biomass flow was calculated over a time period when the silos were filled the same amount of times as they were emptied. This was done by investigating the change in level in the two biomass silos with time, which can be seen in Figure 4.4. When the slope of the level function is positive the silo is filled with biomass and when the slope is negative the silo is emptied. The area in between the dashed lines, in Figure 4.4, include four complete filling and emptying cycles.

Because of the filling cycles of the silos, the logged data of the biomass flow was discontinuous and the flow was zero for periods. This meant that the calculated standard deviation of the biomass flow from the data was too large. Therefore, the standard deviation was set to 5 % of the mean biomass flow.

5 EXPERIMENTS

The two experimental cases studied in this thesis are presented in Table 5.1. The exported data for the cases, C1 and C2, are from operation during 21 h and 18 h, respectively. C1 is from the second day of operation, where the first day was starting up of the process, and C2 is from the third day.

As mentioned in Chapter 3, the gasification process can be operated under different conditions and loads. During operation without the methanation process, several unknown streams in the energy and carbon balance over the gasification process are left out. The unknown streams are off-gases from the methanation process that are combusted in the post combustion chamber and CO₂ as an inert gas. In the first case, only the gasification process was in operation, implying that N₂ was used as an inert gas and there were no off-gases. In the second case, the gasification process and the tar adsorbers in the methanation process were in operation, meaning that tar off-gases entered the post combustion chamber. N₂ was used as an inert gas in the second case, as well, since the CO₂ separation step in the methanation process was not in operation.

Table 5.1: Description of the different experiment cases.

Case	Date	Description	Load	Off-gases	Inert gas
C1	23/4 -15	Gasification process	90%	No	N ₂
C2	24/4 -15	Gasification process and tar off-gases	90%	Only tar	N ₂

6 RESULTS

In this chapter, the results from the energy and carbon balances over the gasification process, gasification section and the gasifier are presented. The performance of the gasification process is presented as the normal distributions and mean values of η_o and η_{cg} and the performance of the gasifier as $\eta_{cg,G}$ and X_{comb} . The maximum cold gas efficiencies of the entire process ($\eta_{cg,tot}$ and $\eta_{cg,G,tot}$) are presented as well.

The normal distributions of the energy flows entering and leaving the system ($\dot{Q}_{in}$ and $\dot{Q}_{out}$) and of the carbon flows entering and leaving the system ($\dot{n}_{C,in}$ and $\dot{n}_{C,out}$) are presented in graphs for the balances over the gasification process and section. The differences between the flows entering and leaving the system are presented in absolute values and in percentile values of the entering flows.

6.1 ENERGY AND CARBON BALANCE OVER THE GASIFICATION PROCESS

The normal distribution of the energy and carbon flow entering and leaving the gasification process calculated from process data are presented for case C1 and C2 in Figure 6.1 and 6.2, respectively. In both cases, the standard deviation (std) of the distribution for the energy entering the system ($\dot{Q}_{in}$) is larger than for the energy leaving the process. The reason for this is because of the large std of the biomass flow (5 % of the average value).

In Table 6.1 the calculated average and normalized energy flows entering and leaving the gasification process for the two cases are presented. The largest energy source was the biomass and the largest energy products were the product gas, hot water and cooling water.

The difference between the energy entering and leaving the system ($\dot{Q}_{in} - \dot{Q}_{out}$) was equal to 0.22 MW in C1, which was 0.8 % of the total energy flow entering the gasification process. In C2, $\dot{Q}_{in} - \dot{Q}_{out}$ was equal to -0.05 MW, which was 0.2 % of the total energy flow entering the gasification process. The heat losses in the gasification process were equal to the energy flow of the flue gas ($\dot{Q}_{FG}$), cooling water ($\dot{Q}_{CW}$), blow down water ($\dot{Q}_{BDW}$) and the convective and radiative heat losses ($\dot{Q}_{HL}$). The heat losses summed up to 5.24 MW in C1, which was

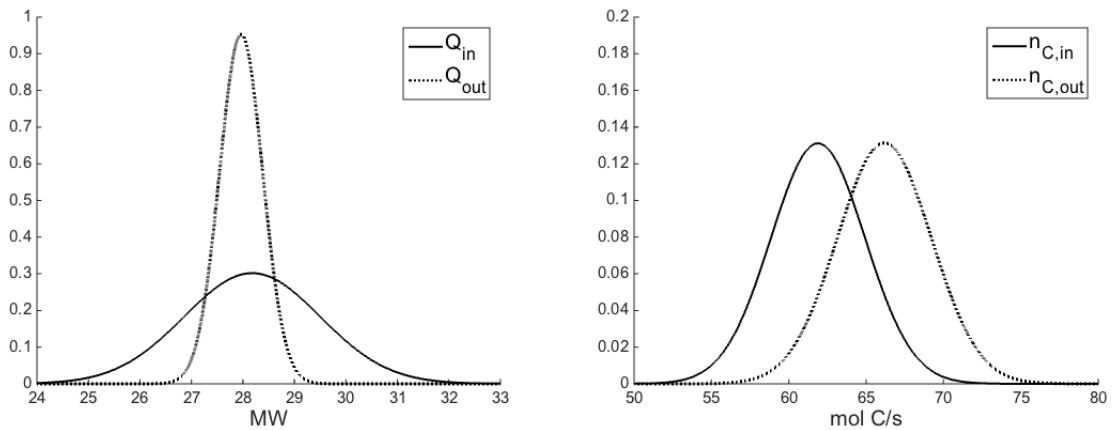


Figure 6.1: Normal distribution of the calculated energy and carbon flows entering and leaving the gasification process for case C1.

