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Comparação de Tecnologias de Gaseificação de Biomassa para Substituição do Gás Natural em Plantas de Pelotização de Minério de Ferro

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Comparison of Biomass Gasification Technologies for Natural Gas Substitution in Iron Ore Pelletizing Plants

ABSTRACT

The iron ore pelletizing process consumes high amounts of energy, including non-renewable sources, such as natural gas. Due to fossil fuel scarcity and increasing concerns regarding sustainability and global warming, at least partial substitution by renewable energy seems inevitable. Gasification projects are being successfully developed in Northern Europe, and large-scale circulating fluidized bed biomass gasifiers have been commissioned in e.g. Finland. As Brazil has abundant biomass resources, biomass gasification is a promising technology in the near future. Biomass can be converted into product gas through gasification. This work compares different technologies, such as air, oxygen and steam gasification, focusing on the use of the product gas in the indurating machine. The use of bio-synthetic natural gas is also evaluated. The main parameters utilized to assess the suitability of product gas were adiabatic flame temperature and volumetric flow rate. It was found that low energy content product gas could be utilized in the traveling grate, but it would require burners to be changed. On the other hand, bio-SGN could be utilized without any adaptions. Economical assessment showed that all gasification plants are feasible for sizes greater than 60 MW. Bio-SNG production is still more expensive than natural gas in any case.

Keywords: Biomass gasification, bio-synthetic natural gas, iron ore pelletizing.

Comparação de Tecnologias de Gaseificação de Biomassa para Substituição do Gás Natural em Plantas de Pelotização de Minério de Ferro

RESUMO

O processo de pelotização de minério de ferro consome grandes quantidades de energia, incluindo fontes não renováveis, como o gás natural. Devido à escassez de combustíveis fósseis e as preocupações crescentes a respeito do desenvolvimento sustentável e aquecimento global, a substituição ao menos parcial por fontes de energia renováveis parece inevitável. Projetos de gaseificação estão sendo desenvolvidos com sucesso nos países Nórdicos, e gaseificadores de biomassa industriais tipo leito fluidizado circulante já são uma tecnologia madura na Finlândia. Como o Brasil tem recursos de biomassa abundantes, a gaseificação é uma tecnologia promissora em um futuro próximo. Neste processo, a biomassa é convertida em gás de síntese. Este trabalho compara diferentes tecnologias, tais como gaseificação a ar, oxigênio e vapor d'água, com ênfase no uso do gás de síntese nos fornos de pelotização. O uso de gás natural sintético também é avaliado. Os principais parâmetros utilizados para avaliar a adequação do uso do gás de síntese foram temperatura adiabática de chama e a vazão volumétrica do gás. Verificou-se que o gás de síntese de baixo poder calorífico poderia ser utilizada no forno tipo grelha, mas isso exigiria mudanças no projeto dos queimadores. Por outro lado, o gás natural sintético poderia ser utilizado sem necessidade de quaisquer adaptações. A avaliação econômica mostrou que todas as plantas de gaseificação são viáveis para tamanhos superiores a 60 MW. Entretanto, a produção de gás natural sintético ainda é mais cara do que o gás natural, em qualquer caso.

Palavras-chave: gaseificação de biomassa, gás natural sintético, pelotização de minério de ferro.

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LIST OF SYMBOLS AND ACRONYMS

Roman Symbols

$\%A$	Ash content on a dry basis	[%]
AF	Annuity factor	[-]
c	Equipment cost	[EUR]
$\%C$	Carbon content on a dry basis	[%]
CC	Carbon conversion	[%]
ER	Equivalence ratio	[-]
GR	Gasifying ratio	[-]
$\%H$	Hydrogen content on a dry basis	[%]
H^0	Standard enthalpy	[J mol^{-1}]
H_f^0	Standard enthalpy of formation	[J mol^{-1}]
HHV	High heating value	[MJ kg^{-1}]
IRR	Internal Return rate	[-]
k	Equilibrium constant	[-]
L	Estimated energy loss	[%]
l	Plant's economic lifecycle	[years]
LHV	Low Heating value	[MJ kg^{-1}], [MJ kmol^{-1}], [MJ Nm^{-3}]
$\dot{m}$	Mass flow	[kg s^{-1}]
MM	Molar mass	[kg kmol^{-1}]
$\%N$	Nitrogen content on a dry basis	[%]
N	Number of moles per GJ of fuel	[kmol GJ^{-1}]
$\dot{n}$	total number of moles in the gas	[kmol kg$_{\text{wood}}^{-1}$]
n	Number of moles	[kmol]
$\%O$	Oxygen content on a dry basis	[%]
p	Pressure	[Pa]
R	Ideal gas constant	[J kmol^{-1} K^{-1}]
RH	Relative humidity	[-]
s	Equipment size	[-]
$\%S$	Sulfur content on a dry basis	[%]
S^0	Standard entropy	[J mol^{-1} K^{-1}]
SB	Steam biomass ratio	[-]

SOR	Steam oxygen ratio	[-]
T	Absolute temperature	[K]
t	Shomate parameter T/1,000	[-]
TD&IC	Total direct and indirect costs	[EUR]
$\bar{V}$	Molar volume	[Nm^3 $kmol^{-1}$]
$\dot{V}$	Volumetric flow rate	[Nm^3 s^{-1}]
V	Volume	[m^3]
w	Mass fraction on a dry basis	[-]
y	Molar fraction in wet basis	[-]

Greek Symbols

α	combustion air ratio	[-]
Δh	Sensible heat	[J mol^{-1}]
$\Delta_c H^0$	Standard enthalpy of combustion	[J mol^{-1}]
$\Delta_r H^0$	Standard enthalpy of reaction	[J mol^{-1} K^{-1}]
$\Delta_r G^0$	Gibbs free energy change of reaction	[J mol^{-1} K^{-1}]
$\Delta_r S^0$	Standard entropy change of reaction	[J mol^{-1} K^{-1}]
Φ	Energy flow	[MW]
η	Efficiency	[-]
θ	Temperature	[°C]
v_i	stoichiometric number	[-]

Superscripts

actual	Real value in the plant, e.g. gas from the balance
dried	Characteristic of the biomass after drying
eq	Equilibrium
flue gas	Flue gas
fuel	Final fuel (solid + gas)
harvested	Characteristic of the biomass after harvesting
input	Gasifier input
NG	Natural gas
product gas	Product gas
reactor	Added to the methanation reactor

SNG	Synthetic natural gas
stoich	Stoichiometric

Subscripts

DAF	Dry ash free basis
dry	Dry basis
NG	Natural gas
product gas	Product gas
ref	Reference
th	Thermal
wet	Wet basis

Abbreviations

AAEM	Alkali and Alkaline Earth Metals
AEI	Automation, electrification and instrumentation
AF	Annuity factor
ANEEL	Brazilian National Electric Energy Agency
ASU	Air Separation Unit
BFB	Bubbling Fluidized Bed
Bio-SNG	Bio-Synthetic Natural Gas
BRL	Brazilian Reais
CEPCI	Chemical engineering plant cost index
CFB	Circulating Fluidized Bed
CFD	Computational Fluid Dynamics
CHP	Combined Heat and Power
DAF	Dry Ash Free Basis
DEA	Diethanolamine
DFB	Dual Fluidized Bed
DNPM	Brazilian National Department of Mineral Production
DOE	Department of Energy
ECN	Energy Research Center of the Netherlands
EF	Entrained Flow

EPE	Brazilian Energy Research Company
ER	Equivalence Ratio
EUR	Euro
FB	Fluidized Bed
FICFB	Fast Internal Circulating Fluidized Bed
GC	Gas Chromatography
GHG	Greenhouse gases
GR	Gasifying Ratio
HHV	High Heating Value
HTHP	High Temperature/High Pressure
IEA	International Energy Agency
IRR	Interest Return Rate
ISBL	Inside battery limits
LHV	Low Heating Value
MC	Moisture Content
MEA	Monoethanolamine
MM	Molar Mass
MME	Ministry of Mines and Energy
NG	Natural gas
PAH	Polycyclic Aromatic Hydrocarbons
RH	Relative Humidity
SB	Steam Biomass Ratio
SOR	Steam Oxygen Ratio
SRF	Solid Recovered Fuel
TDC	Total direct costs
TD&IC	Total direct and indirect costs
TIC	Total indirect costs
USD	United States Dollar

1 INTRODUCTION

Iron and its derivative products are present, to some extent, in the production chain of all contemporary goods. The world's main iron ore producers, in decreasing order, are: China, Australia, Brazil, India, Russia and Ukraine (Jesus 2012). These six countries account for almost 90% of the world's total iron ore production. In 2011, the Brazilian iron ore production was 2.8 billion tons, which corresponds to 14.2% of the global production. According to data provided by the Brazilian National Department of Mineral Production (DNPM), the iron ore reserves in the country in 2011 were 29.6 billion tons, equivalent to approximately 17% of the world's total reserves.

Although large amounts of iron ore are extracted every year, only a fraction of this material can be directly utilized in pig iron manufacturing. Most of the crude iron ore is composed of small particles, which are not suitable for conventional reduction processes. Therefore, iron ore agglomeration processes, such as pelletizing and sintering, are fundamental in order to guarantee the maximum use of this natural resource.

The iron and steel industry is a great energy consumer, demanding globally about 28.6 EJ per year (International Energy Agency 2013). In particular, the iron ore pelletizing process consumes considerable amounts of electricity, natural gas (NG) and solid fossil fuels, such as coal and coke breeze. The Brazilian Energy Research Company (EPE) estimates that, on average, pelletizing processes consume 49 kWh of electricity per ton of iron pellets (EPE 2011), 129 kWh t^{-1} of NG, and 182 kWh t^{-1} of solid fossil fuels, mainly coke breeze (EPE 2012). It is also estimated that the mining and pelletizing sector consumed 1.3% of the country's primary energy.

Apart from electricity, which is mainly produced from renewable sources in Brazil, iron ore pelletizing utilizes large amounts of non-renewable energy in the firing stage. However, fossil fuel reserves are finite, and their scarcity might lead to high prices and uncertain supply in the future. In addition, greenhouse gas (GHG) emissions and consequences to global warming are a matter of great concern (Rockström et al. 2009). Also, in the specific case of NG, Brazil is still not self-sufficient. According to the Ministry of Mines and Energy (MME), 51% of the NG consumed in the country in 2013 was imported, from which 68% came from Bolivia

(MME 2013). Therefore, the Brazilian mining companies are interested in new technologies which are able to provide sustainable energy solutions for its processes.

Regarding NG substitution, one of the most promising technologies is biomass gasification followed (or not) by methanation reaction. In this process, charcoal and other suitable biomasses can be thermally converted into product gas, which mainly contains H_2, CO, CO_2 and CH_4. If a fuel with properties similar to NG is required, bio-synthetic natural gas (bio-SNG) can be obtained by e.g. catalytic hydrogenation of carbon oxides (Gao et al. 2012). In fact, the idea of utilizing product gas in iron pellet manufacturing has been published before. Zahl and Nigro (1979) proposed the utilization of low heating value product gas from lignite coal gasification in a rotary indurating machine. Further studies demonstrated that commercial quality iron pellets could be successfully produced from a product gas with low heating value (LHV) of approximately 6 MJ Nm^{-3} (Nigro 1982). Although preliminary results seemed promising, no further references were found, probably due to new NG reserve discoveries and the end of the oil crisis. Nevertheless, after many years this topic rises again, and the application of biomass-derived product gas in iron ore pelletizing might be a solution to sustainability and environmental issues in this sector.

In this context, the objective of this work is to investigate the substitution of NG with product gas/bio-SNG in the iron ore pelletizing process. Firstly, the current Brazilian scenario and the iron ore pelletizing process are briefly presented. Energy consumption is also addressed. Next, a literature review regarding biomass gasification and bio-SNG is provided. The fuel substitution analysis is developed under technical and economical perspectives, and the project feasibility is discussed. Finally, main findings and future work are summarized.

2 NATURAL GAS AND IRON ORE PELLETIZING

Iron ore mining, transporting and crushing operations generate large quantities of fine particles, which are not suitable for conventional reduction in blast furnaces. Therefore, agglomeration processes are required to recover these fine particles and assure the maximum economic mine yield. Pelletizing is one of the main iron ore agglomeration processes currently utilized and it is suitable for ultra-fine particles, ranging between 0.01 and 0.15 mm (Castro et al. 2004). This process is based on thermal treatment at high temperatures, usually varying between 1,200 and 1,400 °C. In general, iron pellets present better metallurgical properties than crude ore lumps; thus, this product is presently an essential feed stock in large blast furnaces, improving process energy efficiency.

In the pellet firing stage, the main fuels are NG or heavy oil (utilized in burners) and coal and/or petcoke, which are added to the green pellet blend. It is estimated that roughly 50% of the total energy input is provided by NG. In this chapter, the current NG scenario in Brazil is briefly presented, as well as pelletizing process details and energy consumption.

2.1 Brazilian Scenario

According to the United Nations International Merchandise Trade Statistics, iron ores and concentrates were the largest export commodity in Brazil in 2012, accounting for more than 30 million USD (United Nations Statistics Division 2013). From the 2.8 billion tons extracted in 2011, the DNPM estimates that 61.4% was composed by sinter feed and 26.6% by pellet feed (Jesus 2012). The national iron pellet production in 2011 was 62.4 Mt, from which approximately 90% were exported to China (51.0%), Japan (11.0%), Germany (5.0%), South Korea (4.0%) and The Netherlands (3.0%).

On the other hand, Brazil still relies on NG imports, which is one of the main fuels utilized by the iron pellet manufacturing process. According to the MME, approximately 50% of all NG consumed in Brazil in the last five years was imported, mainly from Bolivia (MME 2013). Proven NG reserves in 2012 are estimated at 459,178 million m^3, which would last 21 years at the current consumption rate (MME 2013). Inferred reserves were 472,155 million m^3 in 2011 (EPE 2012). The Brazilian NG grid in 2007 is shown in Figure 1; red lines correspond to pipelines under operation and green lines to those under construction.

Figure 1: Brazilian NG grid in 2007. (Source: ANEEL, n.d.)

According to the National Energy Balance, the share of NG in the country's primary energy consumption was 10% in 2011 (EPE 2012). In that year, 24.1 billion m^3 were produced in the country and 10.5 billion m^3 were imported (EPE 2012). Accounting all losses, NG actual consumption in Brazil in 2011 was 28.7 billion m^3, from which 40% was used by industry. The mining and pelletizing sector accounted for 789 million m^3, corresponding to a 7% share of the total industrial use. Despite the abundance of indigenous resources, no biomass was consumed as an energy resource by the mining and pelletizing sector in 2011.

NG prices have been rising since 2002, as shown in Figure 2. Data in the chart corresponds to prices to industrial consumers (including taxes) in some Brazilian states, converted to USD utilizing the average currency in each year (EPE 2012).

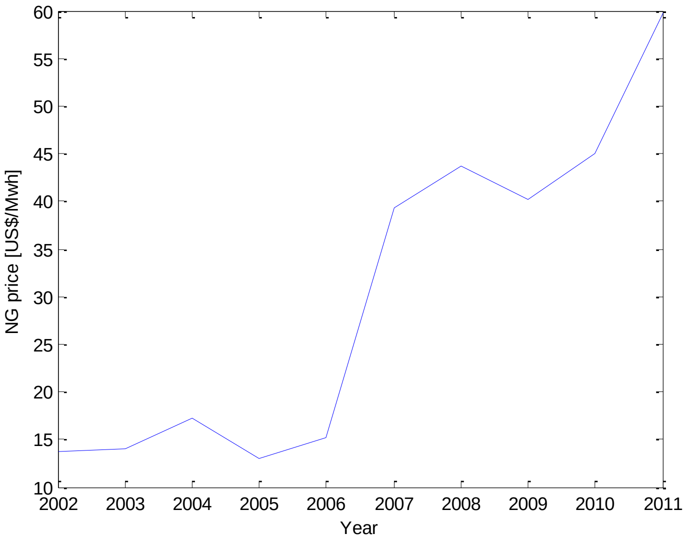

Figure 2: Average NG prices in Brazil. (Data source: EPE, 2012)

In Brazil, NG prices vary considerably depending on the consumption, geographical region and type of contract, e.g. if the gas is national or imported. Values from July 2013 are shown in Table 1. The US dollar exchange rate in the same month was considered as BRL 2.2522 per USD (MME 2013).

Table 1: NG prices in Brazil in July 2013. (Source: MME, 2013)

Region	**Contract**	**PRICE USD/MWh**		
		2,000 m^3/day	**20,000 m^3/day**	**50,000 m^3/day**
Northeast	National	53.14	51.17	49.84
Southeast	National	65.32	53.22	51.04
Southeast	International	65.32	53.22	51.04
South	International	61.64	55.85	54.66
Center-west	International	73.06	62.27	61.56

In addition to increasing prices, security of supply is also a concern in Brazil. Until 2019, a supply agreement with Bolivia assures the purchase of 30 million m^3 of NG per day (Kaup 2010), which corresponds to 40% of the daily availability. Back in 2006, hydrocarbon resource nationalization in Bolivia caused combined royalties and taxes to increase from 18% to 50% (Kaup 2010). At that time, it was not possible to modify the amount supplied. In addition, President Morales' government openly declared that the priority is to use NG to promote industrialization in his country. In this scenario, alternative technologies such as biomass gasification and bio-SNG production are fundamental in order to guarantee long-term NG supply in Brazil at reasonable costs and low environmental footprint.

2.2 Iron Ore Pelletizing Process

The iron ore pelletizing process is comprised of three main stages: (1) raw material preparation, (2) green pellets production and (3) thermal treatment of green pellets (Castro et al. 2004). Pelletizing plants usually receive concentrated ore sludge from the mining site. This sludge contains about 70%-w of solids, and it is sometimes delivered through pipelines (SAMARCO n.d.). Once the sludge arrives at the plant, it is further concentrated by a gravimetric tower, thickeners and filters. After this processes, the moisture content (MC) in the pellet feed is about 10%. Further comminution is also performed by a roller press in order to increase specific surface area and consequently improve pelletizing properties. The pellet feed is then blended with additives and coal. The aim of additives is to improve green pellets' cold agglomeration properties. Mainly limestone and bentonite are utilized for this purpose although organic binders might also be used.

There are different methods of producing green pellets, for instance: balling disc, drum or cone. Balling discs are the most common in Brazil (Castro et al. 2004). They consist of an inclined, rotary disc which promotes particle spherical agglomeration. Since homogeneity is important, green pellets are classified before thermal treatment; balls with diameter below 8 mm and above 18 mm are recycled to the process (SAMARCO n.d.).

Green pellets are sent to the induration furnace by a conveyor belt. There are different types of indurating machines, such as pellet shaft furnaces, grate kilns, which combines a traveling grate with a cylindrical rotary furnace, and traveling grates, which are recommended for large scale facilities (Castro et al. 2004). In Brazil, there is only one grate kiln currently in operation; all other pelletizing plants utilize traveling grate. In this system, green pellets are submitted to several stages: updraft drying, downdraft drying, preheating, firing (or combustion), post-firing and cooling (SAMARCO n.d.). Figure 3 shows a schematic picture of the iron ore pelletizing process. After firing, iron pellets are usually screened in order to assure the right granulometry, which is mainly between 8 and 16 mm diameter.

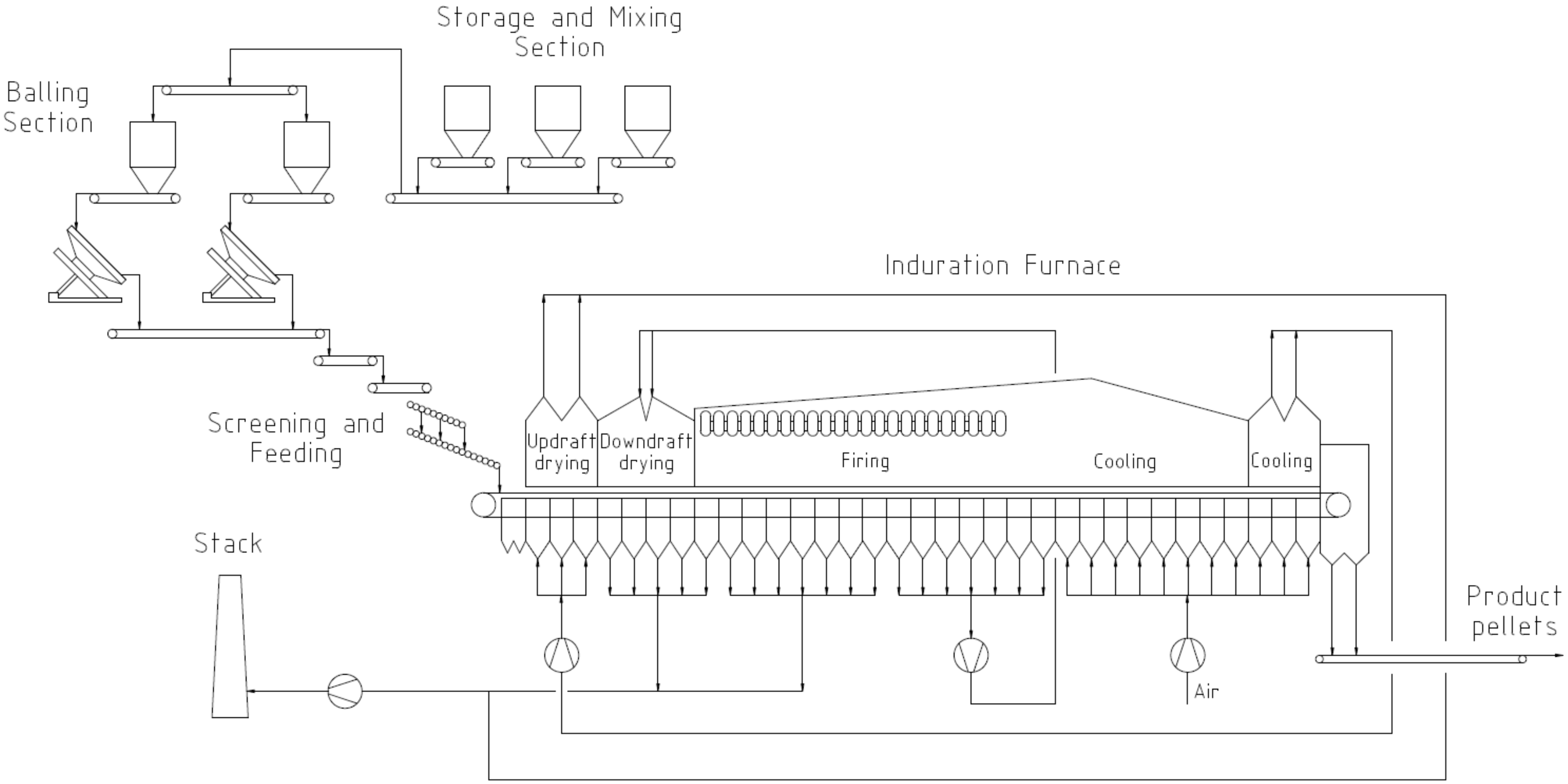

Figure 3: Scheme of iron ore pelletizing process.

2.3 Energy Consumption in Iron Ore Pelletizing

Iron ore pelletizing plants consume four main energy sources: electricity, NG and/or heavy oil and a solid fossil fuel. Electricity is extensively utilized in pumps, compressors, fans, balling discs, conveyors, traveling grates and illumination, among others. However, the electricity consumption in the induration furnace itself should be quite low. Fuel properties utilized in the calculations are summarized in Table 2.

Table 2: LHV and MC values utilized in energy consumption calculations.

Fuel	**LHV [kcal/kg]**	**LHV [MJ/kg]**	**MC [%]**
Coal	6 600	27.6	10%
Petcoke	8 550	35.8	10%
Heavy Oil	9 400	39.4	2.0%
NG*	9 300	39.0	0.0%

*kcal/m^3 and MJ/m^3

NG is the main fuel in the burners since heavy oil is being substituted due to environmental issues. Since 2010, more than 95% of all fuel in the burning zone is NG, which contributes to reduce CO_2 and SO_x emissions. Currently, NG average consumption in all kilns is 0.465 GJ t^{-1} of iron pellets (111 Mcal t^{-1}). Heavy oil consumption is negligible (0.011 GJ t^{-1} of pellets), and only occurs when there are problems with NG supply. Practically only NG is used in the third kiln, in which the consumption is the lowest, 0.405 GJ of NG per ton (96.6 Mcal t^{-1}).

The solid fuel, mainly coal and petcoke, is blended in the pellet feed and provides approximately 50% of the thermal energy required. The estimative is that 71% of the solid fuel is coal; however, the coal/petcoke ratio varies considerably between the three furnaces. The average consumption of coal and petcoke is 0.321 GJ t^{-1} (76.6 Mcal t^{-1}) and 0.133 GJ t^{-1} (31.8 Mcal t^{-1}) respectively. The third kiln is the one that utilizes less coal; it consumes 0.165 GJ t^{-1} (39.3 Mcal t^{-1}) of coal and 0.294 GJ t^{-1} (70.1 Mcal t^{-1}) of petcoke. The energy consumption in all kilns is summarized in Table 3.

Table 3: Indurating machine typical thermal energy consumption.

Fuel	**Furnace 01 [GJ/t]**	**Furnace 02 [GJ/t]**	**Furnace 03 [GJ/t]**	**Average [GJ/t]**
Coal	0.449	0.348	0.165	0.321
Petcoke	0.000	0.105	0.294	0.133
Total Solid	**0.449**	**0.453**	**0.458**	**0.454**
Heavy oil	0.016	0.016	0.000	0.011
Natural gas	0.488	0.503	0.405	0.465
Total burners	**0.504**	**0.520**	**0.405**	**0.476**
Total thermal	**0.953**	**0.973**	**0.863**	**0.930**

In order to compare, the Brazilian average specific energy consumption in the pelletizing process was extrapolated from the National Energy Balance (EPE 2012). As the data included mining and pelletizing sectors together, it was assumed that all NG, coal, coke and petcoke were utilized exclusively for pelletizing and other fuels were utilized only for mining. As a result, the average consumption is estimated at 0.466 GJ t^{-1} of NG and 0.656 GJ t^{-1} of solid fuel. Total thermal energy consumption is estimated at 1.122 GJ t^{-1}, which is 17% higher than values considered in this thesis.

3 CONVERSION OF BIOMASS TO FUEL GAS

In this chapter, practical issues regarding biomass gasification and bio-SNG production are discussed. Firstly, biomass harvesting, handling, and pretreatment are covered. Secondly, a theoretical background regarding biomass gasification is provided. In the sequence, possible gasifying media and gasifier models are presented. Special attention is given to circulating fluidized bed (CFB) gasification, which is the main large-scale option. In addition, downstream operations such as gas cleaning, methanation and gas upgrading are discussed. Finally, the commercial plants currently under operation and their main characteristics are listed.

3.1 Biomass Supply Chain

Fuel supply is possibly one of the main issues in biomass energy conversion systems. Cost-effective harvesting, transportation, comminution, storage, handling pretreatment and feeding are crucial to assure a project’s feasibility. In contrast to fossil solid fuels, biomass characteristics may vary according to location (soil), climate and season. Therefore, quality control is necessary in order to guarantee a fuel’s minimum requirements, especially regarding MC, particle size and ash characteristics. Quality control measures can be taken from the growing stage. For instance, type and quantity of fertilizers can influence chlorine and nitrogen contents of the biomass (Van Loo et al. 2008).

The supply chain characteristics depend on the type of biomass utilized, e.g. harvested or non-harvested. The latter generally includes granular materials, such as wood chips, rice husk and bark (Basu 2010). The former group refers to high MC fuels, such as straw, grass, and bagasse. Harvested biomass is beyond the scope of this work, in which the focus is eucalyptus chips and their utilization in large-scale CFB gasification facilities. For detailed information concerning other fuels, refer to Basu (2010) or Van Loo (2008).

3.1.1 Harvesting

Many options are available for cost-effective energy biomass harvesting, and some of them can be similar to those utilized in the pulp industry (Van Loo et al. 2008). For timber, extraction of whole trees is recommended, utilizing one-grip harvesters at the stump for cutting and delimbing (Kallio et al. 2005). Technologies for mechanized harvesting of short rotation eucalyptus forests (after 2 years) are also available (Kallio et al. 2005, Ghaffariyan et al. 2011, Cortez et al. n.d.). In such cases, feller-bunchers can be utilized for extraction of whole trees, including stems, branches, leaves and bark (Ghaffariyan et al. 2011). Grapple skidders are used to transport tree bunches to the chipping machine, usually located on the road side. Whole trees are chipped, without any delimbing operation. Wood chips are then transported by trucks to the final use.

Forestry residues, such as tops, bark, branches, and stumps represent a great potential for lowering costs, and should be considered if soil characteristics are favorable (Kallio et al. 2005). In case residues are utilized, the choice of adequate logistics is likely the main aspect to make their use economically attractive. Studies have evaluated the utilization of eucalyptus tops from the pulp industry for energy production in the South region of Brazil (do Canto et al. 2011). Tops represented from 8 to 10% of tree total volume. It was observed that the use of residues can produce from 94 to 162 times more energy than what is consumed to collect and comminute, depending on the MC. Therefore, efficiency can be increased if forestry residues are left in the field for some time to dry. In addition, leaves would fall during this period, decreasing nutrient losses.

There are many options for forest biomass harvesting. The harvesting method choice is site specific, varying according to terrain conditions, forest age (height and diameter of trees) and transport logistics. Detailed description of harvesting methods is beyond the scope of this work. For more information concerning eucalyptus harvesting for energy purposes in Brazil, refer to Do Canto (2009).

3.1.2 Transportation Logistics

As biomass has lower energy density than solid fossil fuels, transportation costs are considerably higher (Van Loo et al. 2008). Short distances and optimized fuel delivery logistics are important in order to ensure project’s feasibility. Several transportation methods are available, depending on the distance and type of biomass. In Austria, for woody biomass and distances up to 120 km, trucks are usually recommended. For very short distances of uncomminuted thinnings or for wood chips (e.g. less than 10 km), tractors with trailers might also be utilized. Shipping is usually a low cost option, allowing biomass procurement from longer distances.

However, these values are specific, and detailed studies should be performed within the Brazilian context.