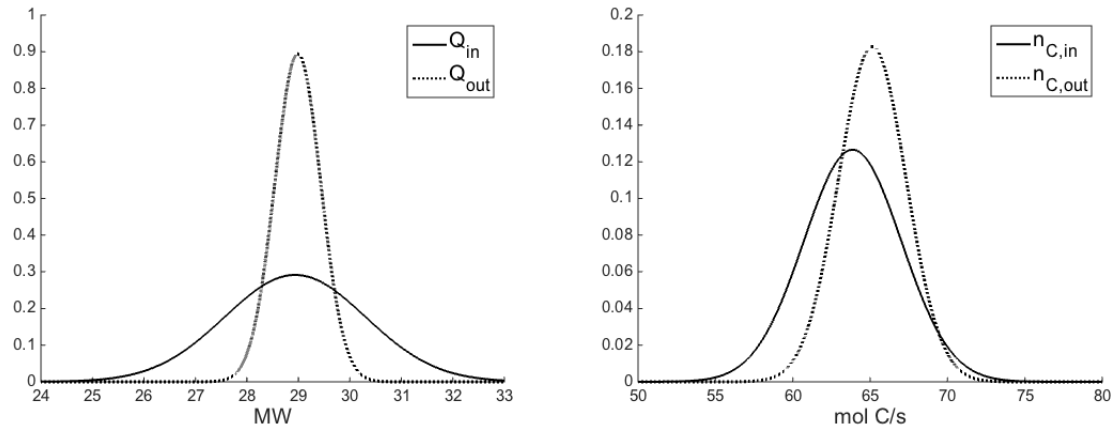


Figure 6.2: Normal distribution of the calculated energy and carbon flows entering and leaving the gasification process for case C2.

18.6 % of the total energy flow entering the gasification process. In C2, the heat losses summed up to 4.24 MW, which was 14.6 % of the total energy flow entering the gasification process.

In Table 6.2 the calculated average and normalized carbon flows entering and leaving the gasification process for the two cases are presented. The largest carbon source was the biomass and the two largest carbon products were the product and the flue gas.

Table 6.1: Average energy flows, in MW, and normalized energy flows entering and leaving the gasification process.

$\dot{Q}_i$ [MW]	C1	C2	$\eta_i = \dot{Q}_i / \dot{Q}_{in}$	C1	C2
$\dot{Q}_{bio}$	26.49	27.36	η_{bio}	0.940	0.945
$\dot{Q}_{RME}$	1.06	1.06	η_{RME}	0.038	0.037
$\dot{Q}_{off}$	-	0.01	η_{off}	-	0.000
$\dot{Q}_{CaCO_3}$	-0.03	-0.03	η_{CaCO_3}	-0.001	-0.001
$\dot{Q}_{BFW}$	0.22	0.10	η_{BFW}	0.008	0.003
$\dot{Q}_{el}$	0.44	0.44	η_{el}	0.016	0.015
$\dot{Q}_{in}$	28.18	28.94	η_{in}	1.000	1.000
$\dot{Q}_{PG}$	20.25	21.05	η_{PG}	0.719	0.727
$\dot{Q}_{FG}$	0.73	0.59	η_{FG}	0.026	0.020
$\dot{Q}_{HW}$	2.45	3.68	η_{HW}	0.087	0.127
$\dot{Q}_{CW}$	3.64	2.78	η_{CW}	0.129	0.096
$\dot{Q}_{DHW}$	0.02	0.02	η_{DHW}	0.001	0.001
$\dot{Q}_{BDW}$	0.01	0.01	η_{BDW}	0.000	0.000
$\dot{Q}_{HL}$	0.86	0.86	η_{HL}	0.031	0.030
$\dot{Q}_{out}$	27.96	28.99	η_{out}	0.992	1.002
$\dot{Q}_{in} - \dot{Q}_{out}$	0.22	-0.05	$\eta_{in} - \eta_{out}$	0.008	-0.002

Table 6.2: Average carbon flows, in mol/s, and normalized carbon flows entering and leaving the gasification process.

$\dot{n}_{C,i}$ [mol C/s]	C1	C2	$v_{C,i} = \dot{n}_{C,i}/\dot{n}_{C,in}$	C1	C2
$\dot{n}_{C,biomass}$	59.81	61.77	$v_{C,biomass}$	0.966	0.967
$\dot{n}_{C,RME}$	1.97	1.97	$v_{C,RME}$	0.032	0.031
$\dot{n}_{C,off}$	-	0.02	$v_{C,off}$	-	0.000
$\dot{n}_{C,CaCO_3}$	0.11	0.11	$v_{C,CaCO_3}$	0.002	0.002
$\dot{n}_{C,in}$	61.89	63.88	$v_{C,in}$	1.000	1.000
$\dot{n}_{C,PG}$	43.09	42.91	$v_{C,PG}$	0.696	0.672
$\dot{n}_{C,FG}$	23.09	22.22	$v_{C,FG}$	0.373	0.348
$\dot{n}_{C,out}$	66.18	65.14	$v_{C,out}$	1.069	1.020
$\dot{n}_{C,in} - \dot{n}_{C,out}$	-4.29	-1.26	$v_{C,in} - v_{C,out}$	-0.069	-0.020

The difference between the carbon flow entering and leaving the system ($\dot{n}_{C,in} - \dot{n}_{C,out}$) was equal to -4.29 mol/s in C1, which was 6.9 % of the total carbon flow entering the gasification process. In C2, $\dot{n}_{C,in} - \dot{n}_{C,out}$ was equal to -1.26 mol/s, which was 2.0 % of the total carbon flow entering the system.

As mentioned in Section 4.6, only the variables with $\gamma \geq 0.2$ from the one-at-a-time sensitivity analysis are considered significant. These variables and the value of γ are presented in Table 6.3 and Table A.6 for C1 and C2, respectively. Some variables influence the energy and carbon balance in the same order of magnitude, such as the biomass and product gas flow. Other parameters influence only the energy or carbon balance, such as the temperatures in the hot water system and the CO₂ concentration in the flue gas, respectively. In Table 6.3 and A.6, $T_{HW,1} - T_{HW,4}$ are four temperatures in the hot water system and $T_{RME,in}$ is the temperature of RME in the scrubber reservoir.

In Figure 6.3 and 6.4, the normal distribution of the total and gasifier cold gas efficiency and overall thermal efficiency are presented for case C1 and C2, respectively. The distributions of

Table 6.3: Variables with $\gamma \geq 0.2$ from the OAT sensitivity analysis of the energy and carbon balance over the gasification process for C1.

	$\gamma_{\dot{Q}_{in}}$	$\gamma_{\dot{Q}_{out}}$	$\gamma_{\dot{n}_{C,in}}$	$\gamma_{\dot{n}_{C,out}}$
$\dot{m}_{biomass}$	0.94		0.97	
$\dot{V}_{PG}$		0.72		0.65
$\dot{V}_{FG}$		0.03		0.35
$Y_{CO_2,FG}$				0.31
$T_{HW,1}$		0.79		
$T_{HW,2}$		-0.75		
$T_{HW,3}$		0.98		
$T_{HW,4}$		-0.87		
$T_{RME,in}$		0.23		

C1 and C2 show that the total and gasifier cold gas efficiencies were within 73 – 80 % and 80 – 88 %, respectively. The average values of the total and gasifier cold gas efficiencies are similar for case C1 and C2, which can be seen in Table 6.5. The overall efficiency on the other hand differs between the two cases, where the mean value was larger in C2 than C1. The distributions of C1 and C2 show that the overall efficiency was within 80 – 86 %.

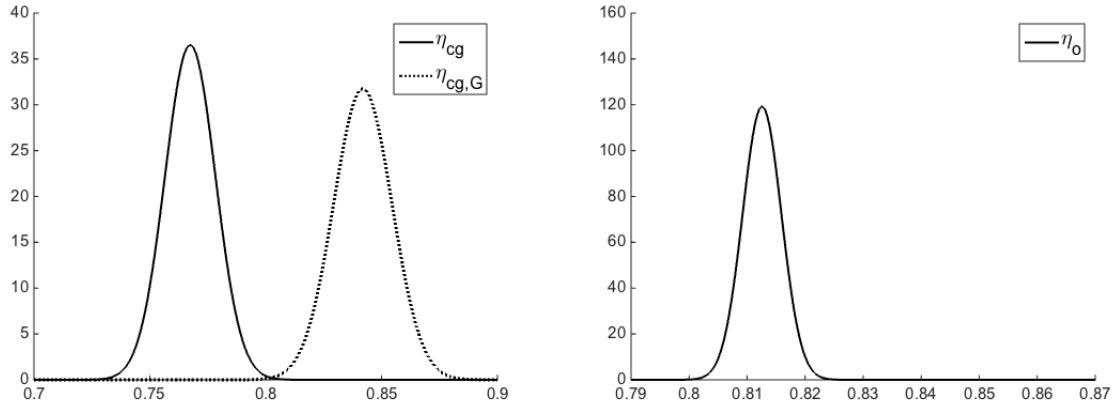


Figure 6.3: Total and gasifier cold gas efficiencies and overall thermal efficiency for case C1.

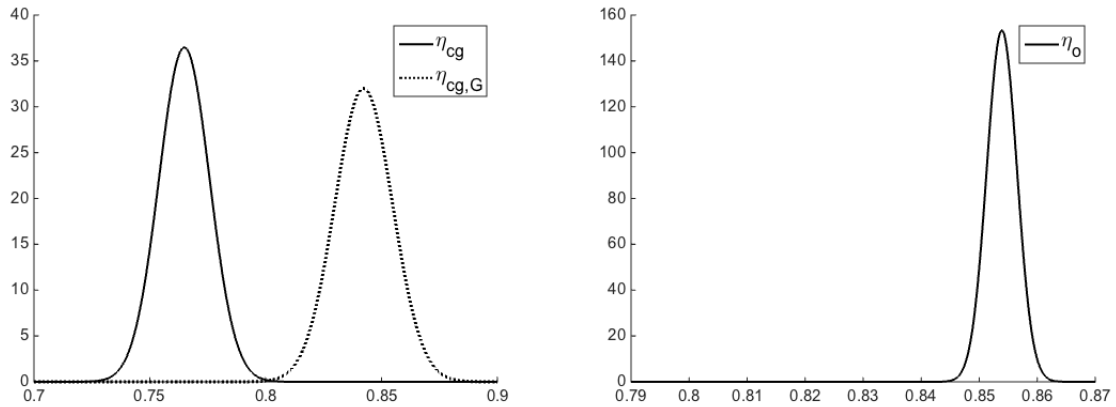
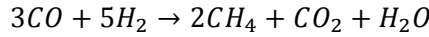


Figure 6.4 Total and gasifier cold gas efficiencies and overall thermal efficiency for case C2.

Table 6.5: Average values of the efficiencies and the maximum efficiencies for C1 and C2.

	C1	C2
η_{cg}	76.7 %	76.5 %
$\eta_{cg,tot}$	62.9 %	62.7 %
$\eta_{cg,G}$	84.2 %	84.2 %
$\eta_{cg,G,tot}$	69.0 %	69.0 %
η_o	81.3 %	85.4 %

The molar fraction of the species in the wet product gas were approximately; 8.5 % CH₄, 1.9 % C₂H₄, 40 % H₂, and 24 % CO. By combining reactions R5 and R6 in Section 2.1:



it is clear that a maximum of 16 mol of CH₄ can be produced from the CO and H₂ in the product gas. Through reaction R7 in Section 2.1, it is clear that 1 mol CH₄ can be created from 1 mol C₂H₄.

This allows $\eta_{CH_4,max}$ to be calculated accordingly:

$$\eta_{CH_4,max} = \frac{(8.5 + 16 + 1.9)M_{CH_4}LHV_{CH_4}}{8.5M_{CH_4}LHV_{CH_4} + 40M_{H_2}LHV_{H_2} + 24M_{CO}LHV_{CO} + 1.9M_{C_2H_4}LHV_{C_2H_4}} \approx 82 \%$$

which is used to calculate $\eta_{cg,tot}$ and $\eta_{cg,G,tot}$ presented in Table 6.5.

6.2 ENERGY AND CARBON BALANCE OVER THE GASIFICATION SECTION

The sensitivity analysis over the gasification process showed that the temperatures in the hot water system had a significant influence on the energy balance. An energy and carbon balance was, therefore, set up over the gasification section to exclude the hot water system.

The normal distribution of the energy and carbon flow entering and leaving the gasification section calculated from process data are presented in Figure 6.5 and 6.6 respectively. As in the balance over the gasification process, the standard deviation of $\dot{Q}_{in}$ is larger than for $\dot{Q}_{out}$. The values of the average calculated and normalized energy and carbon flows entering and leaving the gasification section are presented in Table 6.4 and 6.5 respectively.