The employed logistics for biomass supply is usually developed based on the location of size reduction, which is commonly defined by type of biomass. For instance, when early thinnings (15 to 20 cm diameter) and loose forest residues are being utilized, comminution at the harvesting site might be the best option (Kallio et al. 2005). For timber, large-scale centralized chipping at the plant site might be economical. In case of centralized chipping, residual biomass can be transported in bales or bundles. It is estimated that transportation of bales can save up to 10% of the costs if compared to the transport of wood chips (Van Loo et al. 2008).

3.1.3 Comminution

CFB gasifiers are considered quite tolerant regarding fuel particle size, and fine particles are not required. Thus, chipping should be the best comminution alternative once particle sizes between 5 and 50 mm can be obtained (Van Loo et al. 2008). Wood chips are approximately 8 mm thick, 25 mm wide and up to 55 mm long. As shown in Figure 4, there are two main types of chipping equipment: disc and drum. Although both types are able to adjust the chip size within a certain range, the former usually results in a more uniform particle distribution. If necessary, size classification can be performed after chipping utilizing trummels (Basu 2010).

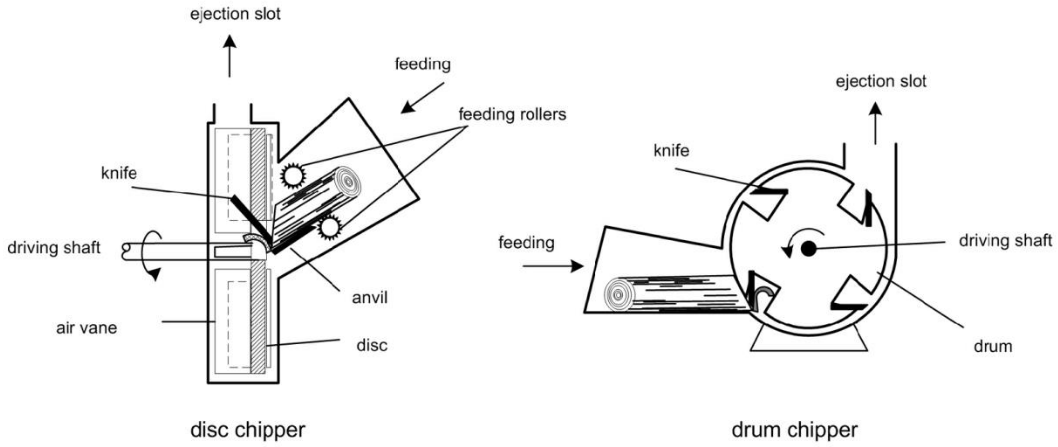

Figure 4: Disc and drum chippers. (Source: Marutzky & Seeger, 1999 cited in Van Loo et al., 2008, p. 64)

The energy required for chipping can be estimated at 1–3% of the total fuel energy content (Van Loo et al. 2008). MC might affect this energy consumption; for instance, higher water content reduces the friction factor, resulting in a lower energy demand. The chipper size depends on the delivering logistics (e.g. centralized or decentralized) and on the diameter of the logs/bundles. Chipper sizes and their respective characteristics are presented in Table 4.

Table 4: Characteristics of various sizes of chippers. (Source: Marutzky & Seeger, 1999 cited in Van Loo et al., 2008, p. 65)

Size	Productivity [m^3 bulk/h]	Diameter [cm]	Feeding system	Power [kW]
Small-scale	3–25	8–35	Manual or crane	20–100
Medium-scale	25–40	35–40	Crane	60–200
Large-scale	40–100	40–55	Crane	200–550

3.1.4 Drying

Fuel MC commonly varies according to the type of biomass, season, harvesting method and time of storage. Therefore, drying the material beforehand provides a more homogenous feedstock, which reduces the costs with process control (Van Loo et al. 2008). Despite beneficial reactions with water occur inside the gasifier, drying improves the overall energy efficiency for several reasons. Firstly, high MC means lower fuel net energy content. Secondly, heat is consumed to vaporize the water fed into the gasifier. Additionally, storage of wet biomass might increase the risk of biological degradation.

Fresh wood might contain between 30 and 60% moisture after cutting (Basu 2010). However, 2,260 kJ of thermal energy are required per each kg of water fed into the gasifier. Water excess also reduces the producer gas energy content. Surface moisture can be effectively removed by pre-drying. MC can be reduced to 10-20%, which is ideal for most gasification processes. Depending on climate conditions, simple outside storage might reduce the MC to 20-30% (Van Loo et al. 2008, Basu 2010); however, exposure to rainfall should be avoided.

Drying systems’ feasibility depends on the biomass price, plant size, and availability of waste heat (Van Loo et al. 2008). In all cases, economic analysis should consider possible indirect savings; for instance, impacts in feeding system size, energy consumption and product gas LHV. In gasification plants, drying medium can be hot product gas, combustion exhausts, hot air or steam, preferably heated by some source of waste heat (Brammer et al. 1999). During the drying process, biomass temperature should be kept below 100 ºC in order to avoid hazardous emissions and/or thermal degradation. In addition, it is recommended that the O_2 content in the drying medium is below 10%-v in order to minimize the risk of fire, especially if higher temperatures are used. Several dryer models are available for biomass. In this work, the most relevant continuous large-scale dryers suitable for wood chips are briefly presented. For more detailed information, refer to Brammer et al. (1999).

Belt dryers consist of a traveling grate transporting a thin material layer, usually from 2 to 15 cm (Brammer et al. 1999). The drying medium, usually hot air or flue gases, passes through the grate in a vertical direction. As the chamber is enclosed, upward and downward gas flow are possible. Single or multiple stage/pass equipment is available (Figure 5), depending on the amount of biomass to be treated, initial MC and available area. Drying medium temperature is commonly limited to 350 ºC and low velocity gas flow is preferable in order to avoid particle carriage.

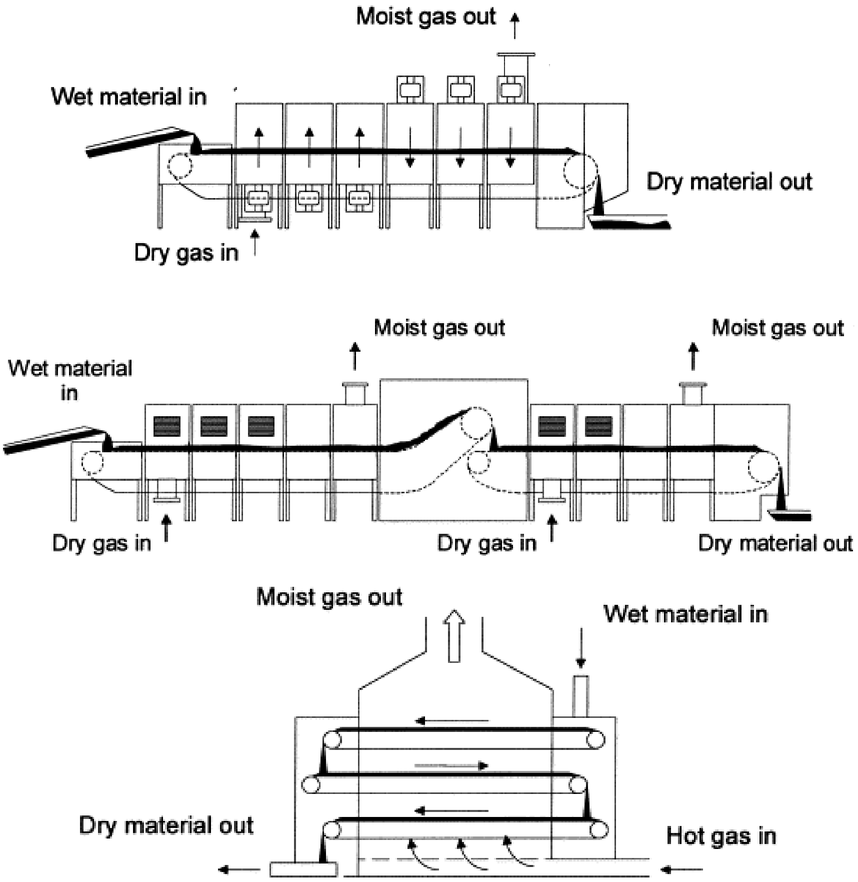

Figure 5: Belt dryer models; from top to bottom: single-stage single-pass, multiple-stage single-pass, and multiple-stage multiple-pass. (Adapted from: Brammer et al., 1999)

Rotary cascade dryers are commonly utilized for wood chips. The main advantage is technology maturity, which decreases risks (Brammer et al. 1999). Air or exhaust gases are the usual drying medium, and flow can be either co-current or counter-current. As shown in Figure 6, it is comprised of a rotary, slightly inclined cylinder equipped with longitudinal flights (Figure 7) able to lift the material and improve mixing. Typical sizes range between 4 and 10 m long, from 1 to 6 m diameter and rotating speed from 1 to 10 rpm. For safety reasons, inlet gas temperature is limited to approximately 250 ºC. Gas velocity is commonly between 2 and 3 m s^{-1} although it can range from 0.5 up to 5 m s^{-1}. Dryer efficiency can be between 50 and 75%, depending on the initial MC and gas inlet temperature. Exhaust gas cleaning is usually necessary in order to separate entrained particles, which can increase capital and operational costs.

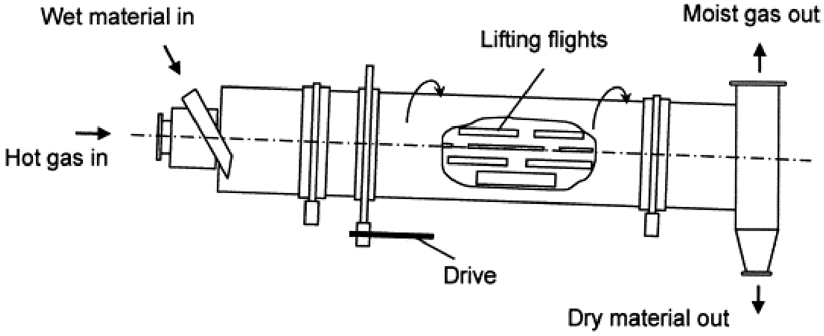

Figure 6: Rotary cascade dryer. (Source: Brammer et al., 1999)

Figure 7: Longitudinal flights in a frontal view. (Source: Brammer et al., 1999)

Other options for wood chip drying are also available. For instance, indirect steam-tube rotary dryers are based on conductive heating of biomass; however, they are only suitable when medium pressure saturated steam is available. Fluidized bed (FB) dryers are also appropriate for wood chips, as long as particle size is relatively uniform. Nevertheless, these two technologies are not extensively used in commercial, large-scale gasification plants yet. In practice, the most common configuration in biomass gasification plants is the belt dryer combined with hot air as drying media.

3.1.5 Storage

Biomass storage is fundamental in large-scale gasification plants in order to guarantee continuous operation. Adequate storage design might significantly reduce costs since large biomass volumes are utilized (Van Loo et al. 2008). Usually, separate long and short-term storage are necessary, the latter directly connected to the feeding system.

Mounds are the most common option for long-term storage. However, studies have shown that storing comminutes materials (e.g. wood chips) in piles may increase biological degradation. As a consequence, wood chips should be stored in piles for no longer than two weeks (Kallio et al. 2005). Long-term storage of uncomminuted material is a possible solution if longer storage is needed. Biological activity increases the pile's temperature and consequently the risk of fire, as well as dry mass losses and health hazards (Van Loo et al. 2008). High MC also increases microbial growth in the fuel; thus if storage of wood chips is necessary, MC should be kept below 30%-w (Van Loo et al. 2008, Kallio et al. 2005). Risk of self-ignition can be reduced when lower piles are utilized; for example, less than 8 m for bark (Van Loo et al. 2008). Piles should not be compacted and should contain homogeneous material. In order to prevent incidents, temperature and CO_2 release along the pile can be monitored. Outside piles should be avoided, especially in the rainy season due to increase in MC and leaching. Construction of paved storage surfaces is also advisable in order to avoid contamination with soil and stones.

Short-term storage provides fuel directly to the feeding system, usually after drying. Silos and hoppers are the most common apparatus. Adequate hopper design is important to ensure continuous operation. Common problems with hopper discharge (Figure 8) are: arching, ratholing and funnel flow (Basu 2010). Mass flow is the best discharge regime, in which the biomass is withdrawn homogeneously.

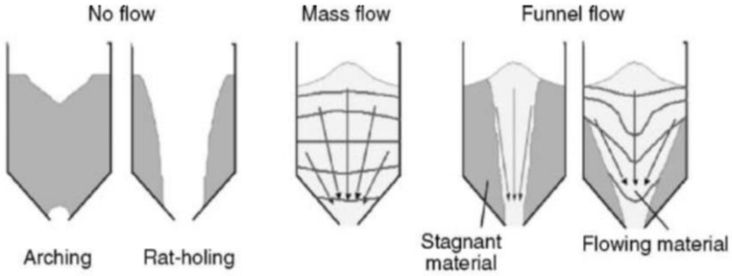

Figure 8: Flow regimes in hoppers. (Source: Basu, 2010)

In order to avoid problems with silo discharge, rotatory screw reclaimers (Figure 9) are recommended, for both outdoor and indoor storage. This equipment extracts the fuel by combining translation and rotation movements, allowing homogeneous discharge. The material is withdrawn in the center by a conveyor, from which it can be directed to the feeding system (METSO PAPER 2010).

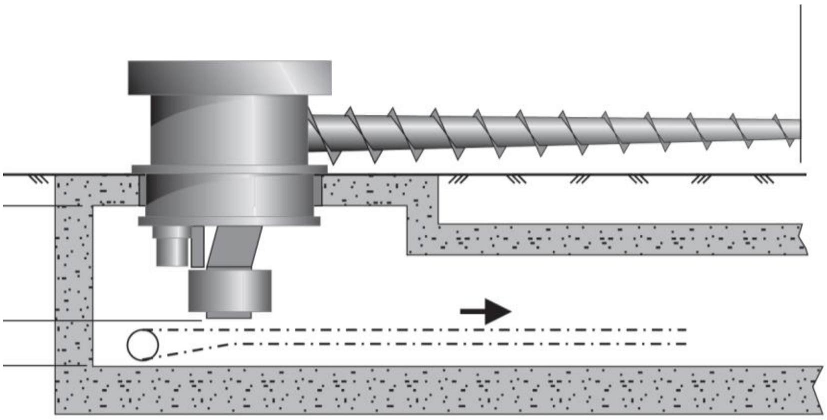

Figure 9: Rotatory screw reclaimer. (Adapted from: Metso Paper, 2010)

3.1.6 Handling

Intermediate biomass transportation between long-term storage, drying and short-term storage is also necessary. Many possibilities are available for this purpose. The main large-scale options, their advantages and disadvantages are summarized in Table 5.

3.1.7 Feeding Systems

Screw conveyors, gravity chutes or a combination of both (Figure 10) are likely the most common feeding system in commercial CFB gasifiers. Among the main advantages of screw conveyors are: avoidance of dust emissions, relatively small dimensions and low cost (Van Loo et al. 2008). They are only suitable for short distances, which is normally the case between short-term storage and gasifier. It requires the particle size to be smaller than 50 mm, which is also not an issue in CFB systems utilizing wood chips.

Table 5: Biomass handling systems. (Source: Van Loo et al., 2008)

Handling system	**Advantages**	**Disadvantages**
Crane	Adequate for wood chips; fully automated.	Problems with nonhomogeneous materials, e.g. mixtures.
Belt conveyors	Suitable for long distances with bulk materials or unit loads; simple and inexpensive construction; allows weight measurement.	Not suitable for inclined conveying; dust emissions; sensitive to temperature and dirt accumulation.
Tube-rubber belt conveyors	Avoid dust emissions; possibility to overcome height differences; distances up to 2,000 m.	Not suitable for long and sharp particles.
Chain conveyor	Suitable for sawdust, bark and wood chips; horizontal or inclined transport, up to 90º; flexible regarding particle size; material charging and discharging possible at any point.	Dust emissions, unless totally encased; relatively high power demand; low conveying capacity; high wear of chain and coating.
Bucket elevators	Inclined and vertical conveying of small and medium-sized particles; capacity up to 400 t h^{-1}; max. transportation height 40 m.	Particle size is limited by bucket dimensions; dust emissions and dirt accumulation, especially in high speeds.

The main disadvantages of screw conveyors are: relatively high energy consumption, high sensitivity to impurities (e.g. metals or stones), bark, and inhomogeneous particle size. There are several types of screw feeders such as variable-pitch, variable diameter, wire, and multiple screws (Basu 2010). Adequate design should avoid plugging and jamming, ensuring uninterrupted operation.

Although gravity chutes are extremely simple, they require the pressure in the gasifier to be lower than in the feeding system; otherwise, fuel is blown back (Basu 2010). In addition, fuel might accumulate near the chute outlet, resulting in corrosion problems. In order to avoid corrosion, extensions may be utilized, as shown in Figure 10; however, insulation and cooling are required. Gravity chutes are usually combined with a screw feeder because they are not metering devices.

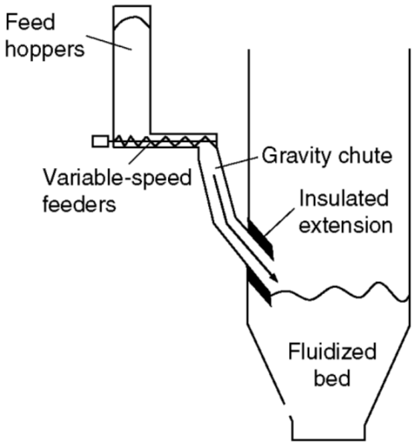

Figure 10: FB feeding system. (Adapter from: Basu, 2010)

More than one feeding system is always necessary for large-scale gasifiers in order to improve fuel distribution in the reaction vessel. For instance, in bubbling fluidized beds (BFB) one screw conveyor usually covers up to 3 m^2 of the bed (Basu 2010). In CFB this number varies, ranging between 10 and 40 m^2 per feeder for different types of coal (Basu 2006).

3.2 Biomass Gasification

Gasification can be defined as a series of thermo-chemical reactions which are able to transform carbonaceous raw materials (e.g. coal, oil, coke, biomass and wastes) into a combustible gas, called product gas or syngas, depending on the application. In the present work, the former term is adopted since the focus is on the gas combustion and not on its use as a feedstock for synthesis processes. Although biomass gasification is a well-established technology, the interest in it has only increased in recent years due to sustainability issues and global warming concerns.

In biomass gasification, renewable carbon resources, such as charcoal, wood chips, energy crops (e.g. short rotation coppice, miscanthus, switchgrass, etc.), forestry residues (e.g. bark, tops, branches, small logs, etc.), agricultural wastes (e.g. cobs, straw, bagasse, stalks, shells, etc.), and other wastes (plastics, municipal solid waste, sawdust, etc.) are transformed into combustible gases. In general, the carbon source (often called fuel) reacts with a gasifying medium, which can be air (or oxygen), steam or a combination of both. Each gasification medium has its own reaction mechanism, which usually defines the reactor's heat source. For instance, when air or oxygen is utilized, biomass partial oxidation provides the heat required by endothermic reactions. In these cases, the system can be considered autothermal. On the other hand if only steam is utilized, an external heat source must be utilized. In such situations, the system is called allothemal. The gasifying medium and its amount have considerable impact on the gasification process, operation and resulting product gas. Therefore, this matter is discussed in more detail in the next section.

The main combustible components in the product gas are H_2 and CO although CH_4 and small hydrocarbons containing 2-3 atoms of C are formed in minor concentrations. However, non-combustible products are also present in this mixture, such as H_2O, CO_2 and possibly N_2, depending on the reagents and/or reactor design. Undesirable products from biomass gasification such as tars and alkali are formed as well.

Although this thermo-chemical process is generally called gasification, several partially overlapping phenomena occur inside a gasifying reactor. A gasifier can be divided into preheating, drying, pyrolysis, char gasification and combustion zones (Chen et al. 2003). These stages are schematically represented in Figure 11. Depending on the gasifier design, there might also be tar cracking and shift reaction zones.

Preheating and drying stages are of special importance for biomass gasification. Although pre-drying is usually performed, final drying can only occur inside the gasifier, at temperatures above 100ºC. Volatiles are also released in the preheating zone, when temperatures are above 200ºC.

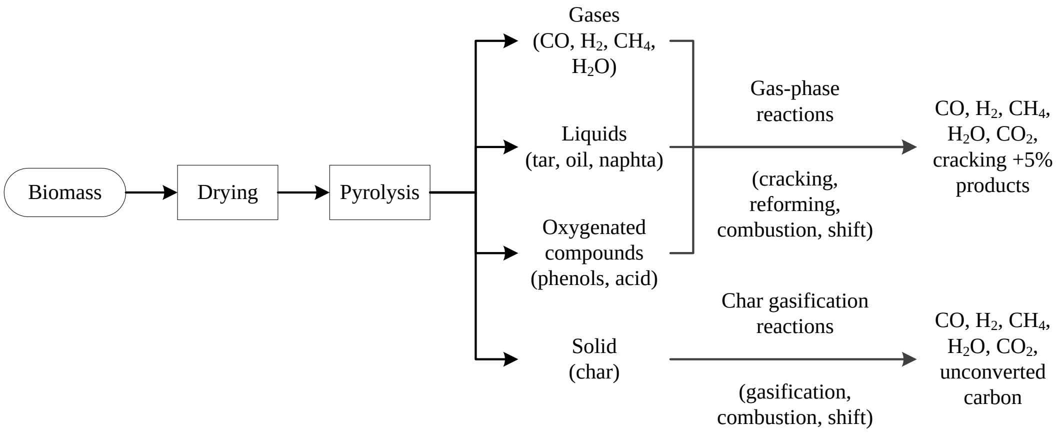

Figure 11: Gasification stages. (Adapted from Basu, 2010, p. 119)

Thermal cracking of large hydrocarbons present in the solid phase occurs in the pyrolysis zone, forming char, liquids and gases. Further decomposition of liquids also occurs in this stage, in which condensable and non-condensable gases are produced (Basu 2010). Condensable gases are generically called tars, and should be avoided as much as possible since their condensation causes clogging in downstream equipment. Measures for tar reduction are discussed later in this chapter.

At the gasification zone, reactions between pyrolysis products and gasifying agents occur, in both solid and gaseous phases. The most typical gasification reactions, as well as combustion reactions, are listed in Table 6, together with their respective enthalpies of reaction ($\Delta_r H^0$) at 25ºC. Enthalpy data in Table 6 were compiled from several sources by Basu (2010, p. 121).

Char gasification reactions are kinetically the limiting ones. This is explained by the fact that they occur in solid phase, and therefore mass transfer plays an important role in the reaction rate. Char reaction speed also depends on the gasifying agent. Reactions with oxygen are the fastest ones, and thus the remaining oxygen is quickly consumed, favoring CO formation (Basu 2010). Those are followed by the char-steam reaction, which may be from 10^3 to 10^5 times slower. The char-carbon dioxide reaction (Boudouard) is even slower, from 10^6 to 10^7 times if compared with char-oxygen. At temperatures below 1,000 K (727 ºC), the rate of Boudouard reaction can be considered insignificant. Nonetheless, the slowest of all char reactions is hydrogasification, which can be 10^8 times slower.

Table 6: Main gasification reactions. (Data compiled by Basu, 2010, p. 121)

Reaction Type	**Reaction**	**$\Delta_r H^0$ at 25°C [kJ/mol]**
Carbon Reactions		
R1 (Boudouard)	$C + CO_2 \leftrightarrow 2\ CO$	+172
R2 (water-gas or steam)	$C + H_2O \leftrightarrow CO + H_2$	+131
R3 (hydrogasification)	$C + 2\ H_2 \leftrightarrow CH_4$	-74.8
R4	$C + ½\ O_2 \rightarrow CO$	-111
Oxidation Reactions		
R5	$C + O_2 \rightarrow CO_2$	-394
R6	$CO + ½\ O_2 \rightarrow CO_2$	-284
R7	$CH_4 + 2\ O_2 \leftrightarrow CO_2 + 2\ H_2O$	-803
R8	$H_2 + ½\ O_2 \rightarrow H_2O$	-242
Water-gas Shift Reaction		
R9	$CO + H_2O \leftrightarrow CO_2 + H_2$	-41.2
Methanation Reactions		
R10	$2\ CO + 2\ H_2 \rightarrow CH_4 + CO_2$	-247
R11	$CO + 3\ H_2 \leftrightarrow CH_4 + H_2O$	-206
R12	$CO_2 + 4\ H_2 \leftrightarrow CH_4 + 2\ H_2O$	-165
Steam-reforming Reactions		
R13	$CH_4 + H_2O \leftrightarrow CO + 3\ H_2$	+206
R14	$CH_4 + ½\ O_2 \rightarrow CO + 2\ H_2$	-36.0

Combustion is usually faster than gasification given the same pressure and temperature. Therefore, the amount of air or oxygen utilized should be carefully determined since unnecessary combustion results into heat wastage and low energy content product gas. Shift reaction is also important for the process, but at temperatures above 1,000 ºC, it rapidly reaches equilibrium, compromising the hydrogen yield (Basu 2010). Methanation reactions are very important in bio-SNG production; nonetheless, they are not favored inside the gasifier, and only occur to a minor extent under normal operating conditions. As the use of catalysts is required, methanation is often carried downstream. This process is also discussed later in this chapter.

3.3 Gasifying Agent

Gasifying media nature and amount are the main parameters influencing the product gas composition and energy content (Bacovsky et al. 2010). Currently, the most utilized gasification agents are air (or oxygen), steam, or a mixture of both. Table 7 shows composition and LHV ranges for different gasifying media.

Table 7: Product gas properties for different gasifying media. (Data source: Bacovsky, 2010)

Gasifying agent	**Air**	**O_2**	**Steam**
H_2 [%-vol, dry]	11-16	23-28	35-40
CO [%-vol, dry]	13-18	45-55	22-25
CO_2 [%-vol, dry]	12-16	10-15	20-25
CH_4 [%-vol, dry]	2-6	<1	9-11
N_2 [%-vol, dry]	45-60	<5	<1
LHV [MJ/Nm^3]	4-6	10-12	12-14

3.3.1 Air or Oxygen

Air is the most utilized gasifying media in biomass gasification. Its main advantage is heat generation inside the reactor, enabling an autothermal process. In this situation, the fuel is partially oxidized, avoiding external heating equipment.

Purified oxygen or simply air can be utilized as an O_2 source. Although air is an inexpensive option, it introduces considerable amounts of nitrogen in the system, causing product gas dilution and consequent reduction in the final LHV. The use of O_2 may increase process efficiency since lower amounts of gas would need to be heated inside the gasifier. On the other hand, oxygen purification can be costly and energy consuming.

The equivalence ratio (ER) in gasification processes indicates the oxygen deficit in the system. The ER in biomass gasification usually varies between 0.2 and 0.3 (Basu 2006), and it is defined by the following equation:

$$ER = \frac{O_2\ content\ in\ air\ supply}{O_2\ required\ for\ complete\ combustion} \tag{1}$$

Although increasing the ER might increase the temperature, char conversion and gas yield, it decreases the product gas' LHV for several reasons. Firstly if air is being utilized, N_2 dilution increases as the ER increases. In addition, it favors oxidation reactions to form CO_2 and H_2O at the expense of H_2 and CO (Devi et al. 2003). Therefore, there is an optimum ER in which char conversion and oxidation are balanced, leading to the maximum product gas LHV.

Table 8: Oxygen supply alternatives. (Adapted from: Gullichsen et al., 1999)

System	Adsorption	Cryogenic	Purchase
Operating temperature	Ambient	-190ºC	-190ºC
Maximum purity [%]	95	99.9	99.9
Usual purity [%]	90-93	98-99.5	98-99.5
Pressure [kPa]	~ atm	70 to 1 400	Up to 1 600
Specific power at site [kWh/t O_2]	265	300	50*

*Vaporization only

If enriched air or pure oxygen is utilized, air separation units (ASU) are required. Two technologies are currently utilized in large-scale pulp production: the cryogenic process and molecular sieve adsorption (Gullichsen et al. 1999). Purchase in liquid state is also possible. A comparison between these options is presented in Table 8. According to Hingman et al. (2008), the decision on the oxygen supply is not only based on technical features, but also on commercial/economic aspects. The use of steam might be necessary in cases where pure oxygen is utilized in order to control temperature inside the reactor.

3.3.2 Steam

Steam gasification results in a H_2-rich product gas, usually over 50%-vol. (Devi et al. 2003). This is an advantage if methanation is performed downstream since this process requires an H_2/CO ratio of at least 3:1 (Gao et al. 2012). On the other hand, steam gasification is an endothermic process, and heat supply complexity is one of its main drawbacks.

The steam-biomass ratio (SB) is defined as the ratio between the water input (including gasifying media and biomass MC) and the biomass input on a dry ash free (DAF) basis. Some studies have shown that the optimal SB ratio for dual fluidized bed (DFB) gasifiers is 0.5 kg/kg (Hofbauer et al. 2000). The SB is defined as:

$$SB = \frac{Total\ H_2O\ input\ [\mathrm{kg/h}]}{Biomass\ input\ [\mathrm{kg_{DAF}/h}]} \tag{2}$$

When the SB increases, H_2 and CO_2 formation are favored, whereas CO concentration decreases (Herguido et al. 1992 cited in Devi et al. 2003, p. 129). Therefore, the product gas LHV decreases with increasing SB. On the other hand, increasing SB improves carbon conversion only up to a certain point (Hofbauer et al. 2000). It is estimated that the water conversion in DFB gasifiers is usually as low as 10-15%. Therefore, increasing the SB might decrease the process thermal efficiency since considerable amounts of sensible heat are lost in water removal from the product gas (Corella et al. 2007).

3.3.3 Steam/O_2

Gasifying with a mixture of steam and oxygen (or air) brings together advantages of both media. Firstly, the presence of oxygen provides the necessary heat. Secondly, costs with oxygen or dilution of nitrogen decreases due to the use of steam. Similarly to the SB, the gasifying ratio (GR) indicates the quantitative relation between media and biomass, as follows (Devi et al. 2003):

$$GR = \frac{Total\ H_2O + O_2\ input\ [\mathrm{kg/h}]}{Biomass\ input\ [\mathrm{kg_{DAF}/h}]} \tag{3}$$

Following the same trend as ER and SB parameters, increasing GR results in a product gas with lower energy content due to formation of more CO_2 instead of H_2 and CO (Devi et al. 2003).

3.4 Biomass Gasifiers

Several gasifier models have been developed in the past 20 years (Bridgwater 2002). Regarding biomass gasification, reactors can be divided into three main groups: fixed (or moving) bed (e.g. downdraft and updraft), FB (e.g. BFB, CFB, DFB), and entrained flow (EF) gasifiers. Each model is suitable for certain fuel quality and input, as shown in Figure 12. This section aims at briefly presenting the most relevant biomass gasifiers. Detailed information concerning this matter can be found elsewhere, such as Olofsson et al. (2005).

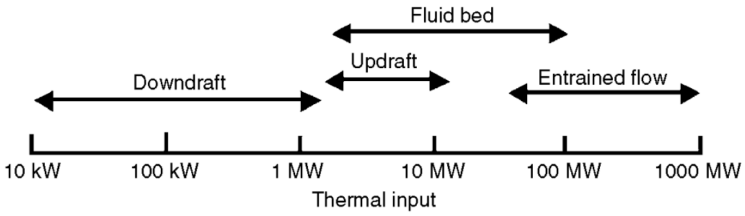

Figure 12: Typical size of biomass gasifiers. (Source: Basu, 2010)

3.4.1 Fixed Bed Gasifiers

Fixed or moving bed gasifiers are probably the oldest gasification technology (Olofsson et al. 2005). The main advantage of this type of gasifier is its simplicity and consequent low capital cost. However due to temperature homogeneity and process control issues, scaling-up to more than 10 MW_{fuel} is not possible. Additionally, ash melting is also a concern since high temperatures are usually achieved. Air is usually the gasifying medium since costs of oxygen or indirect heating would compromise the economic feasibility of such equipment. Three models are currently relevant (Figure 13): updraft, downdraft and crossdraft gasifiers.

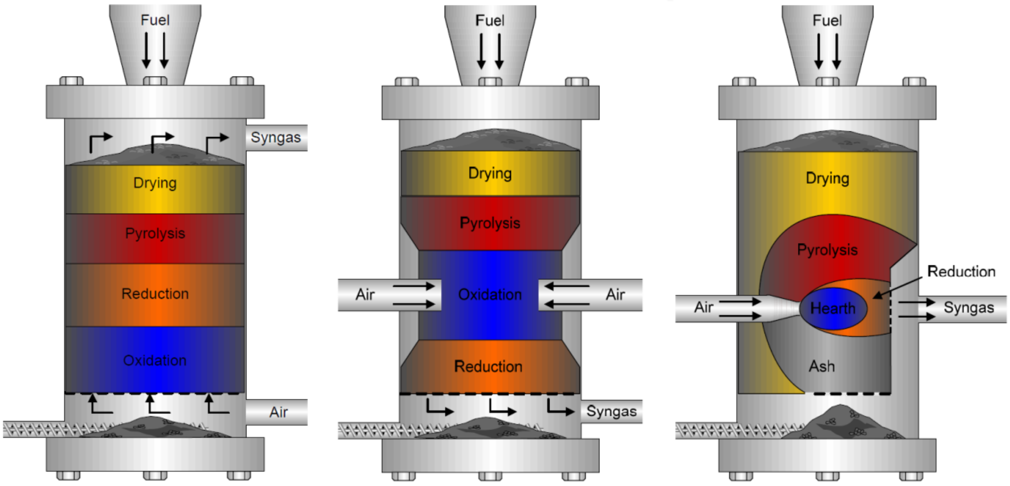

Figure 13: From left to right: updraft, downdraft and crossdraft gasifiers. (Source: Olofsson et al., 2005)

In updraft models, fuel is fed from the top and air from the bottom. The product gas is collected at the top of the equipment. As combustion reactions occur in the lower part, a gas with relatively high energy content can be obtained – from 4 to 5 MJ Nm^{-3} when air is utilized (Olofsson et al. 2005). Another advantage is its relative low sensitivity to fuel size and MC. However, the product gas' tar content is extremely high, reaching 10 to 20%-w.

In crossdraft gasifiers, air is blown in the middle part, and the product gas is collected on the opposite side (Olofsson et al. 2005). Despite its simple design, the product gas has low energy and high tar content. In addition, the product gas leaves the reactor at high temperatures, and considerable amounts of sensible heat might be lost. Therefore, this model is possibly the least attractive among fixed bed gasifiers.

Downdraft gasifiers correspond to approximately 75% of all models available in the market (Bridgwater 2003). In this case, air is fed in the middle and the product gas is collected from the bottom. They are probably the most suitable gasifier for small-scale applications due to low investment and maintenance costs. However, the scale-up is limited to 500 kg h^{-1} (2.5 MW_{fuel}) of wood input due to control and bed height issues. Although relatively low tar content product gas is obtained, LHV is usually low since a large part of the fuel is oxidized (Olofsson et al. 2005). As a result, the overall efficiency is quite low. Another disadvantage is the high particle content of the gas (McKendry 2002).

3.4.2 Fluidized Bed Gasifiers

FB systems consist of a granulated bed and a gaseous stream, which is forced through it. When the gas reaches a certain velocity, it becomes capable of moving the bed material, and the gas-solid mixture starts behaving like a fluid (Rayaprolu 2010). In general, this technology can be utilized in solid fuel combustion and gasification processes with considerable increase in efficiency since intense turbulence improves homogeneity and process control. Moreover, FB gasifiers are usually flexible regarding fuel as long as suitable particle size and MC can be achieved.

Usually, FB gasifiers are divided into two parts: bed and riser. The bed consists of some inert material, commonly quartz sand with 250 µm particle size (Olofsson et al. 2005). However, active bed materials such as dolomite might also be utilized. Catalytic bed materials are discussed later in this chapter. Combustion reactions usually occur in the bed. The riser (or freeboard) is the space above the bed, where most of gas-phase reactions take place.

There are two types of FB gasifiers: BFB and CFB. BFB gasifiers (Figure 14) operate with lower gas velocities, usually between 2 and 3 m s^{-1} (Olofsson et al. 2005). In such velocities, bed particles are moved but not carried through the riser. As a result, turbulence is much lower, as well as required fan power. Despite the presence of a cyclone, the product gas usually has large amounts of suspended particles. On the other hand, tar content is relatively low, though not as low as e.g. downdraft gasification. Depending on the application, gas cleaning and/or tar reforming may be necessary.

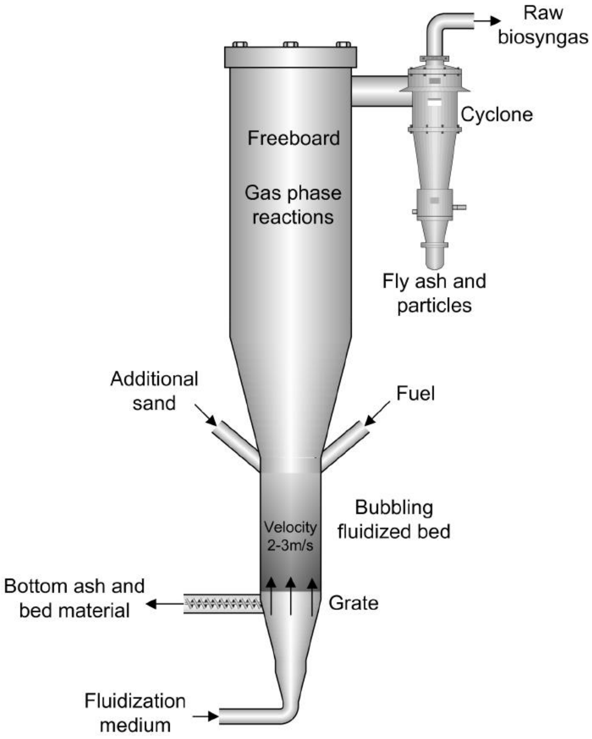

Figure 14: BFB gasifier. (Source: Olofsson et al., 2005)

CFB gasifiers are characterized by higher gas velocity, which increases turbulence. The bed material is partially carried by the fluidization medium. Suspended bed material, together with most of the ash and unreacted char are retained by the cyclone and recycled to the gasifier. Consequently, particle residence time is longer and carbon conversion increases.

According to Bridgewater (2003), “for large-scale applications the preferred and most reliable system is the circulating fluidized bed gasifier”. However, conventional CFB systems are only suitable for air or oxygen. For steam gasification, external heat supply is necessary. A promising technology is the DFB gasifier, which consists of two separate zones: steam gasification and air combustion. As these two technologies are of special interest in this study, they are discussed in detail later in this chapter.

3.4.3 Entrained Flow Gasifiers

In EF gasifiers, a mixture of fuel and gasifying medium (e.g. oxygen or a mixture of steam/O_2) is injected in the reactor, generating a high temperature flame (Basu 2010, Olofsson et al. 2005). In EF models, the fuel must be in the form of gas, powder or slurry (Olofsson et al. 2005). Therefore, there are still some issues in utilizing biomass in this type of reactor since it must be either ground to powder, pyrolysed to liquid and/or gas, or transformed into charcoal slurry. In addition, the high alkali content in biomass ash makes it harmful to refractory and metal parts.

As ash melting is not a concern in EF gasifiers, temperatures between 1,200 and 1,400ºC and pressures from 20 to 70 bars are commonly utilized (Basu 2010, Olofsson et al. 2005). Under such conditions, the product gas is almost free of tars. On the other hand, large amounts of sensible heat might make this process inefficient if proper heat recovery systems are not available (Olofsson et al. 2005).

There are several types of EF gasifiers. The CHOREN's Carbo-V® process is the most successful EF model utilizing biomass (Higman et al. 2008). In this process, fuel is pre-gasified at low temperature, and a mixture of pyrolysis gas, tars and char are further gasified in a down EF gasifier (Figure 15).

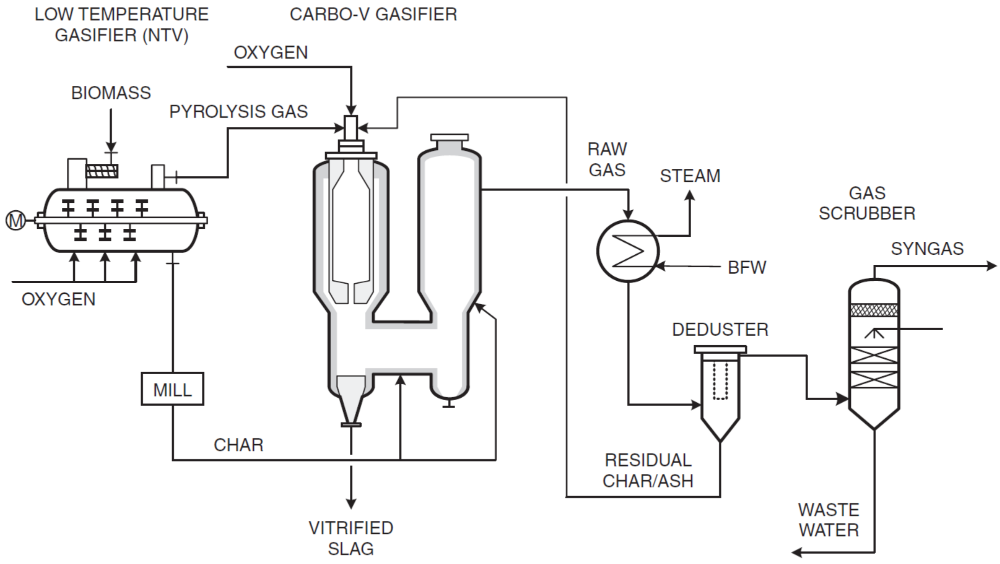

Figure 15: Choren process for biomass gasification. (Source: Higman et al., 2008)

3.5 CFB Gasifiers

CFB technology (Figure 16) is currently the main commercial alternative for large-scale oxygen/air biomass gasification. It is especially suitable for biomass since it provides long particle residence times and has relative high tolerance to the presence of volatiles (Basu 2010).

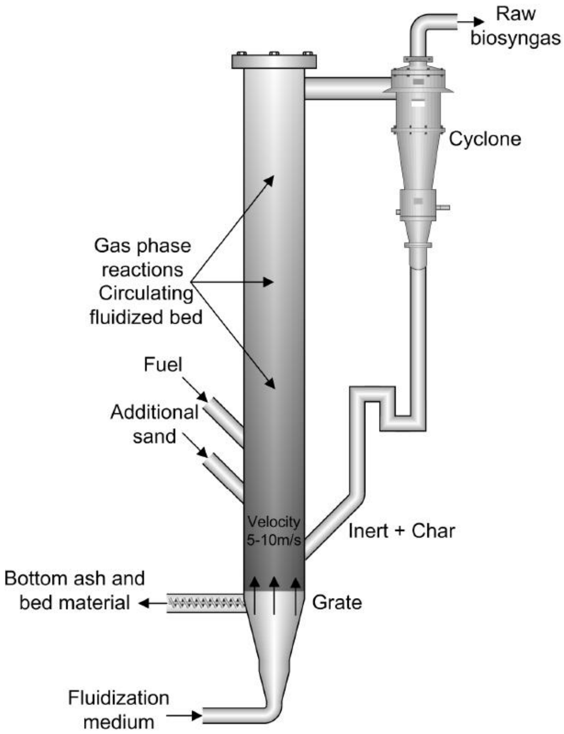

Figure 16: CFB gasifier. (Source: Olofsson et al., 2005)

In CFB gasifiers, the fluidization medium velocity is relatively high, usually between 5 and 10 m s^{-1} (Olofsson et al. 2005). Therefore, bed particles are partially carried by the gas stream and dispersed over the riser. These solids are then separated from the product gas by a cyclone and recycled to the reactor. The hydrodynamic regime inside a CFB is called a fast fluidized bed (Basu 2010). In such a regime, relatively uniform temperatures can be achieved inside the reactor, typically between 800 and 1,000 ºC for biomass. In addition, intense turbulence effectively promotes heat and mass transfer. CFB is also flexible regarding different fuels such as agricultural residues, wood and solid recovered fuels (SRF). Within a certain range, CFB gasifiers are tolerant to variations in particle size and MC (Olofsson et al. 2005). Tar content in the product gas can be considered quite low although not as low as downdraft or EF gasifiers. On the other hand, the product gas might contain relatively high amounts of particulate matter. Additionally, bed agglomeration is still an issue, which limits the operating temperature.

Appropriate bed material extraction must be performed periodically in order to avoid bed agglomeration. Despite the tolerance to particle size, larger particles require a higher fluidization velocity, which increases the energy consumption and may result in equipment erosion.

If high H_2-content product gas is required, air and oxygen gasification are not recommended. This is the case of bio-SNG production, in which the methanation stage is performed downstream. In this situation, DFB steam gasification is the best solution. In DFB gasification, two FB reactors are interconnected. In the first one, the gasification process occurs and only steam is injected as the fluidization/gasifying medium. In the second reactor, non-gasified char and tars are burnt, and only air is utilized as fluidizing gas. The bed material circulates between the two reactors, acting as a heat carrier. Figure 17 shows one possible configuration for this concept, which utilizes two CFB reactors. However, other configurations are also possible, e.g. CFB gasifier and BFB combustor.

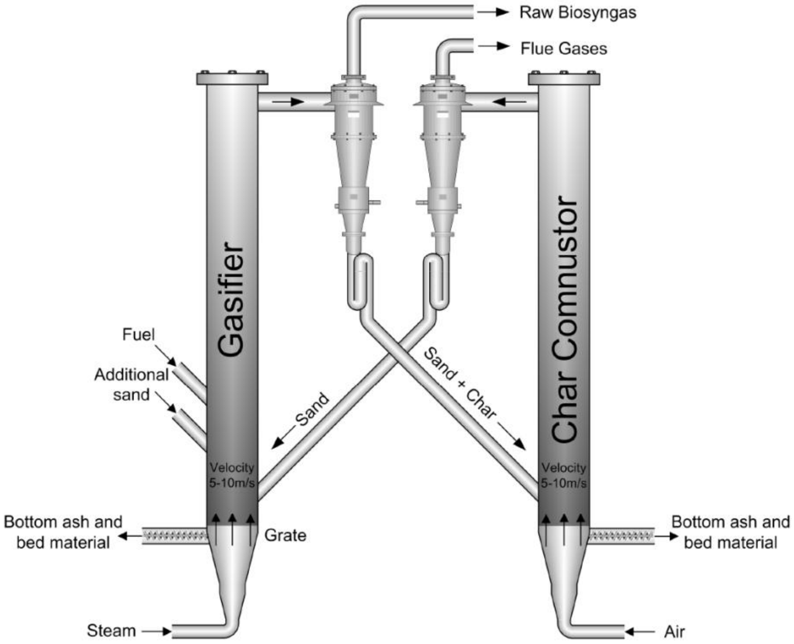

Figure 17: DFB gasifier. (Source: Olofsson et al., 2005)

The product gas is almost free of N_2, resulting in higher energy content. In addition, equilibrium calculations and practical data have shown that methane content is higher than in air/oxygen gasification (Olofsson et al. 2005), which also contributes to the increasing product gas' LHV. On the other hand, the use of two interconnected reactors increases construction complexity and investment costs, making these systems feasible only for large-scale processes. In addition, it is estimated that only 10 to 15% of the steam is actually converted in DFB gasifiers. Therefore, the product gas must be dried and high amounts of sensible heat might be lost in this stage, decreasing the process overall efficiency (Hofbauer et al. 2000, Corella et al. 2007).

Many commercial CFB gasifiers are currently under operation in Europe (Vakkilainen et al. 2013). The main CFB biomass gasifiers are presented later in this chapter. However, the feasibility of DFB gasifiers is still contradictory. According to Corella et al. (2007), several DFB plants commissioned before mid-90s were dismantled for unknown reasons. On the other hand, successful results of an 8 MW_{fuel} demonstration plant in Güssing (Austria) have been recently reported by Hofbauer and co-workers (2002). Unlike CFB gasification, large-scale plants utilizing the DFB concept are still not available. The largest plant is located in Herten (Germany), with 15 MW_{fuel} input (Göransson et al. 2011).

3.6 Product Gas Contaminants

Biomass fuels are complex mixtures, mainly composed of organic C and O, but also H, N, and possibly S in very small quantities (compared to e.g. coal). The organic fraction consists of cellulose, hemicellulose and lignin. Inorganic materials such as alkali and alkaline earth metals (AAEM), silica and chlorine can also be found (Turn et al. 1998). The amount of each component depends on several factors, such as species, environment, harvesting conditions, handling, and the part of the plant from which the fuel is made. This variable composition, combined with operating conditions, contributes to formation of certain contaminants, such as tars (organic), inorganic oxides (ash), ammonia, and other undesirable substances. This section introduces possible contaminants, their impacts and primary measures that can be taken to reduce or even avoid their formation.

3.6.1 Tars

Tar formation is one of the main issues in biomass gasification (Devi et al. 2003). The term ‘tar’ has different definitions, depending on the application or research group. According to Devi et al. (2003, p.126), tar can be defined as “a complex mixture of condensable hydrocarbons, which includes single to 5-ring aromatic compounds along with other oxygen-containing hydrocarbons and complex polycyclic aromatic hydrocarbons” (PAH). Many relevant institutions such as the International Energy Agency (IEA), the U.S. Department of Energy (DOE) and the European Commission define tar as any gasification byproduct with a molecular weight higher than benzene (Basu 2010, Devi et al. 2003). Compounds present in tars depend on the type of biomass and gasification process that is being utilized (Abu El-Rub et al. 2004). Due to the diversity of compounds present in tars, they are usually divided into five groups (Li et al. 2009), as shown in Table 9.

The main problem caused by the presence of tars is plugging or clogging of pipes and ducts due to condensation when the product gas is cooled down. Other issues involve aerosol formation and possibility of polymerization (Basu 2010, Li et al. 2009). Tar deposition also reduces the effectiveness of heat transfer surfaces (Li et al. 2009). Additionally, tars may contain toxic substances and decrease the overall gas yield since they contain considerable amounts of energy (Abu El-Rub et al. 2004).

Table 9: Composition of tars. (Source: Li & Suzuki, 2009)

Class	Class name	Property	Representative compounds
1	GC-undetectable	Very heavy tars, cannot be detected by GC	Total gravimetric tar minus GC-detectable fraction
2	Heterocyclic aromatics	Contain hetero atoms; highly water soluble	Pyridine, phenol, cresols, quinoline, isoquinoline, dibenzophenol
3	Light aromatic (1 ring)	Light hydrocarbons with single ring; usually not condensable	Toluene, ethylbenzene, xylenes, styrene
4	Light PAH compounds (2–3 rings)	2 and 3 ring compounds; condense at low temperature even at very low concentration	Indene, naphthalene, methylnaphthalene, biphenyl, acenaphthalene, fluorene, phenanthrene, anthracene
5	Heavy PAH compounds (4–7 rings)	Larger than 3-ring; condense at high-temperatures even at low concentrations	Fluoranthene, pyrene, chrysene, perylene, coronene

In order to reduce tar content in the product gas, primary and secondary measures can be applied. Primary tar removal measures basically consist of changes in operation parameters.

Secondary measures are comprised of utilization of cleaning operations downstream. Primary methods are often preferable since the costs are usually lower and no further unit operations are required (Devi et al. 2003).

In their review, Devi et al. (2003) point out that the main primary methods for tar reduction involve appropriate operation conditions, the use of bed additives and/or adequate gasifier design. The most important operating parameters affecting tar formation are temperature, pressure, gasifying medium nature and amount, and residence time. Regarding temperature, operation above 800 ºC is preferred in order to obtain high char conversion and low tar content. However, increasing temperature might result in a product gas with lower heating value due to increase in H_2 content. Also, the risk of sintering and ash melting increases. Figure 18 summarizes the temperature effect in the gasification process. (Devi et al. 2003)

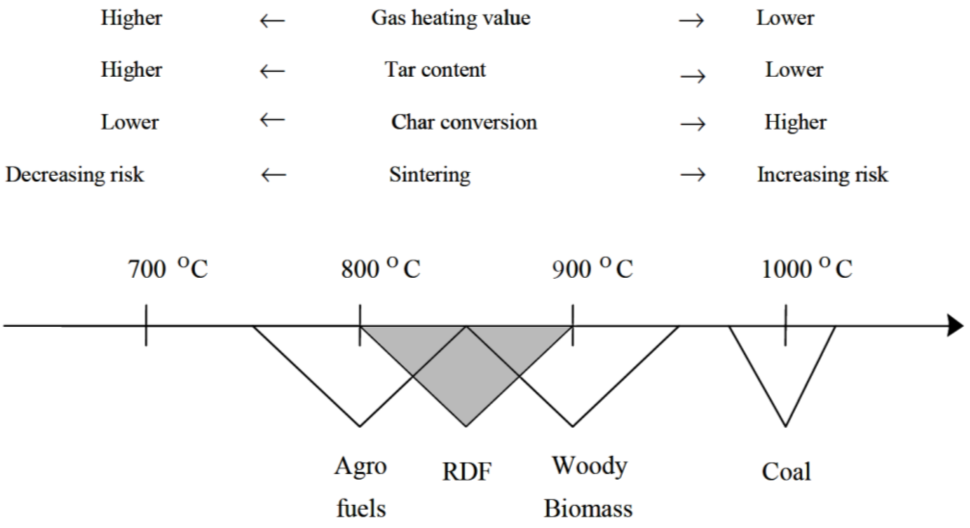

Figure 18: Temperature effect in gasification process. (Source: Devi et al., 2003)

Operating pressure has less influence in tar formation although it has been observed that pressure might reduce phenol formation. However, higher pressures might cause increase in PAH content (Knight 2000, cited in Devi et al. 2003, p. 128). On the other hand, the ER plays an important role. It is obvious that ER increase favors oxidation reactions and therefore tar reduction. However, the product gas' heating value also decreases due to CO_2 formation and/or N_2 dilution. In addition, it has been reported that increase in ER might cause PAH formation. In the case of other gasifying media (e.g. steam or steam/O_2), SB or GR increase usually decreases the tar content; however, LHV might be compromised. (Devi et al. 2003)

According to Devi, et al. (2003) the main bed additives currently utilized for tar reduction are Ni-based catalysts, olivine, and calcined dolomite or magnesite. Limestone and iron catalysts are also being investigated. The advantages of such additives are: promotion of char gasification, decrease in tar content and consequent changes in gas composition. Decrease in agglomeration risk is also observed when catalysts are utilized.

Dolomite is currently the most utilized bed additive in biomass gasification since it is effective, inexpensive and easily available (Basu 2010). Amounts from 3% to 10% of dolomite in bed material have proven to be effective for tar reduction and improvement of gas quality (Devi et al. 2003). However, dolomite is a soft material, and considerable amounts of fines are carried during the gasification. Therefore, studies have been conducted regarding the use of olivine, which has similar properties but is more resistant to abrasion.

Table 10: Tar reducing catalysts. (Adapted from: Abu El-Rub et al., 2004)

Catalyst	Advantages	Disadvantages
Dolomite	Inexpensive and abundant; attain high tar conversion (up to 95%); often used as guard beds for expensive catalysts; most popular for tar elimination	Fragile and quickly eroded from FB
Olivine	Inexpensive and high attrition resistance	Lower catalytic activity than dolomite
Iron	Inexpensive and abundant	Rapidly deactivated in absence of hydrogen; less active than dolomite
Char	Inexpensive; natural production inside the gasifier; high tar conversion (comparable to dolomite)	Consumed by gasification reactions
Ni-based	Able to attain complete tar elimination at ~ 900 °C; 8-10 times more active than dolomite	Rapid deactivation because of sulfur and high tar content; relatively more expensive

Iron in different forms has been reported as a catalyst for gasification, pyrolysis and tar cracking. In their review, Abu El-Rub et al. (2004) point out that metallic iron has proven to be more effective than iron oxides. However, hydrogen should be present in the medium in order to avoid coke formation and deposition. In addition, hematite ores have shown better results than magnetite ones for coal gasification (Cypers et al. 1980 cited in Abu El-Rub et al. 2004, p. 6 914). Table 10 summarizes possible tar reducing catalysts and their main features.

The main reactor design characteristics that promote tar reduction and improve gas yield are: secondary air injection, which also contributes to higher temperature; utilization of two-stage gasification, e.g. separation of pyrolysis and reduction zones; and addition of char during the gasifier start-up (Devi et al. 2003). Fast internal fluidized bed (FICFB) gasifiers, in which gasification and combustion zones are separate, were reported to provide a high LHV gas with low tar content (Fercher et al. 1998). Nevertheless, it seems that a compromise between tar content and gas LHV must usually be made.

3.6.2 Alkali/Alkaline Earth Metals and Other Inorganic

Besides Alkali and Alkaline Earth Metals (AAEM - mainly Na, K, Ca and Mg), inorganic substances such as Al, Cl, Fe, P, Si and Ti are commonly present in biomass. During gasification, these elements can form ash, attach to the bed material or be carried by the gaseous stream, contaminating the product gas. The distribution of inorganics in the gasification system depends on their content in the fuel, gasifier model, bed material type, and possibly on operation time. (Turn et al. 1998)

Alkali compounds usually present in biomasses are Na and K, the latter in higher concentration, up to 5%-w of dry matter (Turn et al. 1998). According to Turn et al. (1998), most alkali after gasification is found either in the ash or associated with the bed material, and only a small fraction (approximately 6%-w of the Na input) is actually released with the gases. Experiments conducted in a bench-scale BFB utilizing alumina silicate as bed material and bagasse as fuel confirmed that the main concern regarding alkali contaminants is the possibility of bed agglomeration due to low melting point ash. Ca, Mg, Al, Fe, P, Si and Ti mostly were found in the solid phase as well, representing no concern to downstream processes.

Studies regarding alfalfa gasification in pressurized FB have shown that only a small fraction of alkali and trace metals present in the fuel are carried by the gaseous stream (Salo et al. 1998). The fuel’s Na and K content was 280 and 24,600 ppm (wt) respectively, whereas the product gas contained only 0.04 ppm (wt) of total alkali content (Na + K) after being cooled to 432ºC and filtered. Additionally, it was demonstrated that the amount of trace metals (Hg, Cd, Pb, Se, Sb, Co, Be and As) in the product gas after high temperature/high pressure (HTHP) ceramic filter was below European Union regulations.

In order to reduce agglomeration risk, periodic substitution of bed material is the most important measure (Basu 2010). Addition of limestone or dolomite also avoids agglomeration once the presence of calcium produces a eutectic, higher melting point ash (Turn et al. 1998). Dolomite is recommended for in-bed tar removal. Biomass pretreatment has been considered as a possible solution for AAEM reduction in gasification systems as well (Turn et al. 1998, Lv et al. 2010). However, it has been shown that AAEM increase char reactivity, and their removal would result in a higher gasification initial temperature (Lv et al. 2010, Yip et al. 2009). Studies conducted by Yip et al. (2009) have shown that K, Na and Ca are (in this order) the most active catalysts present in bio-char when steam gasification is performed. Particularly, it was observed that the presence of AAEM has a positive effect in the water-gas shift reaction rate.

Another relevant inorganic substance formed during gasification is ammonia. The same study conducted by Turn et al. (1998) has shown that the greater the amount of nitrogen present in the fuel, the higher is the NH_3 concentration on a dry product gas. Chlorine is also a concern since most of it is released with the gas in the form of HCl. Temperatures should therefore be kept above the product gas dew point in order to avoid corrosion.

3.7 Gas Cleaning

Depending on the product gas application and environmental regulations, gas cleaning must be performed downstream of the gasifier. In case product gas is directly combusted, as occurs in the indurating machine, tars are unlikely to cause any problems to the system as long as they are maintained above the dew point (Milne et al. 1998), which is usually between 150 and 350 ºC (Cummer et al. 2002). The only issue in this case concerns air emissions. However, other contaminants such as AAEM and chlorine may cause problems in the induration furnace. No literature reference was found regarding this matter and further studies concerning the fate of alkali and chlorine are needed. In addition, particulate matter is usually present in

CFB gasification product gas, and its removal may be necessary. On the other hand, methanation catalysts might be deactivated due to tar/soot deposition, and previous gas cleaning is mandatory.