The difference between the energy entering and leaving the system ($\dot{Q}_{in} - \dot{Q}_{out}$) was equal to 0.28 MW in C1, which was 0.8 % of the total energy flow entering the gasification section. In C2, $\dot{Q}_{in} - \dot{Q}_{out}$ was equal to 0.63 MW, which was 1.8 % of the total energy flow entering the gasification section. The difference between the carbon flow entering and leaving the

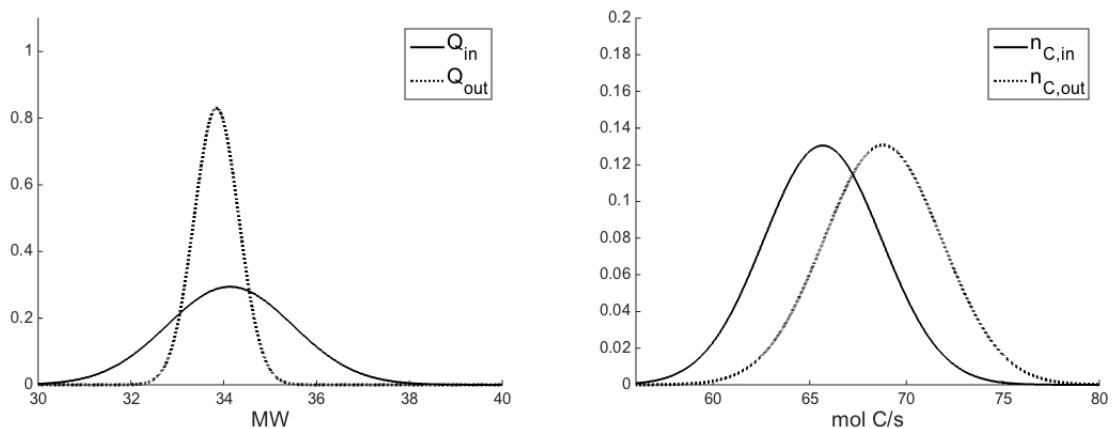


Figure 6.5: Normal distribution of the calculated energy and carbon flow entering and leaving the gasification section for C1.

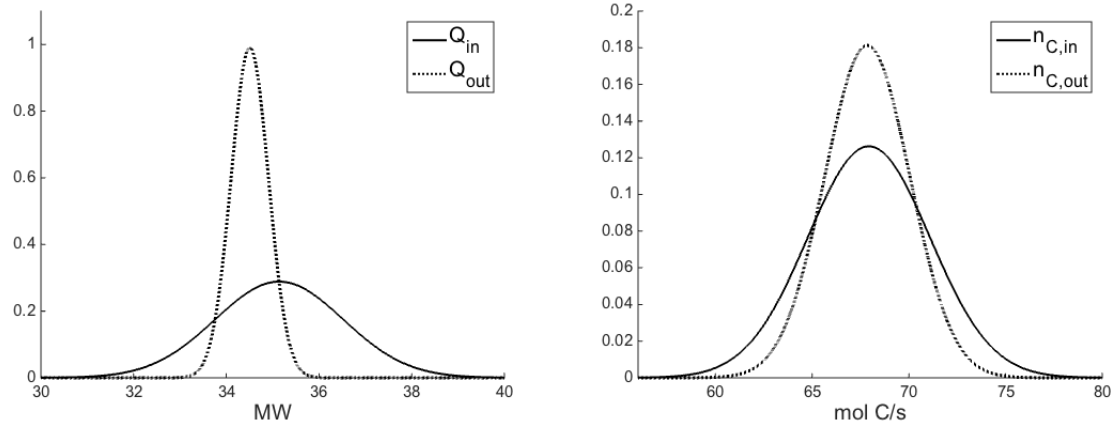


Figure 6.6: Normal distribution of the calculated energy and carbon flow entering and leaving the gasification section for C2.

system ($\dot{n}_{C,in} - \dot{n}_{C,out}$) was equal to -3.11 mol/s in C1, which was 4.7 % of the total carbon flow entering the gasification section. In C2, $\dot{n}_{C,in} - \dot{n}_{C,out}$ was equal to -0.05 mol/s, which was 0.1 % of the total carbon flow entering the gasification section.

Table 6.4: Average energy flows, in MW, and normalized energy flows entering and leaving the gasification section.

$\dot{Q}_i$ [MW]	C1	C2	$\eta_i = \dot{Q}_i / \dot{Q}_{in}$	C1	C2
$\dot{Q}_{bio}$	26.49	27.36	η_{bio}	0.776	0.779
$\dot{Q}_{RME}$	1.06	1.06	η_{RME}	0.031	0.030
$\dot{Q}_{off}$	-	0.01	η_{off}	-	0.000
$\dot{Q}_{PG,rec}$	1.86	2.06	$\eta_{PG,rec}$	0.054	0.059
$\dot{Q}_{tar}$	0.50	0.52	η_{tar}	0.015	0.015
$\dot{Q}_{steam,in}$	3.26	3.26	$\eta_{steam,in}$	0.096	0.093
$\dot{Q}_{air}$	0.96	0.87	η_{air}	0.028	0.025
$\dot{Q}_{in}$	34.13	35.14	η_{in}	1.000	1.000
$\dot{Q}_{PG,G}$	23.16	24.14	$\eta_{PG,G}$	0.679	0.687
$\dot{Q}_{FG}$	4.91	4.58	η_{FG}	0.144	0.130
$\dot{Q}_{steam,out}$	4.91	4.93	$\eta_{steam,out}$	0.144	0.140
$\dot{Q}_{HL}$	0.86	0.86	η_{HL}	0.025	0.024
$\dot{Q}_{out}$	33.85	34.51	η_{out}	0.992	0.982
$\dot{Q}_{in} - \dot{Q}_{out}$	0.28	0.63	$\eta_{in} - \eta_{out}$	0.008	0.018

Table 6.5: Average carbon flows, in mol/s, and normalized carbon flows entering and leaving the gasification section.