Although other methods are available, hot gas cleaning is usually preferred since tar condensation is thereby avoided. This operation is carried out at approximately 550 ºC; thus, the product gas must be first cooled (Olofsson et al. 2005). After sulfur removal, HTHP ceramic filters are utilized at pressures between 15 and 25 bars. The inorganic fraction is removed downstream, usually by adsorption (Cummer et al. 2002). Nonetheless, efficient and economical product gas cleaning is still an issue.

Studies conducted at the Energy Research Centre of the Netherlands (ECN) developed a promising technology for product gas cleaning (Zwart et al. 2006). In this concept, the product gas is cooled to approximately 400 ºC in a fire tube cooler before dust removal. In this manner, particles in the gas continuously erode the heat exchanger surface, preventing tar deposition. After cooling, particulate matter is separated by a cyclone or hot-gas filter. Tars are removed by an oil-based gas washer (OLGA). This system is comprised of two absorption columns, in which the scrubbing liquid is oil-based. Regeneration is done by air stripping, which is further recycled to the gasifier. Afterwards, chloride and sulfur components are removed by commercial adsorbents. One advantage of this system is that no water is condensed, avoiding contamination and need of waste water treatment. This system has been successfully implemented in a 4 MW_{th} demonstration plant in France (Zwart et al. 2009).

3.8 Methanation and Gas Upgrading

The product gas mainly consists of H_2 and CO, which leads to a relatively low energy content compared to NG. Methanation process can be used to convert carbon oxides (mainly CO) into methane. The most important reactions are water-gas shift and methanation, shown previously by equations R9 to R12 in Table 6. It is relevant to note that reactions R10 and R12 are a linear combination of R09 and R11 (Kopyscinski et al. 2010).

$$CO + H_2O \leftrightarrow CO_2 + H_2 \ \Delta_r H^0 = -41\ kJ/mol \tag{R09}$$

$$2CO + 2H_2 \leftrightarrow CH_4 + CO_2 \ \Delta_r H^0 = -247\ kJ/mol \tag{R10}$$

$$CO + 3H_2 \leftrightarrow CH_4 + H_2O \; \Delta_r H^0 = -206\ kJ/mol \qquad \text{(R11)}$$

$$CO_2 + 4H_2 \leftrightarrow CH_4 + 2H_2O \; \Delta_r H^0 = -165\ kJ/mol \qquad \text{(R12)}$$

As can be seen from reaction R11, the minimum H_2/CO molar ratio is 3:1. Nonetheless, the achievable ratio is approximately 1:1 for DFB steam gasification (Zwart et al. 2006). Gas conditioning might be necessary in order to adjust the relative amount of hydrogen and oxygen. The water-gas shift reaction (R9 in Table 6) can be carried out in fixed bed reactors utilizing commercial Ni-based catalysts before methanation. According to Zwart et al. (2006), a steam content of 33%-mol (wet basis) and temperatures about 370 ºC are required in order to achieve a H_2/CO ratio of 3:1. On the other hand, equilibrium calculations have indicated that higher temperatures (e.g. 900-1,000 K) and slightly high pressure are advised in order to increase H_2 content. However, CH_4 content decreases to practically zero under high temperature conditions, decreasing process efficiency (Haryanto et al. 2009).

Methanation reactions are exothermic, and therefore low temperatures are desired. In practice, temperatures between 300 and 450 ºC are achievable with cooling and/or recirculation (Higman et al. 2008). Catalysts are needed at lower temperatures, and the most utilized are NiO-based. In addition, high pressures (approximately 30 bars) also contribute to methane formation (Gao et al. 2012).

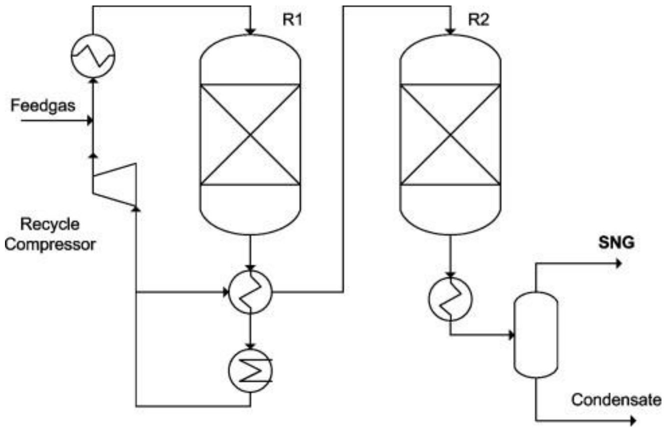

Figure 19: Lurgi methanation process. (Source: Kopyscinski et al., 2010)

Fixed and FB reactors can be used for methanation on a commercial scale. This topic has been recently revised by Kopyscinski et al. (2010). Fixed bed methanation reactors are currently the most utilized in ammonia plants in order to remove CO from H_2-rich streams (Kopyscinski et al. 2010). Lurgi’s methanation process (Figure 19) consists of two fixed bed adiabatic reactors with internal recycle. Temperature range is between 250 and 400 ºC. The first (and only) commercial coal-SNG plant in the USA utilizes this system.

The TREMP^{TM} (Topsøe’s Recycle Energy Efficient Methanation) technology is comprised of three adiabatic fixed bed reactors with recirculation and intermediate cooling (Kopyscinski et al. 2010). This process is shown in Figure 20. Methanation is performed at temperatures between 250 and 700 ºC and pressures up to 30 bars. High pressure superheated steam is a by-product of this process.

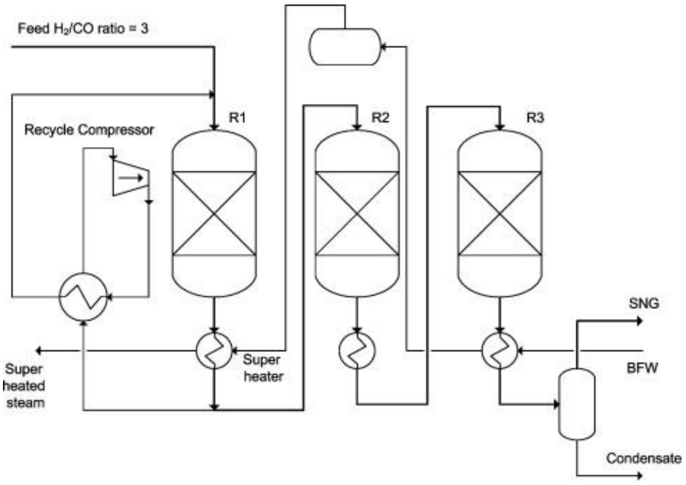

Figure 20: TREMP^{TM} methanation process. (Source: Kopyscinski et al., 2010)

Other fixed bed methanation processes were also developed on commercial/semi-commercial scales. For instance, HICOM process is able to perform gas conditioning and methanation in the same unit, consisting of several adiabatic reactors (Kopyscinski et al. 2010). Linde’s methanation process utilizes one isothermal indirect heated fixed bed and one adiabatic reactor, but this concept is currently utilized for methanol synthesis. The RMP process does not utilize gas recycling, comprising 4-6 reactors with intermediate gas cooling. Although it has not reached commercial scale, ECN has proposed the utilization of a similar concept recently (Zwart et al. 2006).

FB methanation has many advantages, for instance: improved mass and heat transfer, isothermal operation, and easy catalyst handling. Although several projects have been developed in the 60-70s, only the Thyssengas Comflux process (Figure 21) still seems promising for large-scale operation (Kopyscinski et al. 2010). In this concept, shift and methanation reactions can be done in a single stage in only one FB unit. Temperatures between 300 and 500 ºC and pressure from 20 to 60 bars are utilized. Another advantage is higher tolerance of the catalyst to short-chain hydrocarbons, such as ethylene. FICFB research in Güssing (Austria) is currently utilizing this process to produce 1 MW of bio-SNG.

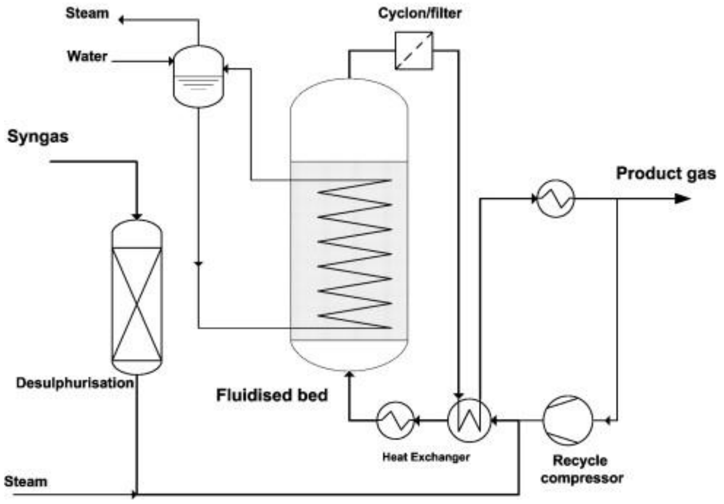

Figure 21: Comflux methanation process. (Source: Kopyscinski et al., 2010)

Gas upgrading may be necessary in order to further remove H_2O and unreacted CO_2. Gas upgrading can be performed before or after methanation (Kopyscinski et al. 2010). Methane upgrading is already a well-developed technology since it is widely utilized in NG production. Chemical, physical and hybrid absorption are the most utilized systems for CO_2 removal. The former typically utilizes amine or carbonate based solvents, such as monoethanolamine (MEA), diethanolamine (DEA), ammonia or hot potassium carbonate (Olofsson et al. 2005). These reactions are irreversible and therefore a large amount of waste is generated. On the other hand, physical absorption allows solvent regeneration, by either heating and/or depressurizing the system. Typical solvents are Selexol (dimethylether of polyethylene glycol) and Rectisol (cold methanol). High pressures are required in physical absorption systems. Solid adsorption is also possible in materials such as alumina, zeolite or activated carbon, but the costs are usually higher. Membrane separation is a promising technology for SNG upgrading, but it has not reached commercial scale yet.

Although biomass conversion into bio-SGN is not yet commercially utilized, Gao et al. (2012) estimate that at least 15 coal-to-SNG projects are being developed in the US and more than 20 are proposed or under construction in China.

3.9 State-of-the-art of Biomass Gasification

Many commercial biomass gasification plants are currently under operation, mainly in Northern Europe. The ten largest gasifiers in 2012 are outlined in Table 11, as well as their location, purpose and fuel.

Table 11: Ten largest biomass gasification plants by 2012. (Data source: Vakkilainen et al., 2013)

No	Company	Purpose	Location	Fuel	MW_{fuel}
1	Lahti Energia	CHP	Lahti, Finland	SRF	160
2	Vaskiluodon Voima	CHP	Vaasa, Finland	Local biomass	140
3	Rüdersdorfer Zement	Calciner	Rüdersdorf, Germany	Biomass waste	100
4	Essent	CHP	Geertruidenberg, Netherlands	Waste wood	85
5	Electrabel	Boiler	Ruien, Belgium	Wood chips	50
6	Metsä Fibre	Lime kiln	Joutseno, Finland	Bark	48
7	Södra Cell Värö	Lime kiln	Väro, Sweden	Bark	35
8	Agnion Technologies	CHP	Pfaffenhofen, Germany	Waste wood	33
9	Corenso United	CHP	Varkaus, Finland	Packaging	32
10	Skive Fjernvarme	CHP	Skive, Denmark	Wood pellets	32

The world’s largest gasifier is actually the number 3 on the list since the two first ones have two identical units in parallel. Figure 22 and Figure 23 show schemes of the two first plants on the list, which are both connected to steam boilers for combined heat and power (CHP) production.

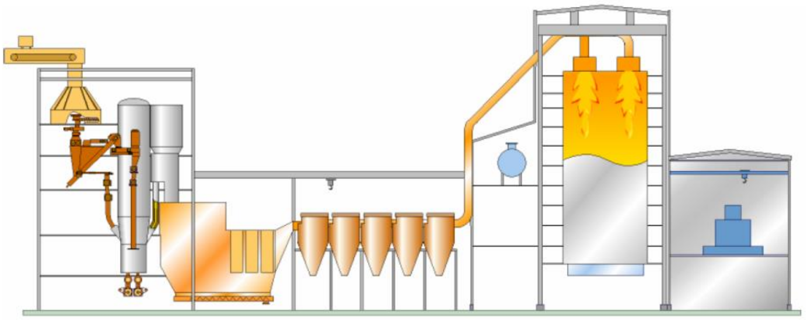

Figure 22: Kymijärvi II gasification process in Lahti, Finland. (Source: Palonen et al., 2005)

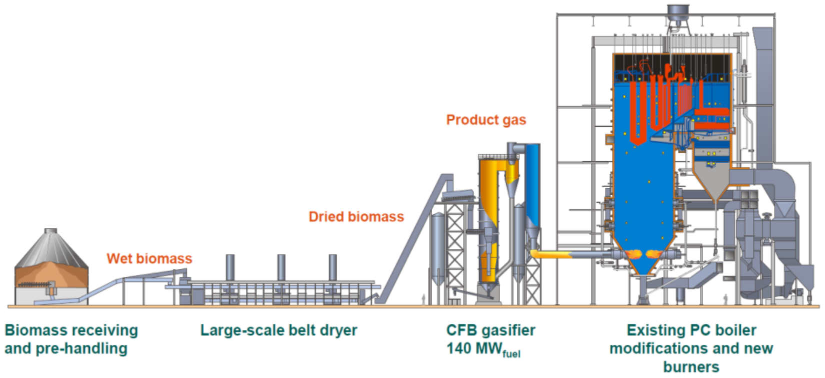

Figure 23: Vaskiluodon Voima biomass gasification plant. (Adapted from: Breitholtz, 2011)

4 CALCULATION METHODOLOGY

In this study, different gasification technologies are compared, focusing on their applications in the iron ore pelletizing process. Mass and energy balances for the current case and several alternative fuels are calculated. Adiabatic flame temperature and gas flows in preheating and firing zones were considered as key technical parameters for this analysis. In this chapter, the calculation methodology is described and the main model assumptions are explained.

4.1 Gas Balance

The gas balance is used to estimate average gas flows through the system, based on design and plant data, under normal operating conditions. The third indurating machine's gas balance was calculated based on compilation of operating data. In situations which conflicting data were found from different sources, those values shown by the gas flow schema were chosen. The main assumptions made in this calculation were:

- Firing and preheating zones were considered as a single control volume in which all NG and combustion air are injected and all fuel (solid and gas) is burnt;
- Combustion of NG and solid fuel is considered to occur uniformly inside the control volume, according to the following chemical reactions:

$$CH_4 + 2O_2 \rightarrow CO_2 + 2H_2O \quad (4)$$

$$C_2H_6 + \frac{7}{2}O_2 \rightarrow 2CO_2 + 3H_2O \quad (5)$$

$$C_3H_8 + 5O_2 \rightarrow 3CO_2 + 4H_2O \quad (6)$$

$$C_{(s)} + O_2 \rightarrow CO_2 \quad (7)$$

$$2H_2 + O_2 \rightarrow 2H_2O \quad (8)$$

$$S + O_2 \rightarrow SO_2 \quad (9)$$

- All gases involved were considered to be ideal, therefore following the equation of state below. This assumption allows streams measured in normal cubic meters (Nm^3) to be simply summed or subtracted. Nonetheless, it is known that after combustion the total number of moles in the gas phase changes due to the reactions represented by Equations 5, 6 and 8. Therefore, the gas flow in Nm^3 in the kiln hood (above the grate) is not necessarily equal to the flow underneath the grate (in the wind boxes);

$$pV = nRT \tag{10}$$

Where: p = pressure [Pa]
V = volume [m^3]
n = number of moles [mol]
R = ideal gas constant [8.314 J mol^{-1} K^{-1}]
T = absolute temperature [K]

- Differences between gas flow values which were lower than 5% were considered negligible due to aforementioned approximations and unknown measurement accuracy.

4.2 Biomass Properties

In order to calculate the adiabatic flame temperature and gas flow rates for all gasification technologies it is first necessary to estimate the product gas composition in each case. This estimation can be done by mass balances, starting with eucalyptus wood ultimate analysis shown in Table 12. The MC corresponds to the specification after drying. As ash does not take part in any reactions, utilizing the wood composition on a DAF basis is a useful simplification.

Table 12: Eucalyptus wood ultimate analysis.

Element	**%-w (dry)**	**%-w (DAF)**	**%-w (wet)**	**Wood kmol/kg$_{wood}$**	**%-mol (wet)**
C	50.1%	50.5%	43.9%	0.0366	30.6%
H	6.00%	6.05%	5.26%	0.0522	43.6%
O	43.0%	43.3%	37.7%	0.0236	19.7%
N	0.10%	0.10%	0.09%	0.0001	0.05%
H_2O	15.0%	15.0%	13.0%	0.0072	6.05%
Ash	0.80%	-	-	-	-
TOTAL	**115%**	**115%**	**100%**	**0.1196**	**100%**

Wood's high heating value (HHV), in MJ kg^{-1}, was calculated based on a unified correlation represented by Equation 11 (Channiwala et al. 2002). LHV on a dry basis and wet basis (15% MC) were calculated according to Equations 12 and 13, accordingly.

$$HHV_{dry} = 0.3491 \cdot \%C + 1.1783 \cdot \%H + 0.1005 \cdot \%S - 0.1034 \cdot \%O - 0.0151 \cdot \%N - 0.0211 \cdot \%A \quad (11)$$

Where: *%C* = carbon content, dry basis [%-w]
%H = hydrogen content, dry basis [%-w]
%S = sulfur content, dry basis [%-w]
%O = oxygen content, dry basis [%-w]
%N = nitrogen content, dry basis [%-w]
%A = ash content, dry basis [%-w]

$$LHV_{dry} = HHV_{dry} - 2.441 \cdot \left(\frac{MM_{H2O}}{MM_{H2}}\right) \cdot \frac{\%H}{100} \quad (12)$$

$$LHV_{wet} = LHV_{dry} \cdot (1 - w_{H2O}) - 2.441 \cdot w_{H2O} \quad (13)$$

Where: *w* = mass fraction, dry basis [-]

4.3 Fuel Composition and Properties Estimation

Several types of fuels were evaluated in this study, such as NG, producer gases from different gasification processes, bio-SNG, coal, petcoke, and a gas/solid blend, which was considered as the final fuel fed into the traveling grate. Each fuel composition calculation and properties estimation is discussed separately.

4.3.1 NG

Average dry NG in Brazil was considered to contain methane, ethane, propane, carbon dioxide and nitrogen (COMPAGAS n.d.). Other hydrocarbons might also be present (e.g. butane and pentane), but their amount was considered negligible. NG composition, as well as each component's chemical properties is shown in Table 13.

Table 13: NG composition (COMPAGAS n.d.) and component's properties.

Component	**%-mol (wet)**	**MM kg/kmol$_{gas}$**	**Mass kg/kmol$_{gas}$**	**Molar V Nm^3/kmol$_{gas}$**	**%-vol (wet)**
CH_4	89.1%	14.29	79.1%	19.94	89.2%
C_2H_6	5.90%	1.774	9.81%	1.308	5.85%
C_3H_8	2.90%	1.279	7.07%	0.6362	2.85%
CO_2	0.90%	0.396	2.19%	0.2004	0.90%
N_2	1.20%	0.336	1.86%	0.2688	1.20%
H_2O	0.00%	0.000	0.00%	0.000	0.00%
TOTAL	**100%**	**18.08**	**100%**	**22.36**	**100%**

NG's LHV is estimated based on the standard enthalpy of combustion of each combustible component at 298.15 K and 1 atm. The net energy content in the gaseous mixtures is therefore defined by:

$$LHV = \sum y_i \cdot \Delta_c H_i^0 \tag{14}$$

Where: y_i = molar fraction of component *i*, wet basis [-]
$\Delta_c H^0$ = standard enthalpy of combustion [J mol^{-1}]

Other properties, such as MM [kg kmol^{-1}] and molar volume [Nm^3 kmol^{-1}], can also be determined similarly to Equation 14, substituting the LHV by the property of interest:

$$MM = \sum y_i \cdot MM_i \tag{15}$$

$$\bar{V} = \sum y_i \cdot \bar{V}_i \tag{16}$$

Where: $\bar{V}$= molar volume [Nm^3 kmol^{-1}]

4.3.2 Solid Blend

Coal and petcoke are the main solid fuels added to the green pellets. According to the third indurating machine's average consumption, approximately 33%-w corresponds to the former, while the remaining 67%-w to the latter. LHV and MC remain as shown in Table 2, resulting in an energy content of 33.1 MJ kg^{-1} (297 MJ kmol^{-1}) for this solid blend. Typical ultimate analysis for coal and petcoke (Fermoso et al. 2009) as well as the final solid fuel composition is shown in Table 14.

Table 14: Solid fuel composition.

Element	**Coal** **%-w$_{DAF}$**	**Petcoke** **%-w$_{DAF}$**	**Solid** **%-w$_{DAF}$**	**Solid** **%-w$_{wet}$**	**Solid** **kmol/kg$_{solid}$**	**Solid** **%-mol**
C	71.2%	87.6%	82.2%	74.7%	0.0622	56.0%
H	4.80%	3.80%	4.13%	3.75%	0.0372	33.5%
O	20.8%	0.90%	7.47%	6.79%	0.0042	3.82%
N	1.50%	1.50%	1.50%	1.36%	0.0010	0.88%
S	1.70%	6.20%	4.72%	4.29%	0.0013	1.20%
H_2O	10.0%	10.0%	10.0%	9.09%	0.0050	4.54%
TOTAL	**110%**	**110%**	**110%**	**100%**	**0.1111**	**100%**

4.3.3 Air/Oxygen Gasification Product Gas Composition

In air or oxygen gasification, the biomass is partially combusted in order to provide sufficient heat for endothermic gasification reactions. Therefore, the amount of O_2 fed into the gasifier must be estimated. The ER, defined by Equation 1, is the ratio between actual and theoretical amount of O_2 necessary for complete combustion. Based on the eucalyptus elemental composition shown in Table 12, the theoretical oxygen required for complete combustion is determined considering that only carbon and hydrogen are oxidized. The oxygen present in the wood accounts negatively. Thus, the stoichiometric amount of oxygen per kilogram of wood is given by:

$$n_{O2}^{stoich} = n_C^{wood} + \frac{n_H^{wood}}{4} - \frac{n_O^{wood}}{2} \tag{17}$$

For air gasification, the stoichiometric air must be determined based on predefined properties. Ambient air entering the gasifier was considered to be at 25 °C, sea level pressure (101,325 Pa) and 60% of relative humidity (RH). Its MC can be calculated based on the RH:

$$RH = \frac{p_{H2O}}{p_{sat}} \tag{18}$$

Where: p_{H2O} = water partial pressure in atmosphere [Pa]

p_{sat} = saturation pressure of water at 25 °C [Pa]

The water saturation pressure at 25 °C can be evaluated utilizing water/steam tables, and it is approximately 1,902 Pa. According to Raoult's Law, the water molar fraction in air equals the ratio between water partial pressure and atmospheric pressure (p_{atm}), as follows:

$$y_{H20}^{air} = \frac{p_{H2O}}{p_{atm}} \tag{19}$$

Assuming that dry air contains 21%-mol of oxygen and 79%-mol of nitrogen, O_2 and N_2 percentages on wet basis are estimated considering the number of water moles (n_{H2O}) present in 1 mol of wet air, as shown by Equations 20 to 22:

$$n_{H2}^{air} = \frac{y_{H2O}^{air}}{1 - y_{H2O}^{air}} \tag{20}$$

$$y_{O2}^{air} = \frac{0.21}{1 + n_{H20}^{air}} \tag{21}$$

$$y_{N2}^{air} = \frac{0.79}{1 + n_{H20}^{air}} \tag{22}$$

Air gasifier inlet composition considers the sum of all wood and wet air components. For instance, no carbon and hydrogen enter with air; therefore, these quantities remain the same (e.g. the carbon input in the gasifier equals the amount of carbon in the wood). However, oxygen, nitrogen and water coming from air must be added. For this purpose, wood and air compositions, as well as the ER are utilized, as follows:

$$n_{O}^{input} = n_{O}^{wood} + 2 \cdot ER \cdot n_{O2}^{stoich} \tag{23}$$

$$n_{N}^{input} = n_{N}^{wood} + 2 \cdot ER \cdot \left(\frac{y_{N2}^{air}}{y_{O2}^{air}}\right) \cdot n_{O2}^{stoich} \tag{24}$$

$$n_{H2O}^{input} = n_{H2O}^{wood} + ER \cdot \left(\frac{y_{H2O}^{air}}{y_{O2}^{air}}\right) \cdot n_{O2}^{stoich} \tag{25}$$

Similar equations were adopted for oxygen gasification. Nevertheless, a mixture of 95% O_2 and 5% N_2 was considered as the gasifying media. In addition, some steam might be added in the riser in order to control the temperature. The amount of steam is represented by the steam oxygen ratio (SOR), which is defined as the mass of steam added per kilogram of oxygen (van der Meijden et al. 2010). In this case, the gasifier input is calculated as follows:

$$n_{O}^{input} = n_{O}^{wood} + 2 \cdot ER \cdot n_{O2}^{stoich} \tag{26}$$

$$n_{N}^{input} = n_{N}^{wood} + 2 \cdot ER \cdot \left(\frac{y_{N2}^{oxygen}}{y_{O2}^{oxygen}}\right) \cdot n_{O2}^{stoich} \tag{27}$$

$$n_{H2O}^{input} = n_{H2O}^{wood} + ER \cdot SOR \cdot \left(\frac{MM_{H2O}}{MM_{O2}}\right) \cdot n_{O2}^{stoich} \tag{28}$$

The product gas was considered to be a mixture of CH_4, CO_2, CO, H_2, N_2 and H_2O for all gasification technologies. Short-chain hydrocarbons, such as ethane, might also be formed; however, their amount was considered negligible. Mass balances relating the gasifier inlet, in terms of C, H, O, N and H_2O, to the product gas composition are presented below. Firstly, carbon balance must contemplate CH_4, CO_2 and CO amounts discounting any unconverted carbon withdrawn with the ashes:

$$\frac{CC}{100} \cdot n_{C}^{input} = n_{CH4}^{PG} + n_{CO2}^{PG} + n_{CO}^{PG} \tag{29}$$

Where: *CC* = carbon conversion [%]

Hydrogen balance takes into account that it is transformed into CH_4, H_2 and also H_2O. In addition, some water is also fed into the system, and this value should be subtracted, as follows:

$$n_{H}^{input} = 4 \cdot n_{CH4}^{PG} + 2 \cdot n_{H2}^{PG} + 2 \cdot (n_{H2O}^{PG} - n_{H2O}^{input}) \tag{30}$$

Oxygen fed into the system completely reacts with C and H, forming CO_2, CO and H_2O. As occurred in Equation 30, the water input must also be subtracted. Finally, nitrogen balance is the simplest one since all N was considered to form N_2:

$$n_O^{input} = 2 \cdot n_{CO2}^{PG} + n_{CO}^{PG} + (n_{H2O}^{PG} - n_{H2O}^{input}) \tag{31}$$

$$n_N^{input} = 2 \cdot n_{N2}^{PG} \tag{32}$$

The amount of water in the product gas equals to the water input, plus water formed by hydrogen oxidation minus water consumed by gasification reactions (refer to reactions R2, R9 and R13 from Table 6). The water balance is based on oxygen and hydrogen balances; for instance, the unreacted oxygen (after carbon reactions) must combine with hydrogen to form water, as no oxygen shall remain in the product gas. At this point, it should be noted that the water balance is a linear combination of oxygen and hydrogen balances. As a result, there are four independent mass balances (Equations 29 to 32) and six unknown variables, corresponding to the amount of each component in the product gas. In order to decrease the problem's degree of freedom and find a solution, it is necessary to introduce two other restricting equalities.

Firstly, it is known that the CO/CO_2 ratio is a function of temperature. For example, a frequently used expression gives this ratio as a function of char surface temperature (Arthur 1951, cited in Basu 2010, p. 126):

$$\frac{[CO]}{[CO_2]} = 2{,}400e^{-\left(\frac{6{,}234}{\theta}\right)} \tag{33}$$

Where: θ = temperature [°C]

Methane molar fraction in the product gas (dry basis) was also specified. According to several studies, methane formation is difficult to predict utilizing simple gasification models, such as those based on equilibrium constants or Gibbs free energy (Baratieri et al. 2008, Puig-Arnavat et al. 2010). Typical methane concentrations within the ranges presented in Table 7 were tested, and the chosen value corresponds to that one resulting in better agreement with the other compounds' concentrations. For example, the acceptable range for methane concentration in air gasification product gas is between 2 and 6%-mol on a dry basis. Several values were tested within this range, and finally the CH_4 concentration was set as 4.70%, which makes the other component's concentrations to be as close as possible to the typical values when the mass balance is closed. Since the methane molar fraction is given on a dry basis, the total number of moles is unknown; thus, an iterative procedure is necessary in order to determine wet basis molar fraction.

Once all required parameters were specified, the system of equations can be solved. CO_2 and CO amounts were calculated by solving Equations 29 and 33 simultaneously, in which the char surface temperature was considered to be equal to the gasifier average temperature. The results were then used in Equation 31, from which the amount of water in the product gas was determined. Finally, H_2 and N_2 amounts are found through Equations 30 and 32, respectively. A new CH_4 number of moles must be introduced if its molar fraction on a dry basis is different from the value established. The calculation procedure starts over, until conversion is reached. Excel goal seek tool was utilized in this case. LHV of product gas was estimated by Equation 14, considering CH_4, CO and H_2 as combustibles. MM and molar volume were also determined by similar equations.

4.3.4 Steam Gasification Product Gas Composition

In steam gasifiers, fuel biomass and steam are fed into the system and heat is provided indirectly; for example, by circulating bed material from a combustion chamber. Therefore, the input composition is the sum of elements present in wood and water, which is given by the SB ratio (Equation 2). The SB ratio is often expressed in terms of total mass of water (including biomass MC) per kilogram of wood on a DAF basis; the water input can be found by converting the SB into kmol per kilogram of wet wood according to the following equation:

$$n_{H2O}^{input} = \frac{SB}{MM_{H2O} \cdot \left(1 - w_{H2O}^{wood}\right)} \tag{34}$$

At this point, CC has to be set. In this case, the value is much lower than for air/oxygen gasification, based on the premise that part of the char is burnt in the combustion zone in order to provide energy to gasification reactions. Contrary to air or oxygen gasification, the combustion products and nitrogen do not mix with the product gas. Thus, the burnt char must be subtracted from the mass balance in order to obtain an accurate product gas composition. Van der Meijden (2010) proposed a relation between CC and gasifier temperature for allothemal systems:

$$CC = 65 + 0.15 \cdot (\theta - 760) \tag{35}$$

Once the gasifier input and CC are defined, the product gas composition can be determined. Equations 29 to 33 and corresponding assumptions were also adopted. Methane content on a dry basis was fixed in 11%, and the iteration follows the same procedure. The product gas LHV was estimated by Equation 14, considering CH_4, CO and H_2 as combustibles. MM and molar volume were determined by Equations 15 and 16.

4.3.5 Bio-SNG Composition

In order to estimate the bio-SNG composition, a chemical equilibrium approach was utilized. Only water-gas shift (R9) and methanation reactions R11 and R12 were considered to occur. Equilibrium constants at a fixed temperature (T) and 101,325 Pa were calculated as follows (Moran et al. 2006):

$$\ln k = -\frac{\Delta_r G^0}{RT} \tag{36}$$

Where: k = equilibrium constant at T and 1 atm

$\Delta_r G^0$ = Gibbs free energy change of reaction [J mol^{-1} K^{-1}]

The standard Gibbs free energy change for each reaction can be calculated utilizing pure species enthalpies and entropies. As entropy and enthalpy are functions of state, their variation always equals the difference between products and reactants:

$$\Delta_r H^0 = \sum \nu_i (H_{f,i}^0 + \Delta h_i) \tag{37}$$

$$\Delta_r S^0 = \sum \nu_i S_i^0 \tag{38}$$

Where: ν_i = stoichiometric number [-]

H_f^0 = standard enthalpy of formation at 298.15 K [J mol^{-1}]

Δh = sensible heat of each specie at T [J mol^{-1}]

S^0 = standard entropy for each specie at T [J mol^{-1} K^{-1}]

$$\Delta_r G^0 = \Delta_r H^0 - T\Delta_r S^0 \tag{39}$$

The stoichiometric number for each species is always equal to its stoichiometric coefficient in a particular reaction, with a positive sign for products and negative sign for reactants. Pure gaseous component enthalpies at T were calculated utilizing the Shomate Equation, with 298.15 K and 1 atm pressure as reference T and p respectively (Chase Jr 1998):

$$\Delta h = H^0 - H_{298}^0 = At + B\frac{t^2}{2} + C\frac{t^3}{3} + D\frac{t^4}{4} - \frac{E}{t} + F - H \tag{40}$$

Where: t = T/1,000

Equation 40 parameters from A to H, for the compounds of interest, can be found in Appendix A. The standard enthalpy of formation is the value in column H, expressed in kJ mol^{-1}. Similarly to the enthalpies, S^0 can also be estimated by Shomate coefficients, through the equation:

$$S^\circ = A \cdot ln(t) + B \cdot t + C \cdot \frac{t^2}{2} + D \cdot \frac{t^3}{3} - \frac{E}{2t^2} + G \tag{41}$$

Once Gibbs free energy changes are determined, equilibrium constants for each reaction can be obtained. The equilibrium constant for a generic chemical reaction (e.g. Equation 42) at fixed pressure (p) and temperature (T) is defined by Equation 43 (Moran et al. 2006):

$$aA + bB \leftrightarrow cC + dD \tag{42}$$

$$K = \frac{n_C^c \cdot n_D^d}{n_A^a \cdot n_B^b} \cdot \left(\frac{\frac{p}{p_{ref}}}{n^{eq}}\right)^{\Sigma \nu_i} \tag{43}$$

Where: $\sum v_i = c + d - a - b$

$n^{eq} = n_A + n_B + n_C + n_D$

n^{eq} = number of moles in the equilibrium [kmol kmol$_{gas}^{-1}$]

Applying this definition to reactions R9, R11 and R12, Equations 45, 47 and 49 are obtained, respectively. However, three equations are written for five variables. Therefore, it is necessary to write the number of moles in the equilibrium in terms of the variables *x*, *y* and *z*, introduced in Equations 44, 46 and 48, respectively.

$$xCO + xH_2O \leftrightarrow xCO_2 + xH_2 \tag{44}$$

$$k_{R9} = \frac{n_{CO2} \cdot n_{H2O}}{n_{CO} \cdot n_{H2}} \tag{45}$$

$$yCO + 3yH_2 \leftrightarrow yCH_4 + yH_2O \tag{46}$$

$$k_{R11} = \frac{n_{CH4} \cdot n_{H2O}}{n_{CO} \cdot n_{H2}^3} \cdot \left(\frac{p}{n^{eq}}\right)^{-2} \tag{47}$$

$$zCO_2 + 4zH_2 \leftrightarrow zCH_4 + 2zH_2O \tag{48}$$

$$k_{R12} = \frac{n_{CH4} \cdot n_{H2O}^2}{n_{CO2} \cdot n_{H2}^4} \cdot \left(\frac{p}{n^{eq}}\right)^{-2} \tag{49}$$

The number of moles of each species when chemical equilibrium is reached can be written as a function of initial amounts in steam gasification product gas and the variables *x*, *y* and *z*. These relations are written below. As can be seen from Equation 54, water might be added to the reactor inlet in order to avoid soot formation.

$$n_{CH4}^{eq} = n_{CH4}^{PG} + y + z \tag{50}$$

$$n_{CO}^{eq} = n_{CO}^{PG} - x - y \tag{51}$$

$$n_{CO2}^{eq} = n_{CO2}^{PG} + x - z \tag{52}$$

$$n_{H2}^{eq} = n_{H2}^{PG} + x - 3y - 4z \tag{53}$$

$$n_{H2O}^{eq} = n_{H2O}^{PG} + n_{H2O}^{reactor} - x + y + 2z \tag{54}$$

$$n^{eq} = n^{PG} - 2y - 2z \tag{55}$$

Where: $n^{reactor}$= water added to the reactor [kmol $kmol_{gas}^{-1}$]

Substituting the expressions above in Equations 45, 47 and 49, a system of three equations and three variables can be solved. Once variables *x*, *y* and *z* are determined, the equilibrium composition in the methanation reactor outlet can be found. After methanation, the gas is cooled to 25 °C and water is removed. The final water molar fraction can be estimated based on the saturation pressure at 25 °C and Equation 19. Carbon dioxide is further removed by absorption. During this process, some methane may be lost. CH_4 and CO_2 amounts are expressed as:

$$n_{CH4}^{SNG} = \eta_{CH4} \cdot n_{CH4} \tag{56}$$

Where: η_{CH4} = methane removal efficiency [-]

$$n_{CO2}^{SNG} = (1 - \eta_{CO2}) \cdot n_{CO2} \tag{57}$$

Where: η_{CO2} = CO_2 removal efficiency [-]

As previously, bio-SNG's LHV was estimated by Equation 14, considering CH_4, CO and H_2 as the only combustible compounds. MM and molar volume were also determined by similar equations.

4.3.6 Combined Gas and Solid Composition

Gaseous and solid fuels are utilized in the pelletizing kiln. In order to simplify the calculation of stoichiometric oxygen required for complete combustion, the elemental composition (in terms of C, H, O, N, S and H_2O) of both gaseous and solid fuels were combined, and the final fuel (gas + solid) composition and its combustion properties were determined. The calculation basis was chosen to be 1 GJ of fuel. Thus, the elemental composition for NG is given by:

$$N_C^{NG} = \frac{(n_{CH4}^{NG} + 2 \cdot n_{C2H6}^{NG} + 3 \cdot n_{C3H8}^{NG} + n_{CO2}^{NG})}{LHV_{NG}\ [\text{GJ kmol}^{-1}]} \tag{58}$$

Where: N = number of moles per GJ of fuel [kmol GJ^{-1}]

$$N_H^{NG} = \frac{(4 \cdot n_{CH4}^{NG} + 6 \cdot n_{C2H6}^{NG} + 8 \cdot n_{C3H8}^{NG})}{LHV_{NG}\ [\text{GJ kmol}^{-1}]} \tag{59}$$

$$N_O^{NG} = \frac{2 \cdot n_{CO2}^{NG}}{LHV_{NG}\ [\text{GJ kmol}^{-1}]} \tag{60}$$

$$N_N^{NG} = \frac{2 \cdot n_{N2}^{NG}}{LHV_{NG}\ [\text{GJ kmol}^{-1}]} \tag{61}$$

$$N_{H2O}^{NG} = \frac{n_{H2O}^{NG}}{LHV_{NG}\ [\text{GJ kmol}^{-1}]} \tag{62}$$

On the other hand, product gas and bio-SNG are considered to contain only CH_4, CO, CO_2, H_2, N_2 and H_2O. Although Equations 61 and 62 remain the same, the elemental composition for C, H and O is determined by the following equations:

$$N_C^{PG} = \frac{(n_{CH4}^{PG} + n_{CO}^{PG} + n_{CO2}^{PG})}{LHV_{PG}\ [\text{GJ kmol}^{-1}]} \tag{63}$$

$$N_H^{PG} = \frac{(4 \cdot n_{CH4}^{PG} + 2 \cdot n_{H2}^{PG})}{LHV_{PG}\ [\text{GJ kmol}^{-1}]} \tag{64}$$

$$N_O^{PG} = \frac{(n_{CO}^{PG} + 2 \cdot n_{CO2}^{PG})}{LHV_{PG}\ [\text{GJ kmol}^{-1}]} \tag{65}$$

Similarly, the solid elemental composition, in kmol GJ^{-1}, can be determined by dividing the amount of each element (in kmol kg^{-1}) by its LHV in GJ kg^{-1}. The final fuel composition, as well as its LHV, is obtained by the weighted average:

$$N_i^{fuel} = \frac{\Phi_{gas}}{\Phi_{th}} \cdot N_i^{gas} + \frac{\Phi_{solid}}{\Phi_{th}} \cdot N_i^{solid} \tag{66}$$

$$LHV_{fuel} = \frac{\Phi_{gas}}{\Phi_{th}} \cdot LHV_{gas} + \frac{\Phi_{solid}}{\Phi_{th}} \cdot LHV_{solid} \tag{67}$$

Where: Φ_{gas} = energy provided by gaseous fuel [MW]

Φ_{solid} = energy provided by solid fuel [MW]

Φ_{th} = total thermal energy input [MW]

4.4 Combustion Air

Stoichiometric oxygen necessary for complete combustion is determined based on the final fuel composition, considering one GJ of fuel as a basis. Oxygen demand was estimated according to combustion reactions presented in Equations 4 to 9. Oxygen content in the fuel accounts negatively. If no nitrogen oxides are considered to be produced during the reaction, and all sulfur is oxidized to SO_2, the theoretical oxygen (in kmol GJ^{-1}) can be determined by:

$$N_{O2}^{stoich} = N_C + \frac{N_H}{4} + N_S - \frac{N_O}{2} \tag{68}$$

The stoichiometric air can be calculated based on the wet air composition and theoretical oxygen. Thus, the minimum amount of air required for complete combustion, in kmol GJ^{-1}, is obtained as follows:

$$N_{air}^{stoich} = N_{O2}^{stoich} + \left(\frac{y_{N2}^{air}}{y_{O2}^{air}}\right) \cdot N_{O2}^{stoich} + \left(\frac{y_{H2O}^{air}}{y_{O2}^{air}}\right) \cdot N_{O2}^{stoich} \tag{69}$$

Excess air ratio is given by the gas balance since the amount of gas leaving cooling zone I does not depend on combustion itself, but on required pellet temperature. As the total energy consumption is known, the air flow can be easily converted from Nm^3 s^{-1} to Nm^3 GJ^{-1}. The combustion air ratio (α) equals the actual air entering preheating and firing zones divided by the theoretical air:

$$\alpha = \frac{\dot{V}_{air}^{actual} \cdot \frac{\Phi_{th}}{1{,}000}}{N_{air}^{stoich} \cdot \bar{V}_{air}} \tag{70}$$

Where: α = combustion air ratio [-]

$\dot{V}_{air}^{actual}$ = actual air volumetric flow rate [Nm^3 s^{-1}]

4.5 Flue Gas Composition

In order to estimate the flue gas composition, complete combustion is considered and the only species present in flue gas are CO_2, N_2, O_2, SO_2 and H_2O. The amount of water in flue gas is the sum of the water present in the gaseous and solid fuels, air moisture and the water formed by hydrogen oxidation during combustion. One approximation made was that the solid fuel is practically not dried in the updraft and downdraft drying zones. The total nitrogen is provided by combustion air and nitrogen content in the fuel. Oxygen corresponds to the amount remaining after combustion due to excess. Thus:

$$N_{CO2}^{FG} = N_C^{fuel} \tag{71}$$

$$N_{N2}^{FG} = \frac{N_N^{fuel}}{2} + \alpha \cdot y_{N2}^{air} \cdot N_{air}^{stoich} \tag{72}$$

$$N_{O2}^{FG} = \alpha \cdot y_{O2}^{air} \cdot N_{air}^{stoich} - N_{O2}^{stoich} \tag{73}$$

$$N_{SO2}^{FG} = N_S^{fuel} \tag{74}$$

$$N_{H2O}^{FG} = \frac{N_H^{fuel}}{2} + N_{H2O}^{fuel} + \alpha \cdot y_{H2O}^{air} \cdot N_{air}^{stoich} \tag{75}$$

Once the number of moles of each flue gas' component per GJ of fuel is determined, their molar fraction on a wet basis can be calculated. Flue gas properties, such as MM and molar volume, are estimate based on Equations 15 and 16.

4.6 Adiabatic Flame Temperature

The adiabatic flame temperature is the hypothetical temperature that exhaust gases would achieve if no energy is lost. Thus, the energy balance is calculated when no other heat output than the flue gas is present. Incoming energy streams are: preheated air from cooling zone I, energy provided by fuel combustion and gas sensible heat. Solid sensible heat is considered negligible. Net calorific values can directly substitute the enthalpies of combustion. Equation 76 shows the energy balance:

$$\dot{m}_{PG} \cdot LHV_{PG} + \dot{m}_{PG} \cdot \Delta h_{PG} + \dot{m}_{solid} \cdot LHV_{solid} + \dot{m}_{air} \cdot \Delta h_{air} = \dot{m}_{FG} \cdot \Delta h_{FG} \tag{76}$$

Where: $\dot{m}$ = mass flow [kg s^{-1}]

As total energy consumption is known, both gaseous and solid mass flows can be determined by dividing the energy input by the LHV in MJ kg^{-1}. Similarly, flue gas mass flow is obtained from the energy consumption. Air mass flow is directly taken from the gas balance. Gaseous fuel, wet air and flue gas sensible heats can be estimated by the weighted average of their components and respective enthalpies at fixed temperature:

$$\Delta h = \sum \nu_i \cdot \Delta h_i \tag{77}$$

Pure gaseous component enthalpies were calculated utilizing the Shomate Equation, with 298.15 K as reference temperature (Equation 40). However, the unknown temperature is an implicit parameter in the Shomate Equation. Therefore, an iterative procedure is necessary in order to obtain the results.

4.7 Product gas/Bio-SNG Flow Rates

In adiabatic flame temperature calculations previously explained, the indurating machine's energy consumption is kept constant regardless the gas fuel utilized. However, some inert components are fed into the system when product gas is utilized, such as nitrogen and/or water vapor. Therefore, the adiabatic flame temperature for product gases is expected to be lower than for NG combustion. Low temperatures in the traveling grate may depreciate important iron pellet qualities, such as mechanical resistance.

It is then necessary to determine the amount of product gas required to maintain the same temperature provided by the NG combustion. The same energy balance presented in Equation 76 is utilized; however, the total energy input is now unknown. As a consequence, product gas and flue gas flow rates are unknown. On the other hand, flue gas temperature is equal to NG's adiabatic flame temperature; thus, flue gas sensible heat is known. Air and solid fuel mass flows are kept constant since their amounts are defined by the cooling zone and green pellet requirements, accordingly. The second equation required to solve this system is the mass balance, from which the flue gas mass flow can be determined:

$$\dot{m}_{PG} + \dot{m}_{solid} + \dot{m}_{air} = \dot{m}_{FG} \tag{78}$$

Flue gas composition and properties depend on energy consumption; thus, an iterative procedure is necessary in order to close these balances. An initial value for product gas mass flow is given, from which the total energy consumption and the percentage correspondent to gas and solid fuel is calculated. Once the energy consumption and proportions are known, flue gas composition and properties can be determined. Product gas mass flow is changed until the system converges. Goal seek tool was utilized in this case.

4.8 Biomass Demand

Biomass demand was determined based on the gas flow, in kmol s^{-1}, for constant temperature. The amount of wood necessary can be obtained simply by dividing the gas flow by the total number of moles in the gas per kilogram of wood. The result, expressed in kg of wood per second, does not include ash, carbon or energy losses. When everything is considered, the total amount of wood necessary is given by:

$$\dot{m}_{wood} = \left[\frac{\dot{n}_{PG}}{\sum n_i^{PG}} \cdot CC \cdot w_C^{wood} \cdot (1 + w_{ash}^{wood})\right] \cdot \left(1 + \frac{L}{100}\right) \tag{79}$$

Where: $\dot{n}$ = total number of moles in the gas [kmol kg_{wood}^{-1}]
L = estimated energy loss [%]

For each gasification technology, the amount of wet biomass, planted area and total area required were estimated, assuming the following data:

- Eucalyptus productivity of 113 wet (56 dry) tons (with bark) per hectare after 5 years[1] – in other words, the total area required corresponds to five times the annual area;
- All harvested biomass is utilized for energy purposes;
- Losses in comminution operation are negligible;

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[1] Studies suggest that this value is quite conservative for Brazil, and it is possible to achieve 96 t ha^{-1} (dry) in 5 years with 3x3 m spacing. Refer to QUÉNO, Laurent M. R., et al. *Custo de produção das biomassas de eucalipto e capim elefante para energia*. pp. 417-426.

Total area, which includes reserves, roads, facilities, etc., doubles the required plated area, which can be estimated as follows:

$$A = \frac{5}{113} \cdot \frac{1 - w_{H2O}^{dried}}{1 - w_{H2O}^{havested}} \cdot \dot{m}_{wood} \cdot \frac{3{,}600\ \mathrm{s/h} \cdot 8{,}400\ \mathrm{h/year}}{1{,}000\ \mathrm{kg/t}} \tag{80}$$

Where: A = total planted area [ha]

4.9 Gasification Efficiency

One manner to validate the model's assumptions is estimating the hot gas efficiency (η_{hg}). According to Basu (2010), the hot gas efficiency is utilized when the product gas is directly injected in a combustion device, without need of cooling:

$$\eta_{hg} = \frac{\dot{m}_{PG} \cdot (LHV_{PG} + \Delta h_{PG})}{\dot{m}_{wood} \cdot LHV_{wood}} \tag{81}$$

In order to estimate the efficiency for oxygen gasification more accurately and be able to compare with other technologies, it is necessary to take into account the energy required for oxygen production, presented in Table 8.

4.10 Substitution Scenarios

All aforementioned calculation procedures considered that the gaseous fuel is either 100% NG or entirely provided by one of the gasification technologies. Total substitution, however, may not be feasible on short or medium terms; therefore, smaller gasification plants are proposed, such as 3, 12, 24 and 60 $\mathrm{MW_{th}}$. Calculations for each substitution scenario follow the same procedure; however, the final fuel composition (Equation 67) includes product gas, NG and solid fuel composition:

$$N_i^{fuel} = \frac{\Phi_{PG}}{\Phi_{th}} \cdot N_i^{PG} + \frac{\Phi_{NG}}{\Phi_{th}} \cdot N_i^{NG} + \frac{\Phi_{solid}}{\Phi_{th}} \cdot N_i^{solid} \tag{82}$$

4.11 Economical Evaluation Approach

The project's economical evaluation is an important parameter to compare different gasification technologies and support decision making process. As a preliminary assessment, the numbers presented here are based on basic design information and values available in the literature. Therefore, this analysis should be regarded only as an order of magnitude estimate and the actual prices may vary within ± 30-50% (Towler et al. 2012).

4.11.1 Investment Costs

The total investment is divided into total direct (TDC) and indirect costs (TIC). Direct costs mainly concern equipment purchase prices, installation and construction. These costs were based on values or correlations available in the literature. Biomass harvesting and chipping are out of the scope. Only the main equipment and installations were considered in this evaluation, divided in the following groups:

- *Biomass pretreatment*: includes wet wood chip handling and storage (H&S), drying operation and conveying after drying;
- *Gasification*: includes pre-feeding H&S, biomass feeding system, CFB gasifier, combustion reactor (for steam gasification and bio-SNG only), bed material and ash H&S, and heat exchangers (for dry product gas from steam gasification only);
- *Gasifying media supply*: includes air compressor, ASU (for oxygen) or steam generator, depending on the gasification technology;
- *Methanation*: includes gas conditioning system, methanation reactor and raw SNG upgrading, for bio-SNG production only;
- *Other inside battery limits (ISBL)*: includes nitrogen supply for instrumentation and safety equipment, as well as equipment installation costs (mainly labor);
- *Other direct costs*: include civil works and foundations, buildings, piping, automation, electrification and instrumentation (AEI), insulation and painting works.

In many cases, costs for equipment of other sizes than those presented in this study were found. In such cases, the cost curve method was utilized, with an exponential factor of 0.7 (Towler et al. 2012):

$$c_2 = c_1 \left(\frac{s_2}{s_1}\right)^{0.7} \tag{83}$$

Where: c = equipment cost
s = equipment size

In addition, some costs were found for years before 2012. In such cases, a correction factor was utilized, based on the ratio between the chemical engineering plant cost index (CEPCI) in 2012 and in the year the value was estimated. Some CEPCI values are shown in Table 15.

Indirect costs are comprised of consulting, engineering, supervision, taxes, freight, training and plant start-up, environmental and legal expenses, capitalized spares and contingency. These categories were selected as the most relevant for a biomass gasification plant by a report published by the National Renewable Energy Laboratory (Worley et al. 2012). Values originally in USD were converted to EUR with an exchange rate of 0.80 EUR/USD.

Table 15: CEPCI values from 2005 to March 2013. (Economic Indicators2013)

Year	CEPCI	Year	CEPCI	Year	CEPCI
2005	468.2	2008	575.4	2011	585.7
2006	499.6	2009	521.9	2012	584.6
2007	525.4	2010	550.8	2013 (Mar.)	568.3

Most equipment prices were taken directly from some source and simply corrected by size and/or time factors. However, some values were estimated based on correlations found in the literature. In addition, indirect costs were mostly calculated based on published percentages of TDC. Detailed references and considerations made on total direct and indirect costs (TD&IC) estimative can be found in Appendix C. All values were rounded to the closest hundred.

4.11.2 Economic Feasibility

The economic feasibility of each gasification technology was assessed based on the summation of capital costs, fuel, electricity, and operation and maintenance costs (O&M). This value corresponds to the plant's total annual costs in € MWh^{-1} of gas produced.

The capital costs of each plant correspond to the investment cost per MWh produced over the whole economic life cycle, for a given annual interest rate. Thus, the total capital cost per year is given by:

$$Capital\ Costs = \frac{TD\&IC}{8{,}000\,\frac{h}{a}\cdot gas\ output\ [MW]}\cdot AF \tag{84}$$

$$AF = \frac{IRR}{1-\left(\frac{1}{1+IRR}\right)^{l}} \tag{85}$$

Where: AF = annuity factor

IRR = interest return rate

l = economic lifecycle

The project is considered feasible if the total costs are lower than the actual NG costs, in € MWh^{-1}. Nonetheless, the NG price cannot be directly utilized in this case because energy consumption increases. For instance if a 24 MW air gasification plant is installed, only 16.8 MW of NG is actually saved since total energy consumption increases to 246 MW. Therefore, the actual NG cost depends on the gasification plant size and on the amount of NG that is actually saved. Thus:

$$NG\ cost = \frac{NG\ saved\ [\mathrm{MW}]}{Plant\ size\ [\mathrm{MW}]}\cdot NG\ price\ [€\ \mathrm{MWh^{-1}}] \tag{86}$$

One commonly utilized economical parameter is the payback period, which is defined as the number of years the investment takes to give positive return. In this case, the positive revenues from the gasification plant are the difference between the plant’s total annual costs and the NG cost. The payback period is then equal to the plant’s economical life cycle when revenues equal to zero. In addition to the payback period, a simple sensitivity analysis is performed in order to evaluate an acceptable range for the most volatile parameters (e.g. wood price and real interest rate) within which the project is still economically attractive.

5 RESULTS AND DISCUSSION

In this chapter, calculation results are presented and discussed. Firstly, the indurating machine's energy consumption and gas balance were evaluated based on operating data. NG combustion parameters, such as adiabatic flame temperature and gas flows were determined based on the methodology presented previously. In the sequence, each gasification technology is compared with NG values, focusing on the main impacts the use of product gas can have in the traveling grate. Finally, different substitution scenarios are discussed.

5.1 Indurating Machine's Energy Consumption

As stated earlier, only the third indurating machine presented is considered in this study. In order to assess its energy consumption, monthly fuel input data were utilized. The total thermal power required was estimated considering an average of 8,400 operation hours and a production of 8.4 Mt of iron pellets per year. The average energy consumption of each fuel, as well as the corresponding percentages, is shown in Table 16.

Table 16: Energy consumption in the indurating machine considered for calculations.

Source	**Energy Consumption**			**Percentage**
	GJ/t	**Mcal/t**	**MW_{th}**	**[%]**
Coal	0.165	29.7	34.6	14%
Petcoke	0.294	79.2	92.1	39%
Total Solid	**0.458**	**109**	**127**	**53%**
Heavy oil	0.000	0.06	0.08	0.0%
Natural gas	0.405	96.6	112	47%
Total burners	**0.405**	**96.7**	**112**	**47%**
Total thermal	**0.863**	**206**	**239**	**100%**

As can be seen from Table 16, at least 112 MW_{th} are required if all NG is to be substituted by renewable resources. The third indurating machine has a total of 46 burners, from which 42 are used under normal operating conditions. As a result, each burner represents an energy load of approximately 2.67 MW_{th}, usually provided by NG. This corresponds to a total of 173 Nm^3 min^{-1} of NG or 4.12 Nm^3 min^{-1} per burner.

It is important to highlight that the use of product gas causes injection of considerable amounts of inert substances in the kiln. Consequently, the thermal input would have to be increased to prevent temperature decrease.

5.2 Gas Balance

The gas balance was calculated based on operating data, under assumptions addressed in Chapter 4. The most important value taken from the gas balance is the air flow in preheating and firing zones. Total gas flow entering the control volume was estimated at 22,240 Nm^3 min^{-1}, including air and NG. In fact, the amount of NG appears to be negligible; however, as it can be easily evaluated, it is subtracted in all cases in order to accurately estimate the actual air input. NG flow rate depends on the kiln energy consumption, which was considered to be the current average (239 MW), from which 112 MW corresponds to NG. As a result, it is estimated that 22,067 Nm^3 min^{-1} or 368 Nm^3 s^{-1} of air flows through the control volume.

5.3 NG and Solid Fuel Combustion

NG composition was shown in Table 13. As can be seen from this table, combustibles in NG are only methane (CH_4), ethane (C_2H_6) and propane (C_3H_8). Their standard enthalpies of combustion are 803 kJ mol^{-1}, 1,560 kJ mol^{-1} and 2,220 kJ mol^{-1} respectively. Utilizing Equation 14, the LHV was estimated at approximately 872 MJ $kmol^{-1}$ of NG or 39.0 MJ Nm^{-3}.

According to the third indurating machine's energy consumption, 47% of the total energy in the preheating and firing zones is provided by NG and the remaining 53% by the solid blend. Following Equation 67, the final fuel (solid + gas) LHV was estimated at 568 MJ $kmol^{-1}$ or 40.2 MJ kg^{-1}. Considering the solid fuel properties shown in Table 14 and Equations 58 to 62, the final fuel composition was determined. In the sequence, the stoichiometric amount of oxygen was calculated, utilizing Equation 68. Table 17 shows the results per GJ of fuel.

Wet air composition was determined based on Equations 18 to 22. The results, as well as some basic properties of wet air (obtained from Equations 15 and 16) are presented in Table 18. These results were combined in Equation 69, from which the amount of stoichiometric wet air is obtained.

Table 17: Fuel composition and stoichiometric oxygen demand for NG as gaseous fuel.

Element	Gas kmol/GJ$_{gas}$	Solid kmol/GJ$_{solid}$	Gas + Solid kmol/GJ$_{fuel}$	Gas + Solid %-mol	Stoch. O_2 kmol/GJ$_{fuel}$
C	1.267	1.879	1.592	34.4%	1.592
H	4.760	1.125	2.833	61.2%	0.708
O	0.021	0.128	0.078	1.68%	-0.039
N	0.028	0.029	0.029	0.62%	0.000
S	0.000	0.040	0.021	0.46%	0.021
H_2O	0.000	0.152	0.081	1.74%	0.000
TOTAL	**-**	**-**	**4.633**	**100%**	**2.283**

The amount of air entering the preheating and firing zones is 368 Nm^3 s^{-1} and the total energy consumption is 0.239 GJ s^{-1}. Based on these values, it was found that the total combustion air is equivalent to 1,539 Nm^3 GJ^{-1}. As the stoichiometric air is estimated at 247 Nm^3 GJ^{-1}, the air ratio is 6.23 (Equation 70). This number is relatively high because the amount of air is controlled by cooling zone requirements, not by the preheating/firing zones.

Table 18: Properties of wet air.

Component	%-mol	MM [kg/kmol]	Molar V [Nm^3/kmol]
H_2O	1.88%	18.02	17.83
O_2	20.6%	32.00	22.39
N_2	77.5%	28.01	22.40
TOTAL AIR (wet)	**100%**	**28.65**	**22.31**

Flue gas composition was obtained through Equations 71 to 75. Mass and volume per GJ of fuel were calculated by the weighted average between the composition and pure component properties. Results are shown in Table 19. Based on the found composition and Equations 15 and 16, the flue gas molar mass is estimated at 28.67 kg kmol^{-1} while the molar volume is 22.21 Nm^3 kmol^{-1}.