$\dot{n}_{C,i}$ [mol C/s]	C1	C2	$v_{C,i} = \dot{n}_{C,i}/\dot{n}_{C,in}$	C1	C2
$\dot{n}_{C,biomass}$	59.81	61.77	$v_{C,biomass}$	0.911	0.910
$\dot{n}_{C,RME}$	1.97	1.97	$v_{C,RME}$	0.030	0.030
$\dot{n}_{C,off}$	-	0.02	$v_{C,off}$	-	0.000
$\dot{n}_{C,PG,rec}$	3.89	4.15	$v_{C,PG,rec}$	0.059	0.061
$\dot{n}_{C,tar}$	0.01	0.01	$v_{C,tar}$	0.000	0.000
$\dot{n}_{C,in}$	65.68	67.91	$v_{C,in}$	1.000	1.000
$\dot{n}_{C,PG,G}$	45.70	45.63	$v_{C,PG,G}$	0.696	0.672
$\dot{n}_{C,FG}$	23.08	22.23	$v_{C,FG}$	0.351	0.327
$\dot{n}_{C,out}$	68.79	67.86	$v_{C,out}$	1.047	0.999
$\dot{n}_{C,in} - \dot{n}_{C,out}$	-3.11	0.05	$v_{C,in} - v_{C,out}$	-0.047	0.001

6.3 ENERGY AND CARBON BALANCE OVER THE GASIFIER

The energy and carbon balance over the gasifier differed from the balances over the gasification process and section. The balance over the gasifier was not an over determined system, since the char and heat flow were not measured. They are solved from the energy and carbon balance, implying that $\eta_{out} = 1$ and $v_{C,out} = 1$. From the calculated energy and carbon flows, presented in Table 6.6 and 6.7, the fraction of char that is combusted X_{comb} was calculated. The normal distributions of X_{comb} , illustrated in Figure 6.7, show that X_{comb} lied within 30 – 100%. The average value of X_{comb} was 64.1 % and 71.1% for C1 and C2 respectively.

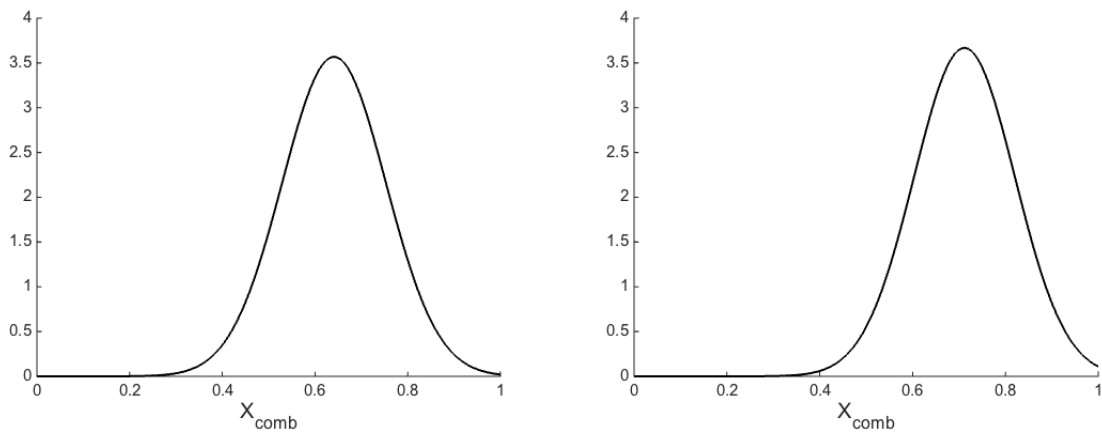


Figure 6.7: Normal distribution of the calculated fraction of char that is combusted, X_{comb} , for C1 (to the left) and C2 (to the right).

Table 6.6: Average energy flows, in MW, and normalized energy flows entering and leaving the gasifier.

$\dot{Q}_i$ [MW]	C1	C2	$\eta_i = \dot{Q}_i/\dot{Q}_{in}$	C1	C2
$\dot{Q}_{biomass}$	26.49	27.36	$\eta_{biomass}$	0.780	0.765
$\dot{Q}_{steam,in}$	3.20	3.20	$\eta_{steam,in}$	0.094	0.089
$\dot{Q}_{heat}$	4.27	5.21	η_{heat}	0.126	0.146
$\dot{Q}_{in}$	33.96	35.77	η_{in}	1.000	1.000
$\dot{Q}_{PG,G}$	23.16	24.14	$\eta_{PG,G}$	0.682	0.675
$\dot{Q}_{steam,out}$	4.91	4.93	$\eta_{steam,out}$	0.145	0.138
$\dot{Q}_{char}$	5.65	6.47	η_{char}	0.166	0.181
$\dot{Q}_{HL,G}$	0.23	0.23	$\eta_{HL,G}$	0.007	0.006
$\dot{Q}_{out}$	33.96	35.77	η_{out}	1.000	1.000

Table 6.7: Average carbon flows, in mol/s, and normalized carbon flows entering and leaving the gasifier.

$\dot{n}_{C,i}$ [mol C/s]	C1	C2	$\nu_{C,i} = \dot{n}_{C,i}/\dot{n}_{C,in}$	C1	C2
$\dot{n}_{C,biomass}$	59.81	61.77	$\nu_{C,biomass}$	1.000	1.000
$\dot{n}_{C,in}$	59.81	61.77	$\nu_{C,in}$	1.000	1.000
$\dot{n}_{C,PG,G}$	45.70	45.63	$\nu_{C,PG,G}$	0.764	0.739
$\dot{n}_{C,char}$	14.11	16.14	$\nu_{C,char}$	0.236	0.261
$\dot{n}_{C,out}$	59.81	61.77	$\nu_{C,out}$	1.000	1.000

7 DISCUSSION

Calculating with randomized input data with a normal distribution is a useful method to include uncertainties in measurements. From the sensitivity analysis it is clear that the four streams that have the largest influence on the balance over the gasification process are; the biomass, product gas, flue gas, and the hot water system. The biomass and product gas are the largest energy carriers, and the biomass, product gas and flue gas are the largest carbon carriers. Since, the temperatures in the hot water system had a large impact on the energy balance, a balance over the gasification section was set up to eliminate the hot water system from the balance. The distributions from the energy and carbon balances over the gasification process and the gasification section do not show any large errors in the measured data or the assumptions.

In both the balance over the gasification process and section, $|v_{C,in} - v_{C,out}|$ was larger for C1 than for C2 and $v_{C,out} > 1$ for C1. An explanation for this could be that the process had not yet reached carbon steady state after starting up the process, implying that there was carbon left in the process during C1, which resulted in more carbon leaving the system than entering. From these results, it is recommended that data for calculations should be exported when the process has been in operation for at least two to three days.