The adiabatic flame temperature for NG and solid fuel was calculated based on the energy balance shown in Equation 76. The air mass flow was obtained by multiplying the total energy consumption (0.239 GJ s^{-1}) by the total amount of air (1,976 kg GJ^{-1}), resulting in 472.2 kg

s^{-1}. Similarly, the flue gas mass flow is given by the total energy consumption times the amount of flue gas (2,002 kg GJ^{-1}), resulting in 478.4 kg s^{-1}.

Table 19: Flue gas generation per GJ of fuel for NG combustion.

Species	**Flue Gas kmol/GJ$_{fuel}$**	**%-mol (wet)**	**Mass kg/GJ$_{fuel}$**	**%-w (wet)**	**Volume Nm^3/GJ$_{fuel}$**	**%-vol (wet)**
CO_2	1.592	2.28%	70.05	3.50%	35.43	2.29%
N_2	53.48	76.6%	1 498	74.8%	1 198	77.2%
O_2	11.93	17.1%	381.7	19.1%	267.1	17.2%
SO_2	0.021	0.03%	1.371	0.07%	0.468	0.03%
H_2O	2.792	4.00%	50.30	2.51%	49.80	3.21%
TOTAL	**69.81**	**100%**	**2 002**	**100%**	**1 551**	**100%**

Air enthalpy was calculated by combining Equations 40 and 77, considering a temperature of 1,000 °C (1,273 K). NG was considered to be at 298 K, and therefore its sensible heat equals to zero. The heat flows from gaseous and solid fuels are also known, 112 and 127 MW accordingly. The only variable is then the flue gas temperature, which is implicit and must be determined by an iterative procedure involving the Shomate Equation. Enthalpy calculation results for all cases are shown in Appendix B.

The adiabatic flame temperature under the described conditions was found to be 1,373 °C (1,646 K). This value is consistent with the traveling grate temperature in the firing zone, which is supposed to be 1,363 °C according to operating data. The detailed energy balance values are summarized in Table 20.

Table 20: Energy balance for NG as gaseous fuel.

Phase	**Flow kg/s**	**Temperature K**	**Enthalpy kJ/kg**	**Heat kW**
Air	472.2	1 273	1 085	512 403
Gas, comb.	2.329	-	48 227	112 330
Solid, comb.	3.827	-	33 103	126 670
Flue gas	478.4	1 646	-1 571	-751 403

Besides the temperature, gas flows are the other important parameter to assess the impacts of gasification technology in the indurating machine. In order to compare values for NG and product gas, flow rates are shown in different units in Table 21.

Table 21: Air, gas, solid and flue gas flow rates in several units.

Phase	**kg/s**	**kmol/s**	**Nm^3/s**	**Nm^3/min**
Air	472.2	16.48	367.8	22 067
Gas	2.329	0.129	2.880	173
Solid	3.827	0.425	-	-
Flue gas	478.4	16.69	370.6	22 237
TOTAL IN	-	-	370.7	22 240

5.4 Air Gasification

Air gasification is the simplest technology among all the options assessed in this study. In this case, biomass is firstly dried from 50% to 15% MC. Inside the gasifier the biomass is partially combusted in order to provide heat for gasification reactions. The ER, defined by Equation 1, was chosen to be 30% based on typical values. In addition, the gasifier temperature was considered to be 850 °C and 95% of the carbon input was assumed to be converted into gas; in other words, 5% of all carbon input is removed with ashes. The product gas leaves the cyclone at 873 K (600 °C) and it is burnt in the indurating machine without any gas cleaning operation. Figure 24 presents the process flow chat and summarizes the main considerations.

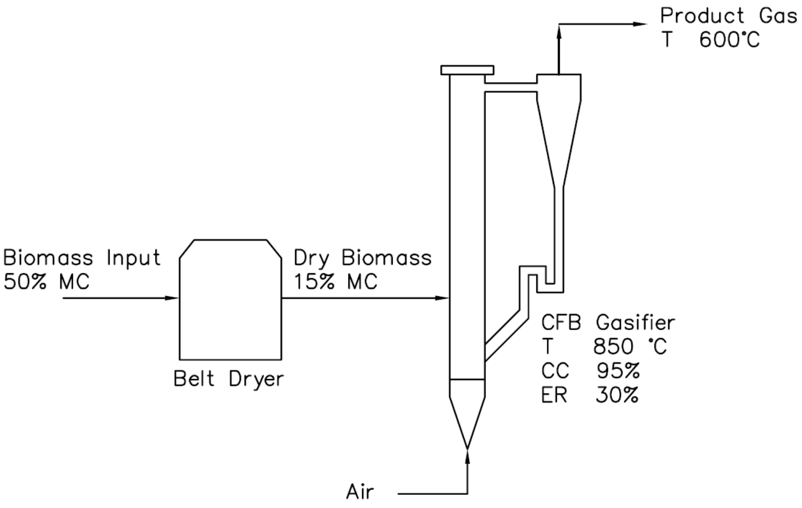


Figure 24: Air gasification flow chart.

In order to determine the gasifier input composition, the stoichiometric amount of oxygen for complete combustion was calculated according to Equation 17. The gasifier input was obtained from Equations 23 to 25 based on the stoichiometric O_2, ER and wet air composition. Results are shown in Table 22.

Table 22: Air gasifier inlet (wood + air) elemental composition.

Element	**Wood** **kmol/kg$_{wood}$**	**Stoch. O2** **kmol/kg$_{wood}$**	**Input (Wood + Air)** **kmol/kg$_{wood}$**
C	0.0366	0.0366	0.0366
H	0.0522	0.0130	0.0522
O	0.0236	-0.0118	0.0463
N	0.0001	0.0000	0.0855
H_2O	0.0072	0.0000	0.0083
TOTAL	**0.1196**	**0.0378**	**0.2287**

The product gas composition was obtained by solving the system of Equations 29 to 33. Methane molar fraction on a dry basis was set as 4.70%. Results are shown in Table 23. Once the composition was determined, the product gas LHV could be estimated based on the heat of combustion from reactions R4, R7 and R8 (Table 6) and Equation 14. As a result, the LHV in this case was found to be 5.32 MJ Nm^{-3} or 116 MJ $kmol^{-1}$.

Table 23: Product gas composition for air gasification.

Component	**Amount** **kmol/kg$_{wood}$**	**%-mol** **(wet)**	**%-mol** **(dry)**	**Mass** **kg/kmol$_{gas}$**	**Volume** **Nm^3/kmol$_{gas}$**
CH_4	0.0043	4.14%	4.70%	0.6649	0.9277
CO	0.0186	18.0%	20.4%	5.043	4.034
CO_2	0.0119	11.5%	13.0%	5.057	2.558
H_2	0.0136	13.2%	14.9%	0.2654	2.955
N_2	0.0427	41.4%	46.9%	11.59	9.267
H_2O	0.0122	11.8%	13.4%	2.130	2.109
TOTAL	**0.1033**	**100.0%**	**113.4%**	**24.75**	**21.85**

The final fuel (gas + solid) composition was calculated in the same manner as NG, considering data shown in Table 14 and Table 23. Gas was considered to provide 47% of the total energy in the preheating and firing zones and the remaining 53% by the solid blend described before. All combustion and flue gas calculations were performed in the same way as for NG. However, when closing the energy balance shown in Equation 76, the gas sensible heat was taken into account since it enters the indurating machine at 600 °C (873 K), immediately after solid removal in the cyclone. Injection of hot gas increases thermal efficiency and also avoids tar condensation and consequent pipeline clogging.

The adiabatic flame temperature for 112 MW of air gasification product gas, under the described conditions, was found to be 1,347 °C (1,620 K). This value is consistently 26 °C lower than the one obtained for NG since the product gas LHV is much lower. In addition, considerable amounts of inert N_2 are injected in the system. Nonetheless, one of the most important operating parameters that assure the iron pellet quality is temperature. Thus, more product gas must be injected into the induration kiln. The required amount of product gas was calculated based on the restriction that NG theoretical temperature (1,373 °C) is maintained. It is important to notice that an increase in the gas flow in the firing zone may increase the heat exchange between gas and solid phase, causing the pellet's temperature to increase. However, these effects were completely neglected in the model proposed here.

In order to determine the required amount of product gas, an iterative procedure is necessary. The total energy consumption and its proportion between gas and solid are not known in this case; therefore, the percentage of gas in the fuel needs an initial guess. From this guessed value, the fuel composition, air ratio and flue gas composition are obtained. After convergence was reached, it was found that 52% of the energy is provided by cold product gas. Table 24 shows the results for fuel composition and stoichiometric oxygen demand per GJ of fuel, after convergence. LHV for the final fuel blend was estimated at 18.5 MJ Nm^{-3} or 204 MJ $kmol^{-1}$.

The stoichiometric amount of air was calculated based on wet air properties shown in Table 18 and in Equation 69. The mass flow of combustion air remains 472.2 kg s^{-1}; however, the new cold gas plus solid consumption is 0.261 GJ s^{-1}. As a result, 1,408 Nm^3 GJ^{-1} of air circulate in the system. The stoichiometric air was estimated at 227 Nm^3 GJ^{-1}, resulting in an air ratio of 6.21.

Table 24: Fuel composition and stoichiometric oxygen demand for air gasification product gas.

Element	**Gas** **kmol/GJ$_{gas}$**	**Solid** **kmol/GJ$_{solid}$**	**Gas + Solid** **kmol/GJ$_{fuel}$**	**Gas + Solid** **%-mol**	**Stoch. O_2** **kmol/GJ$_{fuel}$**
C	2.893	1.879	2.401	21.8%	2.401
H	3.690	1.125	2.447	22.2%	0.612
O	3.525	0.128	1.878	17.0%	-0.939
N	7.117	0.029	3.680	33.4%	0.000
S	0.000	0.040	0.020	0.18%	0.020
H_2O	1.017	0.152	0.598	5.42%	0.000
TOTAL	**-**	**-**	**11.02**	**100%**	**2.094**

The flue gas flow rate was calculated based on the same premises and equations utilized for NG. The results are shown in Table 25. In this case, the molecular mass of the flue gas was estimated at 28.80 kg kmol^{-1} while the molar volume was found to be 22.19 Nm^3 kmol^{-1}.

Table 25: Flue gas generation per GJ of fuel for air gasification case.

Species	**Flue Gas** **kmol/GJ$_{fuel}$**	**%-mol** **(wet)**	**Mass** **kg/GJ$_{fuel}$**	**%-w** **(wet)**	**Volume** **Nm^3/GJ$_{fuel}$**	**%-vol** **(wet)**
CO_2	2.401	3.58%	105.7	5.47%	53.46	3.58%
N_2	50.75	75.7%	1 422	73.6%	1 130	75.7%
O_2	10.91	16.3%	349.1	18.1%	242.9	16.3%
SO_2	0.020	0.03%	1.254	0.06%	0.436	0.03%
H_2O	3.005	4.48%	54.14	2.80%	66.91	4.48%
TOTAL	**67.09**	**100%**	**1 932**	**100%**	**1 494**	**100%**

As temperatures and composition of all gaseous streams were known, the specific enthalpies could be calculated by means of the Shomate Equation. Specific enthalpy results after iteration are shown in Appendix B. Other known parameters in Equations 76 were: LHV of solid and gaseous fuels, and air and solid mass flows, which were kept as constants from NG calculations. Thus, the only unknowns were then the product gas and flue gas mass flows, which could be found by solving Equations 76 and 78 simultaneously. The results are shown in Table 26.

Table 26: Energy balance for air gasification product gas.

Phase	Flow kg/s	Temperature [K]	Enthalpy kJ/kg	Heat kW
Air	472.2	1 273	1 085	512 403
Gas, sensible	28.64	873	780	22 337
Gas, comb.	28.64	-	4 698	134 564
Solid, comb.	3.827	-	33 103	126 670
Flue gas	504.7	1 646	1 577	-795 973

At this point, the total energy consumption and the solid/gas percentages could be determined. If they were not equal to initial values, the calculation restarted with new values, until they converge. In this case, it was found that 284 MW are required to maintain the NG temperature, from which 157 MW (55%) is provided by hot product gas. It represents an increase of 18.6% in energy consumption.

Specific and volumetric flows of air, product gas and flue gas are shown in different units in Table 27. The increased percentages compared to NG values (Table 21) are also presented. As can be seen from the table, 778% more fuel gas would be injected in the case of air gasification. Although this increase is very high, the fuel gas inlet is still much smaller than the air flow rate. In fact, the total gas input in preheating and firing zones increases only approximately 6% and flue gas flow rate increases less than 5%. Significant impact is therefore expected only in the burners, which would possibly require substitution/adaption; for example, flow area should increase by increasing pipe diameter. In addition, the burner’s material would be required to resist higher temperatures (600 °C in this case).

Table 27: Air, gas, solid and flue gas flow rates in several units, for air gasification case.

Phase	kg/s	kmol/s	Nm^3/s	Nm^3/min	Increase [%]
Air	472.2	16.48	367.8	22 067	0.00%
Gas	28.64	1.157	25.29	1 517	778%
Solid	3.827	0.425	-	-	-
Flue gas	504.7	17.53	388.9	23 332	4.93%
TOTAL IN	-	-	393.1	23 585	6.05%

The wood input was calculated based on Equation 79. If the losses are estimated at 2% of the net energy content in wood, 11.8 kg s^{-1} (42.4 t h^{-1}) of wood chips with 15% MC would be required. This corresponds to approximately 27,000 ha of planted area, and a total area of 54,000 ha. It is relevant to highlight that this value can be further reduced by short-rotation forests and/or denser planting.

Finally, the LHV of wood was calculated through Equations 11 to 13. It was found that the energy content of eucalyptus wood considered in this study is 15.6 MJ kg^{-1} with 15% MC. Thus, the gasifier energy input and output are 184 MW and 157 MW, respectively. The hot gas efficiency for air gasification is 85.4%. This value does not take into consideration other possible incoming energy streams, such as air preheating.

5.5 Oxygen Gasification

Oxygen gasification is one alternative to avoid dilution of the product gas by nitrogen. Same calculations were performed as for air gasification although some assumptions were changed. For instance, gasifier temperature was estimated at 950 °C, as less cold gases are fed into the system. ER and CC were kept at the same values as air gasification. In order to control temperature and favor hydrogen formation, SOR was set at 1 kg of steam per kg of oxygen (van der Meijden et al. 2010). Figure 25 shows the oxygen gasification main parameters.

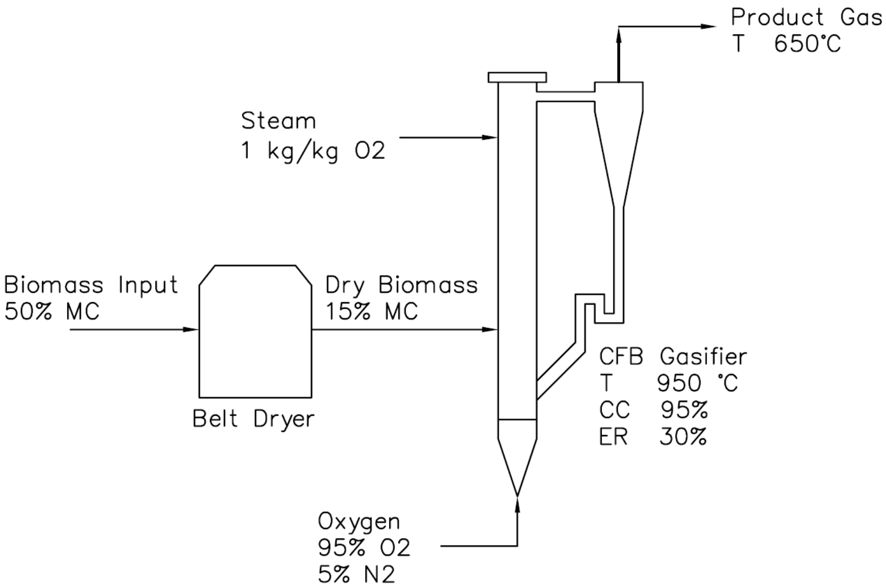


Figure 25: Oxygen gasification flow chart.

The calculations of product gas composition for oxygen gasification are very similar to those for air as gasifying medium. Equation 17 is still valid, and it was considered that the gasifying medium is composed of 95% O_2 and 5% N_2. Stoichiometric oxygen was determined according to Equations 26 to 28. Results are shown in Table 28.

Table 28: Oxygen gasifier inlet (wood + oxygen) elemental composition.

Element	**Wood** **kmol/kg$_{wood}$**	**Stoch. O_2** **kmol/kg$_{wood}$**	**Wood + O_2 + H_2O** **kmol/kg$_{wood}$**
C	0.0366	0.0366	0.03656
H	0.0522	0.0130	0.05218
O	0.0236	-0.0118	0.04626
N	0.0001	0.0000	0.00126
H_2O	0.0072	0.0000	0.01363
TOTAL	**0.1196**	**0.0378**	**0.1499**

The product gas composition was obtained from Equations 29 to 33. Methane molar fraction was considered to be 1.0% on a dry basis. All other assumptions remained the same. The iterative procedure was developed in the same manner, and its results are presented in Table 29. LHV was found to be 8.22 MJ Nm^{-3} or 175 MJ $kmol^{-1}$.

Table 29: Product gas composition for oxygen gasification.

Component	**Amount** **kmol/kg$_{wood}$**	**%-mol** **(wet)**	**%-mol** **(dry)**	**Mass** **kg/kmol$_{gas}$**	**Volume** **Nm^3/kmol$_{gas}$**
CH_4	0.0006	0.76%	1.00%	0.122	0.170
CO	0.0264	35.7%	47.1%	9.995	7.995
CO_2	0.0078	10.5%	13.9%	4.631	2.343
H_2	0.0207	27.9%	36.9%	0.563	6.274
N_2	0.0006	0.85%	1.12%	0.238	0.190
H_2O	0.0179	24.2%	32.0%	4.367	4.323
TOTAL	**0.0740**	**100%**	**132%**	**19.92**	**21.30**

For comparison with NG, the adiabatic flame temperature for product gas combustion was calculated considering data shown in Table 14 and Table 29. Gas was considered to provide

47% of the total energy in preheating and firing zones and the remaining 53% by the solid blend, as described before. All combustion and flue gas calculations were performed in the same way as for air gasification. However, product gas was considered to enter the indurating machine at 650 °C (923 K) since gasification temperature is higher.

The adiabatic flame temperature for 112 MW of product gas, under the described conditions, was found to be 1,365 °C (1,638 K). This value is only 8 °C lower than that one obtained for NG, even though product gas net energy content is quite lower. In addition, it is 18 °C higher than the value for air gasification since very low amounts of inert N_2 are injected in the system.

The product gas mass flow rate required to maintain NG temperature was calculated by the same iterative procedure. The proportion of gas and solid was first guessed. From this value, the final fuel composition, air ratio and flue gas composition were obtained. After convergence was reached, it was found that 48% of the energy is provided by cold product gas if the solid fuel mass flow is kept constant. Table 30 shows the results for fuel composition and stoichiometric oxygen demand per GJ of fuel. The final fuel LHV was estimated at 21.4 MJ Nm^{-3} or 238 MJ $kmol^{-1}$.

Table 30: Fuel composition and stoichiometric oxygen demand for oxygen gasification.

Element	**Gas** **kmol/GJ$_{gas}$**	**Solid** **kmol/GJ$_{solid}$**	**Gas + Solid** **kmol/GJ$_{fuel}$**	**Gas + Solid** **%-mol**	**Stoch. O_2** **kmol/GJ$_{fuel}$**
C	2.683	1.879	2.267	32.7%	2.267
H	3.366	1.125	2.207	31.8%	0.552
O	3.241	0.128	1.630	23.5%	-0.815
N	0.097	0.029	0.062	0.90%	0.000
S	0.000	0.040	0.021	0.30%	0.021
H2O	1.385	0.152	0.747	10.8%	0.000
TOTAL	**-**	**-**	**6.93**	**100%**	**2.024**

The stoichiometric amount of air was calculated based on the wet air properties shown in Table 18 and Equation 69. Combustion air mass flow rate remains 472.2 kg s^{-1}; however, the energy consumption is 0.245 GJ s^{-1}. As a result, 1,502 Nm^3 GJ^{-1} of air circulates in the system. The stoichiometric air was estimated at 219 Nm^3 GJ^{-1}, resulting in an air ratio of 6.85.

Flue gas flow rate was calculated based on the same premises and equations utilized for air gasification. Results are shown in Table 31. In this case, flue gas molar mass was estimated at 28.78 kg $kmol^{-1}$ while the molar volume was found to be 22.19 Nm^3 $kmol^{-1}$.

Table 31: Flue gas generation per GJ of fuel for oxygen gasification case.

Species	**Flue Gas kmol/GJ_{fuel}**	**%-mol (wet)**	**Mass kg/GJ_{fuel}**	**%-w (wet)**	**Volume Nm^3/GJ_{fuel}**	**%-vol (wet)**
CO_2	2.267	3.26%	99.77	4.99%	50.47	3.26%
N_2	52.22	75.2%	1 463	73.2%	1 163	75.2%
O_2	11.85	17.1%	379.2	19.0%	263.8	17.1%
SO_2	0.021	0.03%	1.338	0.07%	0.465	0.03%
H_2O	3.114	4.48%	56.10	2.81%	69.33	4.48%
TOTAL	**69.47**	**100%**	**1 999**	**100%**	**1 547**	**100%**

After flue gas composition was determined, mass and energy balances could be solved by combining Equations 76 and 78. As temperatures and composition of all gaseous streams were known, the specific enthalpies were calculated by the Shomate Equation. Final specific enthalpy results are shown in Appendix B. Other known parameters in Equations 76 were: LHV of solid and gaseous fuels, and air and solid mass flows, which were kept constant from NG calculations. As a result, the only unknowns were then product gas and flue gas mass flows. The results are shown in Table 32.

Table 32: Energy balance for oxygen gasification product gas.

Phase	**Flow kg/s**	**Temperature [K]**	**Enthalpy kJ/kg**	**Heat kW**
Air	472.2	1 273	1 085	512 403
Gas, sensible	13.44	923	1 054	14 171
Gas, comb.	13.44	-	8 790	118 159
Solid, comb.	3.827	-	33 103	126 670
Flue gas	489.5	1 646	1 576	-771 403

For oxygen gasification, it was found that 259 MW are required to maintain the NG temperature, from which 132 MW (51%) are provided by hot product gas. These values consider that the solid fuel mass flow rate is kept constant from NG combustion. It represents an increase of 8.37% in the energy consumption compared to NG.

Air, product gas and flue gas specific and volumetric flows are shown in different units in Table 33. The increase percentages compared to NG values (Table 21) are also presented. As can be seen from the table, 339% more fuel gas would be injected in the case of oxygen gasification. Although this increase is very high, the fuel gas inlet is still much smaller than the air flow rate. The total gas input in preheating and firing zones increases only 3% and flue gas flow rate increases less than 2%. Therefore, significant impact is expected only in the burners, which would probably require substitution/adaption. In addition, burner's material would be required to resist higher temperatures (650 °C in this case).

Table 33: Air, gas, solid and flue gas flow rates in several units, for oxygen gasification case.

Phase	kg/s	kmol/s	Nm^3/s	Nm^3/min	Increase [%]
Air	472.2	16.48	367.8	22 067	0.00%
Gas	13.44	0.675	14.37	862	399%
Solid	3.827	0.425	-	-	-
Flue gas	489.5	17.01	377.4	22 644	1.83%
TOTAL IN	-	-	382.2	22 930	3.10%

The wood input was calculated based on Equation 79. Again, the losses were estimated at 2% of wood's net energy content, and 9.59 kg s^{-1} (34.5 t h^{-1}) of wood chips with 15% MC would be required. This corresponds to approximately 22,000 ha of planted area, and a total area of 44,000 ha.

The oxygen gasification energy input and output are 150 MW and 132 MW, respectively. Thus, the hot gas efficiency for oxygen gasification equals 88.4%. This value does not take into consideration other possible incoming energy streams such as the energy required for oxygen production. According to Table 8, 300 kWh per ton of oxygen are necessary for the air purification process with highest energy consumption, resulting in 3,761 kW required for oxygen purification. As a result, the energy input reaches 153 MW. Thus, the net efficiency for oxygen gasification equals 86.3%, which is only slightly higher than air gasification.

5.6 Steam Gasification

In steam gasification, only wood and steam are fed into the system and the process energy is provided indirectly by char combustion associated with bed material circulation. Gasifier temperature is usually lower in steam gasification, and it was estimated at 800 °C. The SB ratio, defined by Equation 2, was set as 0.5 kg per kg of dry wood (0.0278 kmol kg^{-1}, dry) according to typical values (Hofbauer et al. 2000). This corresponds to a total input of 0.0319 kmol kg^{-1} of wood with 15% MC.

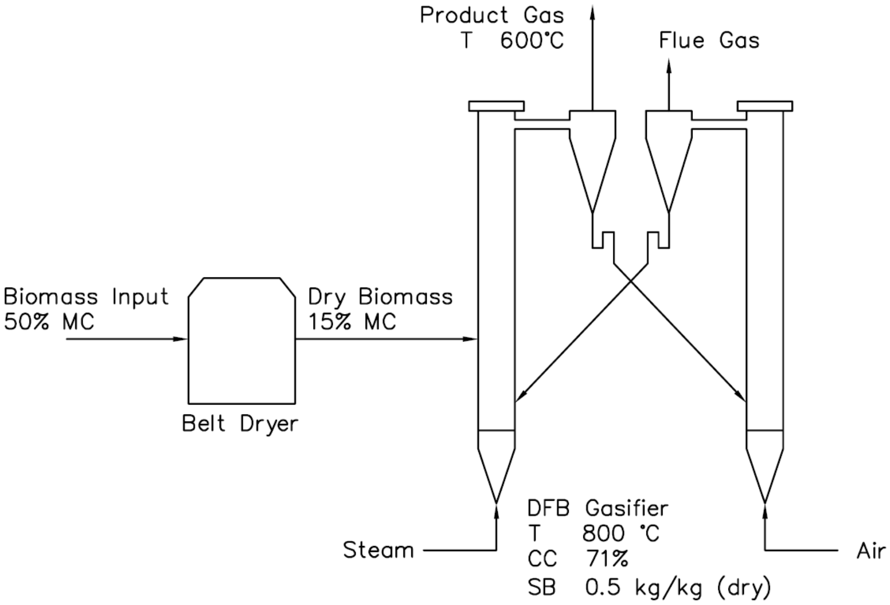


Figure 26: DFB steam gasification flow chart.

5.6.1 Wet Product Gas

As temperature was set at 800 °C, CC in this case is estimated at 71% according to Equation 35. The elemental composition of the product gas is calculated similarly to preceding cases, and results are shown in Table 34. Note that only steam is added in a certain amount to obtain a total of 0.0319 kmol per kg of wood in the gasifier inlet.

Table 34: Steam gasifier inlet (wood + steam) elemental composition.

Element	Wood kmol/kg$_{wood}$	Steam kmol/kg$_{wood}$	Total kmol/kg$_{wood}$
C	0.03656	0.00000	0.0366
H	0.05218	0.00000	0.0522
O	0.02356	0.00000	0.0236
N	0.00006	0.00000	6.26E-05
H_2O	0.00724	0.02468	0.0319
TOTAL	**0.1196**	**0.02468**	**0.1443**

Equations 29 to 33 were utilized to calculate the product gas composition. Methane molar fraction was considered to be 11% on a dry basis. Results are presented in Table 35. LHV was found to be 8.32 MJ Nm^{-3} or 174 MJ kmol^{-1}.

Table 35: Product gas composition for steam gasification.

Component	Amount kmol/kg$_{wood}$	%-mol (wet)	%-mol (dry)	Mass kg/kmol$_{gas}$	Volume Nm^3/kmol$_{gas}$
CH_4	0.0054	7.32%	11.0%	1.174	1.638
CO	0.0103	14.0%	21.0%	3.919	3.135
CO_2	0.0103	14.1%	21.2%	6.215	3.144
H_2	0.0228	31.1%	46.7%	0.626	6.971
N_2	0.0000	0.04%	0.06%	0.012	0.010
H_2O	0.0245	33.5%	50.3%	6.031	5.971
TOTAL	**0.0733**	**100%**	**150.3%**	**17.98**	**20.87**

In order to compare the adiabatic flame temperature, final fuel (gas + solid) composition was calculated in the same manner as for air/oxygen gasification, considering data shown in Table 14 and Table 35. Combustion and flue gas calculations were also performed in the same way as for previous gasification technologies. The product gas was considered to enter the indurating machine at 600 °C (873 K). The adiabatic flame temperature for 112 MW of steam gasification wet product gas was found to be 1,361 °C (1,634 K). This value is 12 °C lower than that obtained for NG although the product gas LHV is notably lower. In addition, it is 4 °C lower than the value for oxygen gasification, probably due to the high water content and lower LHV.

The amount of product gas able to maintain the NG temperature was calculated by the same iterative procedure. The proportion between gas and solid was initially guessed. From this value, the final fuel composition, air ratio and flue gas composition were obtained. After convergence was reached, it was found that 49% of the energy is provided by cold product gas if the solid fuel mass flow is kept constant. Table 36 shows the results for fuel composition and stoichiometric oxygen demand per GJ of fuel, after convergence. LHV for the fuel blend was estimated at 21.6 MJ Nm^{-3} or 237 MJ kmol^{-1}.

Table 36: Fuel composition and stoichiometric oxygen demand for steam gasification.

Element	**Gas** kmol/GJ$_{gas}$	**Solid** kmol/GJ$_{solid}$	**Gas + Solid** kmol/GJ$_{fuel}$	**Gas + Solid** %-mol	**Stoch. O$_2$** kmol/GJ$_{fuel}$
C	2.040	1.879	1.958	26.4%	1.958
H	5.262	1.125	3.154	42.4%	0.788
O	2.432	0.128	1.258	16.9%	-0.629
N	0.005	0.029	0.017	0.23%	0.000
S	0.000	0.040	0.021	0.28%	0.021
H2O	1.928	0.152	1.023	13.8%	0.000
TOTAL	**-**	**-**	**7.431**	**100%**	**2.138**

The stoichiometric amount of air was calculated based on wet air properties presented in Table 18 and Equation 69. The mass flow of combustion air remains 472.2 kg s^{-1}; however, the cold gas and solid fuel consumption is 0.248 GJ s^{-1}. As a result, 1,480 Nm^{3} GJ^{-1} of air circulate in the system. The stoichiometric air is estimated at 232 Nm^{3} GJ^{-1}, resulting in an air ratio of 6.39.

Flue gas flow rate calculation is based on the same assumptions and equations utilized for air/oxygen gasification. The results are shown in Table 37. In this case, the molar mass of the flue gas was found to be 28.59 kg kmol^{-1}, while the molar volume is 22.14 Nm^{3} kmol^{-1}.

Table 37: Flue gas generation per GJ of fuel for steam gasification case.

Species	Flue Gas kmol/GJ$_{fuel}$	%-mol (wet)	Mass kg/GJ$_{fuel}$	%-w (wet)	Volume Nm^3/GJ$_{fuel}$	%-vol (wet)
CO_2	1.958	2.85%	86.18	4.38%	43.60	2.85%
N_2	51.43	74.8%	1 441	73.3%	1 145	74.8%
O_2	11.53	16.8%	368.9	18.8%	256.7	16.8%
SO_2	0.021	0.03%	1.319	0.07%	0.458	0.03%
H_2O	3.845	5.59%	69.26	3.52%	85.60	5.59%
TOTAL	**68.78**	**100%**	**1 966**	**100%**	**1 531**	**100%**

After the flue gas composition was obtained, mass and energy balances could be solved. As temperatures and composition of all gaseous streams are known, the specific enthalpies can be calculated by the Shomate Equation. Final specific enthalpy results are shown in Appendix B. As in the previous calculations, product gas and flue gas mass flow rates were obtained solving Equations 76 and 78 simultaneously. Results are shown in Table 38.

Table 38: Energy balance for steam gasification product gas.

Phase	Flow kg/s	Temperature K	Enthalpy kJ/kg	Heat kW
Air	472.2	1 273	1 085	512 403
Gas, sens.	12.61	873	1 145	14 447
Gas, comb.	12.61	-	9 659	121 837
Solid, comb.	3.827	-	33 103	126 670
Flue gas	488.7	1 646	1 587	-775 358

For steam gasification wet product gas combustion, 263 MW are required to maintain the NG temperature, from which 136 MW (52%) are provided by hot product gas. It represents an increase of 10.0% in the energy consumption.

Specific and volumetric flows of air, product gas and flue gas are shown in different units in Table 39. The increase percentages compared to NG values (Table 21) are also presented. As can be seen from the table, 408% more fuel gas would be injected in the case of steam gasification, without product gas drying. Although this increase is quite high, the fuel gas inlet is still much lower than the air flow rate. In fact, the total gas input in preheating and firing zones increases approximately 3% and flue gas flow rate increases approximately 2%. Thus, significant impact is expected only in the burners, which would likely require substitution/adaption. In addition, the burner's material would be required to resist higher temperatures (600 °C in this case).

Table 39: Air, gas, solid and flue gas flow rates in several units, for steam gasification case.

Phase	kg/s	kmol/s	Nm^3/s	Nm^3/min	Increase [%]
Air	472.2	16.48	367.8	22 067	0.00%
Gas	12.61	0.702	14.64	879	408%
Solid	3.827	0.425	-	-	-
Flue gas	488.7	17.09	378.4	22 704	2.10%
TOTAL IN	-	-	382.4	22 946	3.17%

The wood input was calculated based on Equation 79. Again, the losses were estimated at 2% of the energy content in the wood, and 11.1 kg s^{-1} (40.0 t h^{-1}) of wood chips with 15% MC would were found to be required. This corresponds to approximately 25,500 ha of planted area, and a total area of 51,000 ha.

The steam gasification energy input and output are 173 MW and 136 MW, respectively. Thus, the hot gas efficiency for steam gasification equals 78.7%. This value does not take into consideration other important incoming energy streams, such as heat from steam. If steam is considered, the net efficiency is estimated at 72.7%.

5.6.2 Product Gas Drying

As shown in Table 35, the amount of water in the product gas from steam gasification is considerably higher than other technologies. Therefore, drying before combustion may be desirable. Drying the product gas would allow the net energy content to increase; on the other hand, sensible heat would be lost. In order to assess the benefits and drawbacks of product gas drying, the adiabatic flame temperature for the dried gas was estimated. It was considered that the gas would be cooled to 50 °C. After defining temperature, the water molar fraction could be determined based on the water saturation pressure, as shown in Equation 19. At 323 K, the saturation pressure is 12,351 Pa and the water molar fraction is 12.19%. To achieve this value, 0.0178 kmol of H_2O per kg of wood needs to be removed. The resulting molar, mass and volume fractions are shown in Table 40. After drying, the LHV increased 24%, from 174 MJ kmol^{-1} to 229 MJ kmol^{-1}.

Table 40: Product gas composition after drying, for steam gasification.

Component	**%-mol (dried)**	**Mass kg/kmol$_{gas}$**	**%-w (dried)**	**Volume Nm^3/kmol$_{gas}$**	**%-vol (dried)**
CH_4	9.66%	1.550	8.63%	2.162	9.90%
CO	18.5%	5.173	28.8%	4.138	18.9%
CO_2	18.6%	8.203	45.7%	4.150	19.0%
H_2	41.0%	0.826	4.60%	9.201	42.1%
N_2	0.06%	0.016	0.09%	0.013	0.06%
H_2O	12.2%	2.196	12.2%	2.174	9.96%
TOTAL	**100%**	**17.96**	**100%**	**21.84**	**100%**

The adiabatic flame temperature calculation was done in exactly the same manner as for the wet product gas. Although drying increases the product gas net energy content, the adiabatic flame temperature actually decreases to 1,354 °C due to sensible heat loses. In order to maintain the temperature achieved by NG combustion, product gas and flue gas mass flow rates were obtained by solving mass and energy balances (Equations 76 and 78) simultaneously. Results are shown in Table 41.

Table 41: Energy balance for steam gasification product gas after drying.

Phase	**Flow kg/s**	**Temperature K**	**Enthalpy kJ/kg**	**Heat kW**
Air	472.2	1 273	1 085	512 403
Gas, sens.	9.97	348	89.23	889
Gas, comb.	9.97	-	12 759	127 191
Solid, comb.	3.827	-	33 103	126 670
Flue gas	486.0	1 646	1 578	-767 153

It was found that 255 MW are necessary in this case, from which approximately 128 MW (50%) would be provided by the dry product gas at 323 K. The energy requirement is only

6.59% higher than for NG. Another advantage of product gas drying is to diminish the gas flow rate since less water is injected into the indurating machine. As can be seen from Table 42, gas fuel increases 321%; total input increases 2.5% and flue gas increases only 1.4%.

Table 42: Air, gas, solid and flue gas flow rates in several units, for dried product gas.

Phase	kg/s	kmol/s	Nm^3/s	Nm^3/min	Increase [%]
Air	472.2	16.48	367.8	22 067	0.00%
Gas	9.97	0.555	12.12	727	321%
Solid	3.827	0.425	-	-	-
Flue gas	486.0	16.94	375.7	22 540	1.36%
TOTAL IN	-	-	379.9	22 794	2.49%

On the other hand, hot gas efficiency decreases to 70.8% if the sensible heat is not recovered. The wood input would be 11.6 kg s^{-1} (41.7 t h^{-1}), corresponding to approximately 26,500 ha of planted area. Apparently, the sensible heat losses outweigh the LHV gain in this case. Another disadvantage is that the cooling process may cause tars to condense on the heat transfer surfaces and gas cleaning steps would possibly be required.

5.7 Bio-SNG Production

In Bio-SNG production, the product gas from steam gasification is filtered, cooled and further cleaned before methanation. Gasification parameters are the same as those for steam gasification, and the gas is considered to be completely free of tars after cleaning. The methanation reaction is performed in a fixed or FB pressure vessel. For simplification, methanation was considered to be performed in a single stage at 673 K and 20 atm. Methanation reactions were assumed to reach chemical equilibrium. Finally, the gas is cooled to 298 K and CO_2 is removed with 98% efficiency. During this process, 1% methane was considered to be lost (van der Meijden et al. 2010). The process summary is shown in Figure 27.

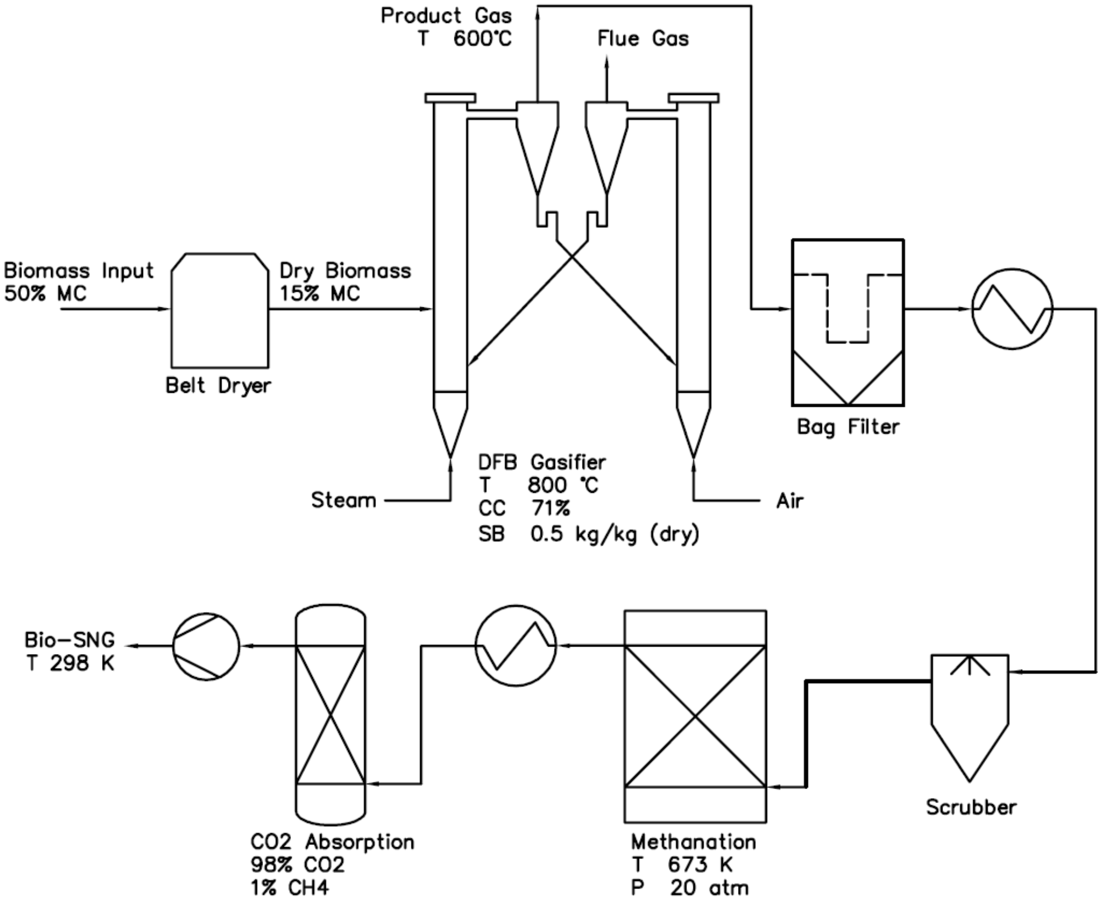


Figure 27: Bio-SNG production flow chart.

The methanation reactor is considered to be fed with product gas obtained from steam gasification (Table 35), without any water removal. Excess water is important in order to promote the water-gas shift reaction (R9 in Table 6) and achieve adequate H_2 to CO ratio. It is also known that excess moisture can avoid carbon formation and consequent catalyst poisoning. Nonetheless, the water content in the product gas (33.5%-mol) was considered to be enough to avoid soot inside the reactor (Deurwaarder et al. 2005) and therefore no more water is added.

In order to evaluate the gas composition after methanation, equilibrium constants were calculated according to Equation 36. Results of enthalpy change, entropy change, Gibbs free energy change and equilibrium constants at 673 K and 1 atm for reactions R9, R11 and R12 are shown in Table 43.

Table 43: Enthalpy, entropy and Gibbs free energy change at 673 K and 1 atm.

Reaction	$\Delta_r H^0$ kJ kmol^{-1}	$\Delta_r S^0$ kJ kmol^{-1} T^{-1}	$\Delta_r G^0$ kJ kmol^{-1}	k -
R9	-38 160	-35.85	-14 031	12.28
R11	-219 944	-245.7	-54 584	17 246
R12	-181 784	-209.8	-40 554	1 405

The number of moles of each compound when the reaction starts and when equilibrium is reached in terms of *x*, *y* and *z* are shown in Table 44. Substituting these expressions in Equations 45, 47 and 49, a nonlinear system can be solved for three equations and three variables. The system was solved utilizing Excel's tool Solver. Initial guesses for *x*, *y* and *z* were 0.01, 0.1 and 0.001 and final results were 3.199E-02, 1.072E-01 and 0.000E-00, accordingly. Apparently, reaction R12 does not occur under such conditions.

Table 44: Methanation reactor input and equilibrium composition at 673 K and 20 atm.

Component	Inlet kmol/kmol$_{gas}$	Equilibrium kmol/kmol$_{gas}$	Equilibrium kmol/kmol$_{gas}$	Outlet %-mol (wet)
CH_4	0.073	$0.073 + y - z$	0.180	23.0%
CO	0.140	$0.140 - x - y$	0.0007	0.09%
CO_2	0.141	$0.141 + x - z$	0.173	22.0%
H_2	0.311	$0.311 + x - 3y - 4z$	0.021	2.66%
N_2	0.0004	0.0004	0.0004	0.05%
H_2O	0.335	$0.335 - x + y + 2z$	0.410	52.2%
TOTAL	**1.000**	**$1.00 - 2y - 2z$**	**0.786**	**100%**

Methanation outlet molar composition is consistent with other data found in literature (Vitasari et al. 2011), indicating that the assumptions made were reasonable. In the next step, methanation output stream is cooled to 298 K and CO_2 is removed with a 98% efficiency absorption tower. The water fraction at this temperature is 3.13%. In addition, 1% of methane is lost in the absorption operation. Results are shown in Table 45. LHV is estimated at 31.7 MJ Nm^{-3} or 706 MJ kmol^{-1}.

Table 45: Bio-SNG composition.

Component	Moles kmol/kmol	%-mol (wet)	Mass kg/kmol	%-w (wet)	Volume Nm^3/kmol	%-vol (wet)
CH_4	0.1786	84.7%	13.60	89.2%	18.97	85.3%
CO	0.0007	0.34%	0.096	0.63%	0.077	0.34%
CO_2	0.0035	1.64%	0.724	4.75%	0.366	1.65%
H_2	0.0209	9.94%	0.200	1.31%	2.231	10.0%
N_2	0.0004	0.20%	0.057	0.37%	0.045	0.20%
H_2O	0.0066	3.13%	0.564	3.70%	0.558	2.51%
TOTAL	**0.2107**	**100%**	**15.24**	**100%**	**22.25**	**100%**

In order to compare the adiabatic flame temperature with NG, the final fuel composition was calculated in the same manner as that for NG, considering data shown in Table 14 and Table 45. Combustion and flue gas calculations were also performed in the same way as for NG. Bio-SNG was considered to enter the indurating machine at 298 K, and therefore no sensible heat is present. The adiabatic flame temperature for 112 MW of bio-SNG was found to be approximately the same as for NG (1,373 °C). The total energy consumption is also the same, as well as the percentages provided by solid and gaseous fuels. Table 46 shows the results for fuel composition and stoichiometric oxygen demand per GJ of fuel for bio-SNG. LHV for the fuel blend was estimated at 39.3 MJ Nm^{-3} or 489 MJ $kmol^{-1}$.

Table 46: Fuel composition and stoichiometric oxygen demand for bio-SNG.

Element	Gas $kmol/GJ_{gas}$	Solid $kmol/GJ_{solid}$	Gas + Solid $kmol/GJ_{fuel}$	Gas + Solid %-mol	Stoch. O_2 $kmol/GJ_{fuel}$
C	1.229	1.879	1.574	32.8%	1.574
H	5.086	1.125	2.988	62.3%	0.747
O	0.051	0.128	0.092	1.92%	-0.046
N	0.006	0.029	0.018	0.38%	0.000
S	0.000	0.040	0.021	0.45%	0.021
H2O	0.044	0.152	0.102	2.12%	0.000
TOTAL	**-**	**-**	**4.795**	**100%**	**2.296**

Flue gas flow rate was calculated based on the same assumptions and equations utilized before. The results are shown in Table 47. In this case, flue gas molar mass was estimated at 28.67 kg kmol^{-1} and molar volume was 22.21 Nm^3 kmol^{-1}.

Table 47: Flue gas generation per GJ of fuel for bio-SNG.

Species	Flue Gas kmol/GJ$_{fuel}$	%-mol (wet)	Mass kg/GJ$_{fuel}$	%-w (wet)	Volume Nm^3/GJ$_{fuel}$	%-vol (wet)
CO_2	1.574	2.25%	69.25	3.46%	35.03	2.26%
N_2	53.44	76.5%	1 497	74.8%	1 197	77.2%
O_2	11.91	17.1%	381.0	19.0%	266.6	17.2%
SO_2	0.021	0.03%	1.370	0.07%	0.468	0.03%
H_2O	2.889	4.14%	52.06	2.60%	51.54	3.32%
TOTAL	**69.83**	**100%**	**2 001**	**100%**	**1 551**	**100%**

As temperatures and composition of all gaseous streams were known, the specific enthalpies can be calculated by means of the Shomate Equation. Final specific enthalpy results are shown in Appendix B. Product gas and flue gas mass flow rates were obtained by solving mass and energy balances (Equations 76 and 78) simultaneously. The results are shown in Table 48.

Table 48: Energy balance for bio-SNG.

Phase	Flow kg/s	Temperature [K]	Enthalpy kJ/kg	Heat kW
Air	472.2	1 273	1 085	512 403
Gas, sens.	2.429	298	0.00	0.00
Gas, comb.	2.429	-	46 309	112 487
Solid, comb.	3.827	-	33 103	126 670
Flue gas	478.5	1 646	1 571	-751 560

In this case, it was found that the energy consumption is only 0.07% higher than that for NG. Specific and volumetric flows of air, product gas and flue gas are shown in different units in Table 49, as well as their increase percentages compared to NG values (Table 21). As can be seen from the table, only 23% more fuel gas would be injected. The total gas input in preheating and firing zones is practically the same, increasing less than 0.2%. The same is observed with flue gas, of which the flow rate increases approximately 0.02%. Therefore, no impacts are expected in the burners or in the fans. The operation with bio-SNG is expected to be the same as for NG.

Table 49: Air, gas, solid and flue gas flow rates in several units, for bio-SNG.

Phase	kg/s	kmol/s	Nm^3/s	Nm^3/min	Increase [%]
Air	472.2	16.48	367.8	22 067	0.00%
Gas	2.429	0.159	3.55	213	23.1%
Solid	3.827	0.425	-	-	-
Flue gas	478.5	16.69	370.7	22 241	0.021%
TOTAL IN	-	-	371.3	22 280	0.180%

The wood input was calculated based on the wood required for steam gasification and the amount of bio-SNG to be produced. For instance, it is known that 0.2107 kmol of bio-SNG are produced per kmol of steam gasification product gas. If 0.159 kmol s^{-1} of bio-SNG is required, 0.757 kmol s^{-1} of product gas needs to be produced. This value was then utilized in Equation 79. Again, the losses were estimated at 2% of the energy content in the wood, and 12.0 kg s^{-1} (43.1 t h^{-1}) of wood chips with 15% MC would be utilized. This corresponds to approximately 27,500 ha of planted area, and a total area of 55,000 ha. It is important to note that the required steam gasifier size in this case is 131 MW.

The steam gasification energy input and output are 187 MW and 112 MW, respectively. Thus, the hot gas efficiency for bio-SNG production equals 60.2%. This value is quite low according to recent literature (Zwart et al. 2006, van der Meijden et al. 2010). This can be attributed to the single stage methanation at a relatively high temperature. If more stages with intercooling were added and the reactor temperature was decreased (and/or pressure increased), the net efficiency for bio-SNG could possibly reach 70%. However, typical values were utilized for e.g. single stage FB methanation (Kopyscinski et al. 2010).

5.8 Substitution Scenarios

Total substitution of NG for renewable energy is not possible in the short or medium term. Therefore, calculations were performed for lower amounts of product gas/bio-SNG in combination with NG and solid fuel. As done before, a final fuel composition was obtained based on the weighted average of each fuel composition. In the same manner, temperature was kept constant and equal to the adiabatic flame temperature for NG and solid. The most important parameters for this comparison are product gas, flue gas flow rate and total inlet flow rates. In addition, the final energy consumption and percentages of each fuel were analyzed. The amount of wood and required planted area are also relevant in order to support short term decisions.

The first substitution scenario considers only 3 MW of product gas or bio-SNG, including sensible heat (when applicable). This corresponds to 2.67% less NG being consumed, and would substitute the energy input in one burner. The results for the 3 MW substitution rate are shown in Table 50. As can be seen from the table, not many changes are expected in this case, from both operational and environmental perspectives.

Table 50: Main parameters for 3 MW substitution.

Parameter	Air	Oxygen	Steam	Steam (dry)	Bio-SNG
NG flow [Nm^3/min]	170	169	169	169	168
Product gas [Nm^3/min]	29	19.6	19.3	17.0	5.7
Increase [%]	15%	9%	9%	8%	0.6%
Gas Inlet [Nm^3/min]	22 266	22 256	22 256	22 253	22 241
Increase [%]	0.12%	0.07%	0.07%	0.06%	0.00%
Flue gas flow [Nm^3/min]	22 258	22 246	22 247	22 244	22 237
Increase [%]	0.09%	0.04%	0.05%	0.03%	0.00%
Energy consump. [MW]	239.9	239.5	239.5	239.4	239.0
Increase [%]	0.36%	0.19%	0.22%	0.15%	0.00%
Wood input [kg s^{-1}]	0.23	0.22	0.24	0.27	0.32
Planted area [ha]	517	499	561	623	733

The second substitution scenario considers 12 MW of product gas or bio-SNG, including sensible heat. This corresponds to 10.7% less NG being consumed. The results are shown in Table 51. Again, not many operational and environmental impacts are expected in this scenario.

Table 51: Main parameters for 12 MW substitution.

Parameter	Air	Oxygen	Steam	Steam (dry)	Bio-SNG
NG flow [Nm^3/min]	160	157	158	157	154
Product gas [Nm^3/min]	116	78.2	77.4	68.1	22.7
Increase [%]	60%	36%	36%	30%	2.5%
Gas Inlet [Nm^3/min]	22 343	22 303	22 302	22 292	22 244
Increase [%]	0.46%	0.28%	0.28%	0.23%	0.02%
Flue gas flow [Nm^3/min]	22 320	22 274	22 278	22 265	22 237
Increase [%]	0.38%	0.17%	0.19%	0.13%	0.00%
Energy consump. [MW]	242.4	240.8	241.1	240.5	239.0
Increase [%]	1.43%	0.76%	0.88%	0.62%	0.01%
Wood input [kg s^{-1}]	0.90	0.87	0.98	1.09	1.28
Planted area [ha]	2 067	1 996	2 243	2 492	2 931

The third substitution scenario considers 24 MW of product gas or bio-SNG, which is already a medium scale facility. This would correspond to 21.4% in NG savings. As can be seen from the results shown in Table 52, impacts in the indurating machine are more relevant; for instance, in air gasification the fuel gas volumetric flow rate already increases 119%. In the same manner, GHG reduction is expected to be significant.

Table 52: Main parameters for 24 MW substitution.

Parameter	Air	Oxygen	Steam	Steam (dry)	Bio-SNG
NG flow [Nm^3/min]	146	141	142	140	136
Product gas [Nm^3/min]	232	156.4	154.7	136.2	45.4
Increase [%]	119%	72%	72%	60%	4.9%
Gas Inlet [Nm^3/min]	22 446	22 365	22 364	22 344	22 249
Increase [%]	0.92%	0.56%	0.56%	0.47%	0.04%
Flue gas flow [Nm^3/min]	22 404	22 311	22 319	22 293	22 238

Increase [%]	0.75%	0.33%	0.37%	0.26%	0.00%
Energy consump. [MW]	245.8	242.6	243.2	242.0	239.0
Increase [%]	2.85%	1.52%	1.77%	1.23%	0.01%
Wood input [kg s^{-1}]	1.80	1.74	1.95	2.17	2.55
Planted area [ha]	4 134	3 991	4 486	4 983	5 861

The last substitution scenario considers 60 MW gasification plants, which would already be within the world's largest facilities. This corresponds to 53.4% in NG savings. The results are shown in Table 53.

Table 53: Main parameters for 60 MW substitution.

Parameter	Air	Oxygen	Steam	Steam (dry)	Bio-SNG
NG flow [Nm^3/min]	107	94	97	92	81
Product gas [Nm^3/min]	580	391.0	386.8	340.6	113.5
Increase [%]	298%	181%	180%	150%	12.3%
Gas Inlet [Nm^3/min]	22 754	22 553	22 551	22 500	22 261
Increase [%]	2.31%	1.41%	1.40%	1.17%	0.10%
Flue gas [Nm^3/min]	22 656	22 422	22 443	22 379	22 239
Increase [%]	1.88%	0.83%	0.93%	0.64%	0.01%
Energy consump. [MW]	256.0	248.1	249.5	246.4	239.1
Increase [%]	7.13%	3.79%	4.41%	3.09%	0.04%
Wood input [kg s^{-1}]	4.50	4.35	4.89	5.43	6.38
Planted area [ha]	10 336	9 978	11 215	12 458	14 653

Further studies also need to be performed in order to assess in which scenarios the burners would require substitution. Changing the burner design and material would be required due to higher temperature and high gas fuel flow rate. However, in case of low NG substitution, the product gas could be injected somewhere else, such as in the kiln hood. Alternative locations for product gas injection and how much it can be injected at a particular place in order to avoid premature combustion also need further analysis.

6 ECONOMICAL ASSESSMENT

In this chapter, the economic assessment of different gasification alternatives is provided, under different NG substitution scenarios. Air-blown CFB gasification and utilization of low energy content gas is discussed first, as this is the simplest and lowest cost large-scale option. Oxygen/steam gasification is presented as another possibility. Finally, DFB steam gasification and conversion of product gas into bio-SNG is taken as the choice with least impact on the system since fuel gas with similar properties compared to NG can be obtained. Detailed results of TD&IC calculations, for all technologies can be found in Appendix D.

The gasification plant's economic life cycle was considered as 20 years in all cases. The interest rate was preliminary considered as the Brazilian interest rate target (SELIC) for 2013, which was 9.5% in November. As a result, the AF in all cases is 11.3%. In order to be more conservative, 8,000 operation hours per year were assumed. The biomass price in Brazil was assumed as 15 € MWh^{-1} and the plant electricity consumption were considered to be 30 kW$_e$ per MWh of gas produced, except for oxygen gasification, in which the electricity consumption of ASU needs to be added, resulting in a total of 60 kW$_e$ MWh^{-1}. In addition, annual O&M costs were considered to be 2% of the investment costs in € MWh^{-1}. Costs with ash disposal and steam production, among others, were included in variable O&M, which was assumed to be 65% of the annual O&M costs. The exchange rate was assumed as 3.00 BRL/EUR. Price of NG (before correction) and electricity in Brazil were considered to be 40 € MWh^{-1} (MME 2013) and 80 € MWh^{-1} (ANEEL 2013) respectively.

6.1 CFB Air Gasification

The CFB air gasification plant is the simplest and therefore the least expensive among all options. In order to calculate the total fuel costs, the hot gas efficiency calculated in Chapter 05 was utilized, which is 85.4%. As can be seen from the results (Table 54), 3 MW and 12 MW gasification plants are not economically feasible since their total costs exceed actual NG costs. For all other sizes the project can be considered feasible; especially for the 157 MW plant, for which the payback is estimated at less than 6 years.

Table 54: Economic analysis for CFB air gasification plant.

Parameter	Unit	Plant Size [MW]				
		3	12	24	60	157
TD&IC	M€	2.21	5.89	9.50	18.1	35.5
Capital costs	€/MWh	10.45	6.96	5.61	4.29	3.21
Fuel costs	€/MWh	17.56	17.56	17.56	17.56	17.56
Electricity	MWh_e/a	720	2 880	5 760	14 400	37 680
Electricity costs	k€	58	230	461	1 152	3 014
Electricity costs	€/MWh	2.40	2.40	2.40	2.40	2.40
Annual O&M	€/MWh	1.84	1.23	0.99	0.76	0.57
Variable O&M	€/MWh	3.42	2.28	1.84	1.40	1.05
O&M Cost	€/MWh	5.26	3.51	2.83	2.16	1.62
TOTAL	€/MWh	35.68	30.43	28.41	26.41	24.79
NG saving	MW	1.82	8.26	16.8	42.6	112
Real NG cost	€/MWh	24.27	27.53	28.00	28.40	28.54
Difference	€/MWh	-11.41	-2.90	-0.41	1.99	3.75
Payback period	Years	-	-	26	9.4	5.4
Max. IRR	%	-	-	8.45%	15.7%	24.3%
Max. Fuel cost	€/MWh	-	-	14.65	16.70	18.20

6.2 CFB Oxygen Gasification

The CFB oxygen gasification plant is one alternative to increase the LHV of the product gas by avoiding excessive dilution by nitrogen. In order to calculate the total fuel costs, a hot gas efficiency of 88.4% was assumed. As can be seen from the results (Table 55), 3 MW and 12 MW gasification plants are not economically feasible. For all other sizes the project can be considered economical; especially for the 132 MW plant, for which the payback is also estimated at less than 6 years.

Table 55: Economic analysis for CFB oxygen gasification plant.