As presented in Section 1.2, the performance goals set up by Göteborg Energi are $\eta_{cg,tot} \geq 65\%$ and $\eta_{o,tot} \geq 90\%$. Assuming a maximum conversion of the product gas to methane, showed that the goal of $\eta_{cg,tot} \geq 65\%$ can be met if less product gas is combusted. If less product gas is combusted, the cold gas efficiency of the product gas from the gasification process (η_{cg}) will increase towards the cold gas efficiency of the product gas from the gasifier ($\eta_{cg,G}$). To be able to meet the goal of $\eta_{o,tot} \geq 90\%$, the heat losses in the process need to be decreased.

The convective and radiative heat losses were about 3% of the energy entering the process. The total calculated heat losses were 18.6 % of the energy entering the process in C1 and 14.6 % in C2. There are uncertainties in the magnitude of the heat losses, as well, since these numbers don't include $\dot{Q}_{in} - \dot{Q}_{out}$, which could be in the form of heat losses. The surface temperatures of the gasification section are high ($\sim 140^\circ\text{C}$) and the convective and radiative heat losses are dependent on the surface temperature. The radiative and convective heat losses could, therefore, be decreased by decreasing the surface temperature, through increased isolation of the walls of the gasification section. The heat losses, not associated with radiation and convection, can be decreased by; decreasing the temperatures in the gasification section, increasing the preheating of steam and air, and combusting less product gas in the combustion chamber.

Product gas is combusted in the combustion chamber as a supplementary fuel to supply the heat demand in the gasifier. If the temperatures in the gasification section were lower, less product gas would have to be combusted. Another result of lowering the temperature would be that less energy, from the gasification section, would have to be used to heat up the fluidization steam and combustion air.

The average value of η_o ($= \eta_{PG} + \eta_{DHW} + \eta_{HW}$) was higher for C2 than for C1. The difference in η_o was caused by the value of η_{HW} being higher for C2 than for C1. η_{CW} was lower for C2 than for C1, which means that the product gas was not cooled as much by the hot water system for C1 as for C2 and, therefore, more cooling was done by the cooling water.

The energy and carbon balance over the gasifier differs from the balances over the gasification process and section in the sense that the gasifier is not an over determined system. The biomass and product gas flow were used to calculate the char and heat flow through the carbon and energy balance, respectively. This implies that there are uncertainties in the results from this balance, which should be kept in mind. The method for calculating the flows entering and leaving the gasifier can be used as a first approximation of the heat flow, which can be used to approximate the flow of bed material. The standard deviations of X_{comb} are large for both C1 and C2, because of the uncertainties in the biomass flow and char content of the biomass. In C1, a smaller fraction of the char in the biomass is combusted in the combustion chamber than in C2. This could also be explained by the carbon steady state hypothesis stating that there was carbon left in the system in case C1.

A more thorough investigation can be done to verify the measured data associated with the streams that affect the balances the most. The measurements affecting the hot and cooling water systems could be verified by setting up a heat balance over the flue gas cleaning system, and the product and flue gas flows can be verified by using a helium tracing method, described by Larsson.

There are uncertainties in the measurement of the biomass flow today through the conveyor belt. One alternative method of calculating the biomass flow, and verifying the results from the conveyor belt, is to measure the amount of biomass that is delivered on site during a longer time of operation. This requires that the process is operated at the same load during the entire operation and that the amount of product gas that is recirculated is constant. This, since, the load of the process is determined by the product gas entering the methanation process, and not the total amount of product gas produced by the gasifier. A larger product gas flow out of the gasifier requires a larger biomass flow, implying that two different operation cases can have the same load, but the biomass flow can differ.

8 CONCLUSIONS

- The method of comparing the normal distributions from the energy and carbon balances does not show any large errors in the measurements and/or the assumptions of the process data.
- The streams that had the largest influences on the energy and carbon balance over the gasification process were; the biomass, product gas, flue gas, and the hot water system.
- The cold gas efficiency of the gasification process and the gasifier cold gas efficiency were within 73 – 80 % and 80 – 88 %, respectively. Reaching a cold gas efficiency of 65 % over the entire process is possible if less product gas is combusted in the combustion chamber.
- The overall thermal efficiency of the gasification process was within 80 – 86 %. To reach an overall thermal efficiency of 90 % over the entire process, the heat losses need to be decreased.
- The losses in the gasification process could be decreased by; increasing the isolation of the gasification section, decreasing the temperatures in the gasification section, increasing the preheating of steam and air, and combusting less product gas.
- The fraction of the char in the biomass that was combusted in the combustion chamber was within 30 – 100 % and the average value was within 64 – 71 %.

9 REFERENCES

- Andersson L, 2015. *GoBiGas - Start-up experiences*. Göteborg: Göteborg Energi.
- Basu P, 2010. Gasification Theory and Modeling of Gasifiers. In: *Biomass Gasification and Pyrolysis - Practical Design and Theory*. Burlington: Elsevier Inc.
- Biofuels Research Advisory Council, 2006. *Biofuels in the European Union - A vision for 2030 and beyond*.
- Elliott J R, & Lira C T, 1999. *Introductory Chemical Engineering Thermodynamics*. New Jersey: Prentice Hall PTR.
- Energimyndigheten, 2013. *Energiläget 2013*.
- European Commission, 2015. *European Commission*. [Online]
Available at: http://ec.europa.eu/clima/policies/transport/index_en.htm
[Accessed 26 01 2015].
- Gómez-Barea A, & Leckner B, 2010. Modeling of biomass gasification in fluidized bed. *Progress in Energy and Combustion Science* (36), 444-509.
- Gröbl T, Walter H, & Haider M, 2012. Biomass steam gasification for production of SNG - Process design. *Applied Energy* (97), 451-461.
- Göteborg Energi, 2015. *GoBiGas*. [Online]
Available at: <http://gobigas.goteborgenergi.se/>
[Accessed 26 01 2015].
- Heyne S, 2013. *Bio-SNG from Thermal Gasification - Process Synthesis, Integration and Performance*. Göteborg: Diss. Chalmers University of Technology.
- Larsson A, 2014. *Fuel Conversion in a Dual Fluidized Bed Gasifier - Experimental Quantification and Impact on Performance*. Energy and Environment. Gothenburg: Diss. Chalmers University of Technology.
- Larsson A, Seeman M, Neves D, & Thunman H, 2013. Evaluation of Performance of Industrial-Scale Dual Fluidized Bed Gasifier Using the Chalmers 2-4-MWth Gasifier. *Energy & Fuels*, 27, 6665-6650.
- Lundberg L, Tchoffor P, Pallarès D, Johansson R, Thunman H, & Davidsson K, 2015. Influence of Conversion Conditions on the Gasification Rate of Biomass Char in a Fluidised Bed. *Fuel Processing Technology*.
- Nguyen T B.D, Ngo S I, Lim Y, Lee J W, Lee U, & Song B, 2012. Three-stage steady-state model of biomass gasification in a dual circulating fluidized-bed. *Energy Conversion and Management* (54), 100-112.
- Nicholson J, 2014. *The Concise Oxford Dictionary of Mathematics*. 5th ed. Oxford University Press.
- Rice J, 2007. *Mathematical Statistics and Data Analysis*. 3rd ed. Belmont: Brooks/Cole.
- Saltelli A, Chan K, & Scott E, 2000. *Sensitivity Analysis*. Chichester: John Wiley & Sons.
- Turns S R, 1996. *An Introduction to Combustion*. New York: McGraw-Hill.

Welty J R, Wicks C E, Wilson R E, & Rorrer G L, 2008. *Fundamentals of Momentum, Heat, and Mass Transfer*. 5th ed. Oregon: John Wiley & Sons.

Zhang W, 2010. Automotive fuels from biomass via gasification. *Fuel Processing Technology* (91), 866-876.

APPENDIX

A CALCULATION OF ENERGY FLOWS

Table A.1: Calculation of energy flows in the energy balance.

Calculation of energy flows	
$\dot{Q}_{BDW}$	$\dot{m}_{BDW} \int_{T_{ref}}^{T_{BDW}} C_{p_{H_2O(l)}} dT$
$\dot{Q}_{BFW}$	$\dot{m}_{BFW} \int_{T_{ref}}^{T_{BFW}} C_{p_{H_2O(l)}} dT$
$\dot{Q}_{bio}$	$\dot{m}_{bio}(LHV_{bio} + \int_{T_{ref}}^{T_{bio}} C_{p_{bio}} dT)$
$\dot{Q}_{CaCO_3}$	$\dot{m}_{CaCO_3} h_{calcination}$
$\dot{Q}_{char}$	$\dot{m}_{char,bio} X_{comb} (LHV_{char} + \int_{T_{ref}}^{T_{char}} C_{p_{char}} dT)$
$\dot{Q}_{CW}$	$\dot{m}_{CW} \int_{T_{ref}}^{T_{CW}} C_{p_{H_2O(l)}} dT$
$\dot{Q}_{DHW}$	$\dot{m}_{DHW} \int_{T_{ref}}^{T_{DHW}} C_{p_{H_2O(l)}} dT$
$\dot{Q}_{el}$	0.443 MW
$\dot{Q}_{FG}$	$\dot{m}_{FG} \int_{T_{ref}}^{T_{FG}} C_{p_{FG}} dT$
$\dot{Q}_{heat}$	$\dot{Q}_{PG,G} + \dot{Q}_{char} + \dot{Q}_{HL,G} - \dot{Q}_{bio} - \dot{Q}_{steam}$
$\dot{Q}_{HL}$	$A(h\Delta T + \epsilon\sigma(T_H^4 - T_c^4))$
$\dot{Q}_{HL,G}$	$A_G(h\Delta T + \epsilon\sigma(T_H^4 - T_c^4))$
$\dot{Q}_{HW}$	$\dot{m}_{HW} \int_{T_{ref}}^{T_{HW}} C_{p_{H_2O(l)}} dT$
$\dot{Q}_{off}$	$\dot{m}_{off}(LHV_{off} + \int_{T_{ref}}^{T_{off}} C_{p_{off}} dT)$
$\dot{Q}_{PG}$	$\dot{m}_{PG}(LHV_{PG} + \int_{T_{ref}}^{T_{PG}} C_{p_{PG}} dT)$
$\dot{Q}_{PG,G}$	$\dot{m}_{PG,G}(LHV_{PG,G} + \int_{T_{ref}}^{T_{PG,G}} C_{p_{PG,G}} dT)$
$\dot{Q}_{RME}$	$\dot{m}_{RME}(LHV_{RME} + \int_{T_{ref}}^{T_{RME}} C_{p_{RME}} dT)$
$\dot{Q}_{steam}$	$\dot{m}_{steam} \int_{T_{ref}}^{T_{steam}} C_{p_{H_2O(g)}} dT$

B INSTALLED EFFECT OF LARGEST ELECTRICITY CONSUMERS

The largest electricity consumers in the gasification process were fans and pumps and their installed effect is presented in Table A.2.

Table A.2: *Installed effect of the largest electricity consumers in the gasification process.*

	Installed effect [kW]
Total air fan	140
Primary air fan	10
Product gas fan	130
Flue gas fan	100
Hot water pumps	51
Scrubber pump	12
Total	443

C CONVECTIVE HEAT TRANSFER COEFFICIENT CALCULATION

Table A.3: Results from calculation of the convective heat transfer coefficient of the gasification section with a surface temperature of 140 °C.

Segment		Area [m ²]	Length [m]	Diameter [m]	Nu_L [-]	h [W/m ² K]
Gasifier	tot		10.4		1546	5.1
	1	1.7	0.4	1.4	69	6.0
	2	18.8	1.1	5.7	171	5.6
	3	61.5	4.2	4.7	635	5.2
	4	52.4	4.2	4.0	636	5.2
	5	3.9	0.6	2.1	102	5.8
Comb ch	tot		14.0		2071	5.1
	1	54.2	8.5	2.0	1265	5.1
	2	13.7	2.5	1.7	389	5.3
	3	8.7	2.0	1.4	309	5.4
	4	3.2	1.1	1.0	174	5.6
Post comb	1	47.2	4.0	3.8	605	5.2
	2	41.3	5.1	2.6	767	5.2
Cyclone	tot		8.2		1224	5.1
	1	2.2	0.5	1.4	86	5.9
	2	36.5	3.8	3.0	585	5.3
	3	23.1	3.8	1.9	589	5.3

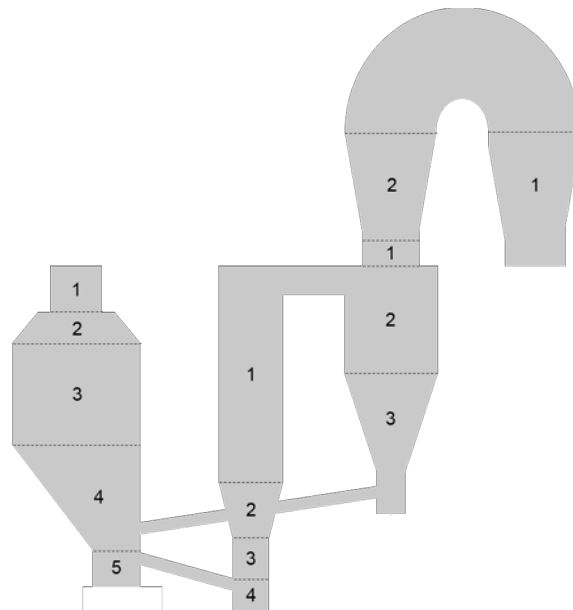


Figure A.1: Illustration of the segments in the gasification section for which the convective heat transfer coefficient was calculated.

D RADIATIVE AND CONVECTIVE HEAT LOSSES

Through calculations according to the equations and correlation in Section 2.5, the convective and radiative heat losses were estimated over the gasification section. The measured surface temperatures and their position can be seen in Figure A.2, where the temperatures next to the dashed lines represent surface temperatures measured on flanges. The overall surface temperature of the gasification section was approximated to 140 °C which was used for the calculations of the convective and radiative heat losses.

From the calculations of the convective heat transfer coefficient, presented in Appendix C, the chosen interval was 5 – 6 W/m^2K . In Table A.4, the approximated heat losses from natural convection with a surface temperature of 140 °C and a surrounding temperature of 30 °C are presented.

Table A.4: Approximated heat losses from natural convection over the gasification section with a surface temperature of 140 °C and surrounding temperature of 30 °C.

	Area [m ²]	h_{min} [W/m ² K]	h_{max} [W/m ² K]	$\dot{Q}_{min}$ [MW]	$\dot{Q}_{max}$ [MW]
Gasifier	140	5	6	0.08	0.09
Comb ch	90	5	6	0.05	0.06
Post comb	210	5	6	0.12	0.14
Cyclone	70	5	6	0.05	0.05
Total	510			0.28	0.34

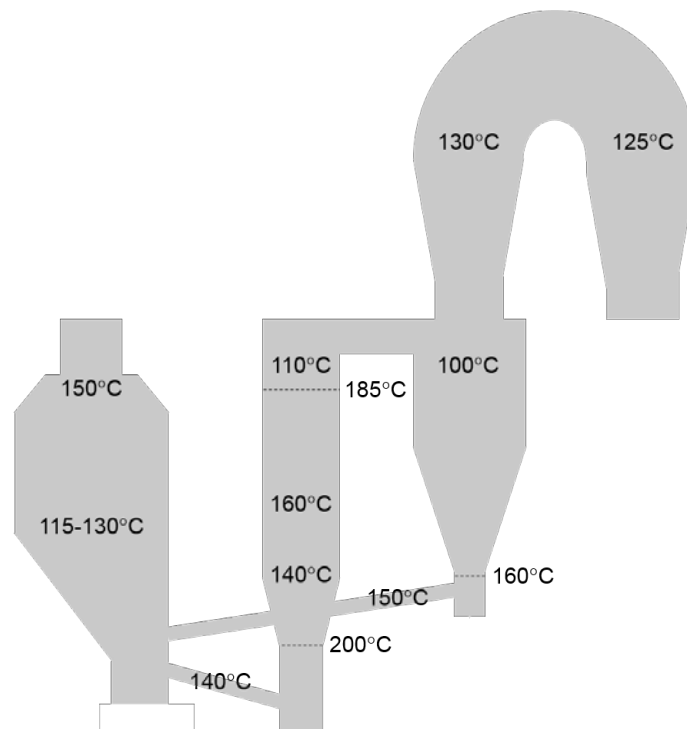


Figure A.2: Measured surface temperatures on the gasification section.

Table A.5: Approximated heat losses from radiation over the gasification section with a surface temperature of 140 °C and surrounding temperature of 30 °C.

	Area [m ²]	ϵ_{min} [-]	ϵ_{max} [-]	$\dot{Q}_{min}$ [MW]	$\dot{Q}_{max}$ [MW]
Gasifier	140	0.9	0.95	0.15	0.16
Comb ch	90	0.9	0.95	0.10	0.10
Post comb	210	0.9	0.95	0.22	0.23
Cyclone	70	0.9	0.95	0.07	0.08
Total	510			0.54	0.57

The measured value of the emissivity coefficient was 0.9 and the chosen interval was 0.9 – 0.95 based on typical values of emissivity coefficients for painted materials. In Table A.5, the approximated heat losses from radiation with a surface temperature of 140 °C and a surrounding temperature of 30 °C are presented.

By summing up the minimum heat losses from natural convection and for radiation and summing up the maximum heat losses, the total convective and radiative heat losses are assumed to lie within the interval 0.82-0.91 MW.

E ONE-AT-A-TIME SENSITIVITY ANALYSIS RESULTS FOR C2

Table A.6: Variables with $\gamma \geq 0.2$ from the OAT sensitivity analysis of the energy and carbon balance over the gasification process for C2.

	$\gamma_{\dot{Q}_{in}}$	$\gamma_{\dot{Q}_{out}}$	$\gamma_{\dot{n}_{C,in}}$	$\gamma_{\dot{n}_{C,out}}$
$\dot{m}_{biomass}$	0.94		0.96	
$\dot{V}_{PG}$		0.72		0.66
$\dot{V}_{FG}$		0.02		0.34
$Y_{CO_2,FG}$				0.29
$T_{HW,1}$		0.83		
$T_{HW,2}$		-0.73		
$T_{HW,3}$		0.95		
$T_{HW,4}$		-0.86		
$T_{RME,in}$		0.19		