Parameter	Unit	Plant Size [MW]				
		3	**12**	**24**	**60**	**132**
TD&IC	M€	2.99	7.94	12.9	24.5	42.5
Capital costs	€/MWh	14.14	9.39	7.59	5.79	4.57
Fuel costs	€/MWh	16.97	16.97	16.97	16.97	16.97
Electricity	MWh_e/a	1 440	5 760	11 520	28 800	63 360
Electricity costs	k€	115	461	922	2 304	5 069
Electricity costs	€/MWh	4.80	4.80	4.80	4.80	4.80
Annual O&M	€/MWh	2.49	1.65	1.34	1.02	0.80
Variable O&M	€/MWh	4.63	3.07	2.49	1.90	1.49
O&M Cost	€/MWh	7.12	4.73	3.82	2.92	2.30
TOTAL	€/MWh	43.02	35.88	33.19	30.48	28.64
NG saving	MW	2.22	9.86	20.0	50.6	112
Real NG cost	€/MWh	29.60	32.87	33.33	33.73	33.94
Difference	€/MWh	-13.42	-3.01	0.15	3.26	5.30
Payback period	Years	-	-	19	8.5	5.4
Max. IRR	%	-	-	9.78%	17.0%	24.2%
Max. Fuel cost	€/MWh	-	-	15.13	17.88	19.69

6.3 DFB Steam Gasification

The DFB steam gasification plant is another alternative to avoid excessive product gas dilution and consequently increase the LHV, especially if the product gas can be dried. Additionally, this system is more flexible since it is ideal for bio-SNG production. In order to calculate the total fuel costs, the hot gas efficiency calculated previously was utilized, which is 78.7% for wet product gas and 70.8% for dry product gas. As can be seen from the results (Table 56 and Table 57), 3 to 24 MW gasification plants are not economically feasible.

Table 56: Economic analysis for DFB steam gasification plant, wet product gas.

Parameter	Unit	Plant Size [MW]				
		3	12	24	60	136
TD&IC	M€	3.19	8.46	13.8	26.13	46.3
Capital costs	€/MWh	15.08	10.00	8.15	6.18	4.83
Fuel costs	€/MWh	19.06	19.06	19.06	19.06	19.06
Electricity	MWh$_e$/a	720	2 880	5 760	14 400	32 640
Electricity costs	k€	58	230	461	1 152	2 611
Electricity costs	€/MWh	2.40	2.40	2.40	2.40	2.40
Annual O&M	€/MWh	2.66	1.76	1.44	1.09	0.85
Variable O&M	€/MWh	4.94	3.27	2.67	2.02	1.58
O&M Cost	€/MWh	7.60	5.04	4.10	3.11	2.43
TOTAL	€/MWh	44.14	36.50	33.71	30.75	28.72
NG saving	MW	2.14	9.56	19.4	49.1	112
Real NG cost	€/MWh	28.53	31.87	32.33	32.73	32.94
Difference	€/MWh	-15.60	-4.63	-1.38	1.99	4.22
Payback period	Years	-	-	-	11	6.5
Max. IRR	%	-	-	-	13.9%	20.8%
Max. Fuel cost	€/MWh	-	-	-	16.56	18.32

Table 57: Economic analysis for DFB steam gasification plant, dry product gas.

Parameter	Unit	Plant Size [MW]				
		3	12	24	60	128
TD&IC	M€	3.462	9.17	14.9	28.29	48.03
Capital costs	€/MWh	16.37	10.84	8.78	6.69	5.32
Fuel costs	€/MWh	21.19	21.19	21.19	21.19	21.19
Electricity	MWh$_e$/a	720	2 880	5 760	14 400	30 720
Electricity costs	k€	58	230	461	1 152	2 458
Electricity costs	€/MWh	2.40	2.40	2.40	2.40	2.40
Annual O&M	€/MWh	2.89	1.91	1.55	1.18	0.94
Variable O&M	€/MWh	5.36	3.55	2.87	2.19	1.74
O&M Cost	€/MWh	8.24	5.46	4.42	3.37	2.68

TOTAL	€/MWh	48.20	39.88	36.79	33.64	31.59
NG saving	MW	2.3	10.2	20.7	52.3	112
Real NG cost	€/MWh	30.67	34.00	34.50	34.87	35.00
Difference	€/MWh	-17.53	-5.88	-2.29	1.22	3.41
Payback period	Years	-	-	-	14	7.9
Max. IRR	%	-	-	-	12.0%	17.9%
Max. Fuel cost	€/MWh	-	-	-	15.87	17.41

6.4 Bio-SNG Production

Producing bio-SNG is by far the most complex and expensive alternative. On the other hand, the fuel has similar properties in comparison to NG, and no adaptions in the indurating machine would be required (e.g. burners and fans). In this case, efficiency was considered to be 60.2%. As shown in Table 58, none of the alternatives for bio-SNG are economically feasible, given the current NG price.

Table 58: Economic analysis for bio-SNG plant.

Parameter	**Unit**	**Plant Size [MW]**				
		3	**12**	**24**	**60**	**131**
TD&IC	M€	10.91	28.74	46.7	88.7	153.2
Capital costs	€/MWh	51.58	33.97	27.60	20.97	16.59
Fuel costs	€/MWh	24.92	24.92	24.92	24.92	24.92
Electricity	MWh$_e$/a	720	2 880	5 760	14 400	31 440
Electricity costs	k€/a	58	230	461	1 152	2 515
Electricity costs	€/MWh	2.40	2.40	2.40	2.40	2.40
Annual O&M	€/MWh	9.09	5.99	4.86	3.70	2.92
Variable O&M	€/MWh	16.88	11.12	9.03	6.86	5.43
O&M Cost	€/MWh	25.98	17.11	13.90	10.56	8.35
TOTAL	€/MWh	104.88	78.40	68.82	58.85	52.26
NG saving	MW	2.67	11.6	23.6	59.6	112
Real NG cost	€/MWh	35.60	38.67	39.33	39.73	40.00
Difference	€/MWh	-69.28	-39.73	-29.48	-19.11	-12.26

7 SUMMARY

This study assessed the use of several gasification technologies for NG substitution in a travelling grate induration furnace. CFB air and oxygen gasification and DFB steam gasification and bio-SNG production were discussed as the most promising large-scale models. The main findings are summarized in this chapter.

First, product gas/bio-SNG compositions for each technology were calculated, based on mass balances. Thus, product gases compositions in molar percentage (wet basis) as well as main thermochemical and combustion properties are summarized in Table 59.

Table 59: Product gas composition and LHV for several gasification technologies.

Property	Air	Oxygen	Steam	Steam (dry)	Bio-SNG
CH_4 [%-mol]	4.14%	0.76%	7.32%	9.66%	84.7%
CO [%-mol]	18.0%	35.7%	14.0%	18.5%	0.34%
CO_2 [%-mol]	11.5%	10.5%	14.1%	18.6%	1.64%
H_2 [%-mol]	13.2%	27.9%	31.1%	41.0%	9.94%
N_2 [%-mol]	41.4%	0.85%	0.04%	0.06%	0.20%
H_2O [%-mol]	11.8%	24.2%	33.5%	12.2%	3.13%
LHV [MJ/Nm^3]	5.32	8.22	8.32	10.5	31.7
MM [kg/kmol]	24.75	19.92	17.98	17.96	15.24
$\bar{V}$ [Nm^3/kmol]	21.85	21.30	20.87	21.84	22.25
Combustion air ratio	6.21	6.85	6.39	6.26	6.19

The most important comparison parameters between different gasification technologies were found to be adiabatic flame temperature for 112 MW of gaseous fuel, the kiln’s final energy consumption, required gasifier size and increase in the gas flow. Wood input and estimated planted area are also important decision making parameters. These values are summarized in Table 60.

Short and medium term substitution scenarios were proposed; for instance 3, 12, 24 and 60 MW. As expected, increase in energy consumption, gas flow rates and other key parameters were found to be a linear function of the gasifier size. Therefore, small gasification facilities are expected to have no or very low impact on the system.

Table 60: Comparison of gasification technologies for total substitution of NG.

Parameter	Air	Oxygen	Steam (wet)	Steam (dry)	Bio-SNG
Flame temperature for 112 MW [°C]	1 347	1 365	1 361	1 354	1 373
Energy consumption [MW]	284	259	263	255	239
Increase in energy consumption [%]	18.6%	8.37%	10.0%	6.59%	0.07%
Gasifier Size [MW]	157	132	136	128	131
Gaseous fuel flow increase [%]	778%	399%	408%	321%	23.1%
Kiln hood gas flow increase [%]	6.05%	3.10%	3.17%	2.49%	0.18%
Flue gas increase [%]	4.93%	1.83%	2.10%	1.36%	0.021%
Hot gas efficiency [%]	85.4%	88.4%	78.7%	70.8%	60.2%
Wood input at 15% MC [$t\ h^{-1}$]	42.4	34.5	40.0	41.7	43.1
Planted area [ha]	27 000	22 000	25 500	26 500	27 500

The main data obtained in the economical assessment were investment costs, payback period, maximum interest rate and maximum wood price for which each technology is still economically feasible. Except for the TD&IC, all other values were calculated based on a corrected NG price since the energy consumption increases differently in each case and size. The results are summarized in Figures 29 to 32.

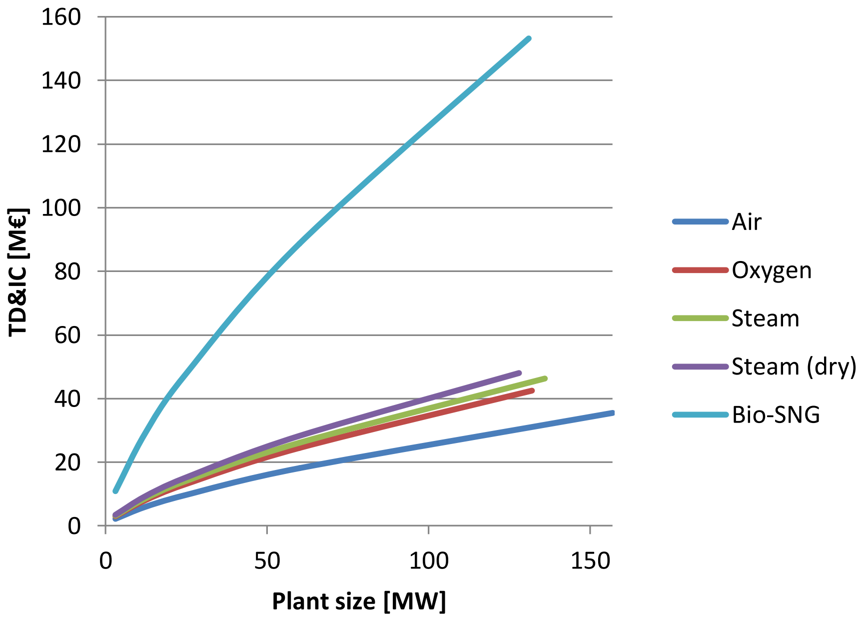


Figure 28: TD&IC as a function of gasification plant size.

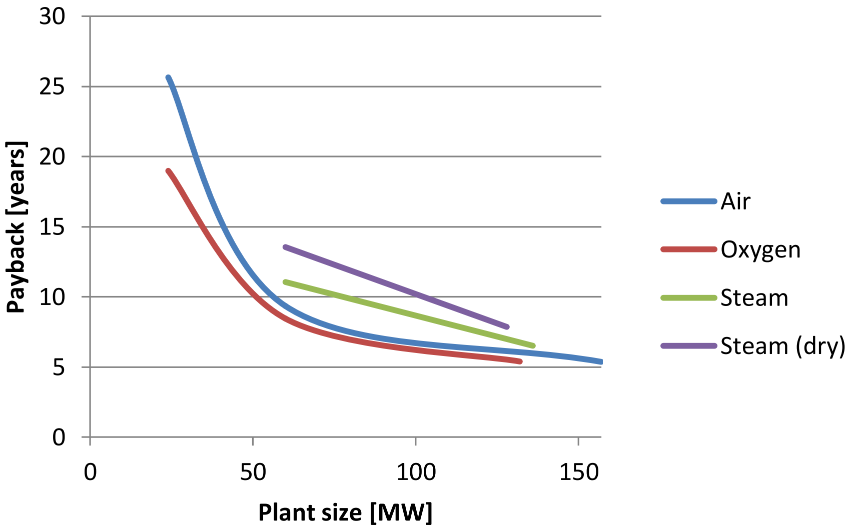


Figure 29: Payback period as a function of gasification plant size, for IR 9.5% and wood price of 15 €/MWh.

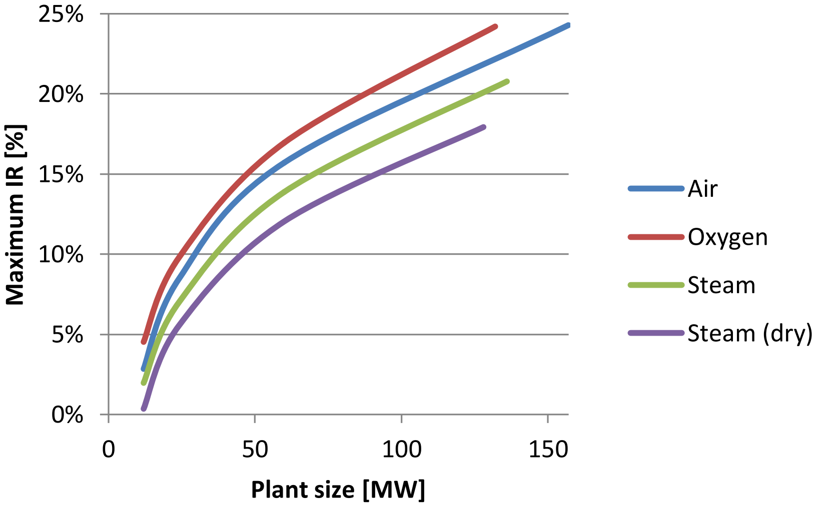


Figure 30: Maximum interest rate as a function of gasification plant size, for 20 years lifecycle and wood price of 15 €/MWh.

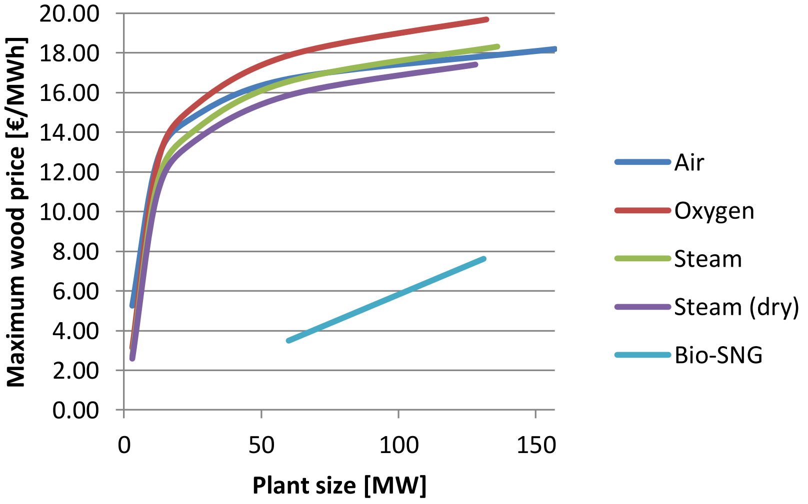


Figure 31: Maximum wood price as a function of gasification plant size, for IR 9.5% and 20 years lifecycle.

Finally, it is important to address that each technology discussed in this study has unique characteristics, advantages and disadvantages which must be taken into account when making the final decision. The most important advantages and disadvantages are summarized in Table 61.

Table 61: Advantages and disadvantages of each gasification technology.

Gasifier type	Advantages	Disadvantages
CFB Air Gasification	Simplest and lowest cost among all studied models; great potential for reducing costs compared to NG; well developed technology for large-scale.	Low energy content, resulting large amounts of product gas injected; highest impact in the kiln; require burners' substitution.
CFB Oxygen Gasification	Higher product gas LHV compared to air; highest thermal efficiency; less wood input required compared to all other cases.	Higher costs compared to air gasification; require burner's substitution; need for ASU; not well developed for biomass gasification.
DFB Steam Gasification (w/o drying)	Higher product gas LHV than air and approx. same LHV as oxygen; high hydrogen content in the product gas; more flexible system.	Higher costs compare to CFB; require burner's substitution; not available in commercial scale yet.
DFB Steam Gasification (with drying)	Higher product gas LHV and lower impact in the system compared to others (except bio-SNG).	Higher costs compare to CFB; require burner's substitution; not available in commercial scale yet; tar condensation in the heat exchanger.
Bio-SNG Production	No impact in the kiln's operation; do not require burner's substitution.	Not economically feasible in any plant size.

8 CONCLUSIONS & FUTURE WORK

The present work discussed the use of product gas/bio-SNG as a renewable substitute for NG in iron ore pelletizing processes. An analysis of the Brazilian NG market has shown that the country is not yet self-sufficient in NG, and that more than half of imported NG comes from the same supplier, namely Bolivia. The supply agreement between these two countries is valid until 2019, and there are uncertainties concerning the amount and price of NG supply after this period. In addition, sustainability and global warming issues contribute even more to the necessity of investments in renewable energy.

Biomass gasification has proven to be one of the most promising technologies for NG replacement. Product gases can be obtained from different gasification technologies assessed in this study. Currently, the most relevant large-scale biomass gasification model is the CFB. It cannot only handle a wide range of alternative fuels but also different gasification media, such as air, oxygen and steam. Each gasifying medium leads to different product gas characteristics, such as composition and LHV. Bio-SNG can be best obtained by DFB steam gasification followed by methanation.

The aforementioned thermochemical processes were compared under technical and economic perspectives. First, adiabatic flame temperature for the current fuels (NG, petcoke and coal) was determined. The amount of product gas and bio-SNG required to maintain this constant temperature was calculated in each case, as well as the gas flow through the indurating machine. The air flow was kept constant since it is defined by the requirements in the cooling zone. It was found that, for all gasification processes except bio-SNG production, the specific energy consumption would increase as well as the gas flow rate through the kiln.

Burner substitution seems inevitable in all cases, except for bio-SNG. On the other hand, it is possible that the fans would not require substitution in any of the cases since the maximum increase in the flue gas flow is approximately 5%. Lower substitution scenarios were also evaluated, such as 3, 12, 24 and 60 MW. As expected, the lower the substitution the less impact product gas would have on the system. However, reductions in GHG emissions may not be relevant, and the plant costs are considerably higher.

Preliminary economic assessment has shown that small-scale gasification plants (e.g. up to 24 MW) are not feasible. On the other hand, from 60 MW the use of product gas appears to be slightly less costly than NG in all cases except bio-SNG production. However, adaptations in the induration furnace were not taken into account in the economic assessment, and it is possible that this enterprise would be less attractive. Nonetheless, sustainability, environmental and security of supply aspects should be also considered in this investment.

Based on the results of this study, it is possible to conclude that air gasification is a suitable technology for NG substitution in iron ore pelletizing traveling grates. However, further studies should be performed before making the final decision:

- Impacts of increasing gas flow rate on the burners should be evaluated by computational fluid dynamics (CFD) analysis. In addition, new burner design and temperature resistant materials must be specified;
- One possible solution for the burners is to inject the gas partially or entirely (depending on the substitution rate) in another location, e.g. in the kiln hood. Further studies are needed to confirm this possibility; for example, determine the maximum amount of gas to avoid premature combustion.
- The economic assessment presented here covers only the order of magnitude of investment costs. Although the numbers are in agreement with recent budgets, all the numbers do not necessarily represent the Brazilian reality;
- Process economics may be improved if an optimized waste heat recovery solution for steam/electricity production is considered, especially in the case of bio-SNG. However, this would require a detailed project;
- Effects on pelletizing gaseous emissions, such as ammonia and nitrogen oxides. Possible costs of air pollution control equipment (if needed) should be taken into account in the economic assessment.

In addition, other studies should be also done in order to improve process economics and reinforce the necessity/advantages of investments in renewable energy, such as:

- Assess the best eucalyptus harvesting technology and optimize biomass costs;
- Evaluate the availability of low-cost alternative fuels, such as energy crops and/or agricultural;
- Life cycle analysis would help other benefits of a biomass gasification plant to be clearly defined and assessed.

REFERENCES

Economic Indicators. Chemical Engineering, 2013. vol. 120, no. 07, p. 60.

ABU EL-RUB, Z., BRAMER, E. and BREM, G., 2004. Review of Catalysts for Tar Elimination in Biomass Gasification Processes. *Industrial & Engineering Chemistry Research*, vol. 43, no. 22, pp. 6911-6919.

ANEEL., n.d. *Atlas De Energia Elétrica do Brasil.* Agência Nacional de Energia Elétrica [viewed 08 July 2013]. Available from: http://www.aneel.gov.br/arquivos/PDF/atlas_par3_cap6.pdf

ANEEL., 2013. *Tarifa Média Por Classe De Consumo e Por Região.* Agência Nacional de Energia Elétrica [viewed 04 November 2013]. Available from: http://relatorios.aneel.gov.br/_layouts/xlviewer.aspx?id=/RelatoriosSAS/RelSampRegCC.xlsx&Source=http://relatorios.aneel.gov.br/RelatoriosSAS/Forms/AllItems.aspx&DefaultItemOpen=1.

BACOVSKY, D., DALLOS, M., WÖRGETTER, M. and TASK, I.B., 2010. Status of 2nd Generation Biofuels Demonstration Facilities in June 2010. *IEA Bioenergy Task 39: Commercializing 1st and 2nd Generation Liquid Biofuels from Biomass*.

BARATIERI, M., BAGGIO, P., FIORI, L. and GRIGIANTE, M., 2008. Biomass as an Energy Source: Thermodynamic Constraints on the Performance of the Conversion Process. *Bioresource Technology*, vol. 99, no. 15, pp. 7063-7073.

BASU, P., 2010. *Biomass Gasification and Pyrolysis: Practical Design and Theory.* Academic press.

BASU, P., 2006. Solid Handling Systems for Fluidized Beds. *In: Combustion and gasification in fluidized beds,* CRC press, pp. 337-357.

BRAMMER, J.G. and BRIDGWATER, A.V., 1999. Drying Technologies for an Integrated Gasification Bio-Energy Plant. *Renewable and Sustainable Energy Reviews*, 12, vol. 3, no. 4, pp. 243-289 ISSN 1364-0321.

BREITHOLTZ, C., 2001. *The Biomass Gasification Plant to Vaskiluodon Voima in Vaasa and Two More Ongoing Gasification Projects.* Presentation given in Solander Symposium

2011 [viewed 14 November 2013]. Available from: http://www.zelenaenergija.org/blobs/eb16f372-fa48-42fe-b621-e9cf0b5fdc7a.pdf.

BRIDGWATER, A.V., 2002. *The Future for Biomass Pyrolysis and Gasification: Status, Opportunities and Policies for Europe*. Birmingham: University of Aston.

BRIDGWATER, A., 2003. Renewable Fuels and Chemicals by Thermal Processing of Biomass. *Chemical Engineering Journal*, vol. 91, no. 2, pp. 87-102.

CASTRO, L.F.A., FIGUEIRA, R.M. and TAVARES, R.P., 2004. *Princípios Básicos is Processos De Fabricação do Gusa Ao Aço Líquido.* Belo Horizonte: Universidade Federal de Minas Gerais.

CHANNIWALA, S. and PARIKH, P., 2002. A Unified Correlation for Estimating HHV of Solid, Liquid and Gaseous Fuels. *Fuel*, vol. 81, no. 8, pp. 1051-1063.

CHASE JR, M.W., 1998. NIST-JANAF Thermochemical Tables. *J. Phys. Chem. Ref. Data*. Fourth ed., pp. 1-1951.

CHEN, G., ANDRIES, J. and SPLIETHOFF, H., 2003. Biomass Conversion into Fuel Gas using Circulating Fluidised Bed Technology: The Concept Improvement and Modelling Discussion. *Renewable Energy*, vol. 28, no. 6, pp. 985-994.

COMPAGAS., n.d. *Manual do Consumidor Industrial.* Companhia Paranaense de Gás [viewed 18 October 2013]. Available from: http://www.compagas.com.br/index.php/web/onde_e_como_usar_o_gas_natural/gn_para_empresa/manual_do_consumidor_industrial.

CORELLA, J., TOLEDO, J.M. and MOLINA, G., 2007. A Review on Dual Fluidized-Bed Biomass Gasifiers. *Industrial & Engineering Chemistry Research*, vol. 46, no. 21, pp. 6831-6839.

CORTEZ, C.L., VELÁZQUEZ, S., COELHO, S.T. and ROBERTO, J., n.d. *Análise do Processo Produtivo do Eucalipto no Sistema "Short Rotation" Para Uso Como Combustível Em Uma Usina Termoelétrica.* Universidade de São Paulo [viewed 30 July 2013]. Available from: http://cenbio.iee.usp.br/download/publicacoes/congbioen_cortez.pdf.

CUMMER, K.R. and BROWN, R.C., 2002. Ancillary Equipment for Biomass Gasification. *Biomass and Bioenergy*, vol. 23, no. 2, pp. 113-128.

DEURWAARDER, E.P., et al., 2005. *Methanation of MILENA Product Gas for the Production of Bio-SNG.* 14th European Biomass Conference & Exhibition: Paris.

DEVI, L., PTASINSKI, K.J. and JANSSEN, F.J., 2003. A Review of the Primary Measures for Tar Elimination in Biomass Gasification Processes. *Biomass and Bioenergy*, vol. 24, no. 2, pp. 125-140.

DO CANTO, J.L., 2009. *Colheita Mecanizada De Biomassa Florestal Para Energia.* Universidade Federal de Viçosa.

DO CANTO, J.L., et al, 2011. Evaluation of a Wood Chipping System for Eucalyptus Tops for Energy. *Revista Árvore*, vol. 35, no. 6, pp. 1327-1334.

EPE., 2012. *Balanço Energético Nacional 2012 - Ano Base 2011.* Brasília: Ministério das Minas e Energia [viewed 29 March 2013]. Available from: https://ben.epe.gov.br/downloads/Resultados_Pre_BEN_2012.pdf.

EPE., 2011. *Projeção Da Demanda De Energia Elétrica Para Os Próximos 10 Anos (2012-2021).* Rio de Janeiro: Ministério das Minas e Energia [viewed 29 March 2013]. Available from: http://www.epe.gov.br/mercado/Documents/S%C3%A9rie%20Estudos%20de%20Energia/20120104_1.pdf.

FERCHER, E., et al., 1998. *Two Years Experience with the FICFB-Gasification Process.* 10th European Conference and Technology Exhibition on Biomass for Energy and Industry.

FERMOSO, J., et al, 2009. High-Pressure Co-Gasification of Coal with Biomass and Petroleum Coke. *Fuel Processing Technology*, vol. 90, no. 7, pp. 926-932.

GAO, J., et al, 2012. A Thermodynamic Analysis of Methanation Reactions of Carbon Oxides for the Production of Synthetic Natural Gas. *RSC Advances*, vol. 2, no. 6, pp. 2358-2368.

GASSNER, M. and MARÉCHAL, F., 2009. Thermo-Economic Process Model for Thermochemical Production of Synthetic Natural Gas (SNG) from Lignocellulosic Biomass. *Biomass and Bioenergy*, vol. 33, no. 11, pp. 1587-1604.

GHAFFARIYAN, M., et al, 2011. Biomass Harvesting in Eucalyptus Plantations in Western Australia. *Southern Forests: A Journal of Forest Science*, vol. 73, no. 3-4, pp. 149-154.

GÖRANSSON, K., SÖDERLIND, U., HE, J. and ZHANG, W., 2011. Review of Syngas Production Via Biomass DFBGs. *Renewable and Sustainable Energy Reviews*, vol. 15, no. 1, pp. 482-492.

GULLICHSEN, J. and FOGELHOLM, C., 1999. Preparation and Handling of Bleaching Chemicals. *In: Book 6B: Chemical pulping*, Helsinki: Fapet Oy, pp. 391-410.

HARYANTO, A., FERNANDO, S.D., PORDESIMO, L.O. and ADHIKARI, S., 2009. Upgrading of Syngas Derived from Biomass Gasification: A Thermodynamic Analysis. *Biomass and Bioenergy*, vol. 33, no. 5, pp. 882-889.

HIGMAN, C. and VAN DER BURGT, M., 2008. *Gasification. 2nd.* Gulf Professional Publishing, USA.

HOFBAUER, H. and RAUCH, R., 2000. Stoichiometric Water Consumption of Steam Gasification by the FICFB-Gasification Process. *Progress in Thermochemical Biomass Conversion*, vol. 1, pp. 199-208.

International Energy Agency, 2013. *Tracking Clean Energy Progress 2013.* Paris: IEA Input to the Clean Energy Ministerial.

JESUS, C.A.G., 2012. *Economia Mineral: Sumário Mineral 2012.* Departamento Nacional de Produção Mineral (DNPM) [viewed 20 May 2013]. Available from: https://sistemas.dnpm.gov.br/publicacao/mostra_imagem.asp?IDBancoArquivoArquivo=7390

KALLIO, M. and LEINONEN, A., 2005. *Production Technology of Forest Chips in Finland.* Jyväskylä: Technical Research Center of Finland (VTT).

KAUP, B.Z., 2010. A Neoliberal Nationalization? the Constraints on Natural-Gas-Led Development in Bolivia. *Latin American Perspectives*, vol. 37, no. 3, pp. 123-138.

KOPYSCINSKI, J., SCHILDHAUER, T.J. and BIOLLAZ, S., 2010. Production of Synthetic Natural Gas (SNG) from Coal and Dry biomass–A Technology Review from 1950 to 2009. *Fuel*, vol. 89, no. 8, pp. 1763-1783.

LI, C. and SUZUKI, K., 2009. Tar Property, Analysis, Reforming Mechanism and Model for Biomass gasification—an Overview. *Renewable and Sustainable Energy Reviews*, vol. 13, no. 3, pp. 594-604.

LV, D., et al, 2010. Effect of Cellulose, Lignin, Alkali and Alkaline Earth Metallic Species on Biomass Pyrolysis and Gasification. *Fuel Processing Technology*, vol. 91, no. 8, pp. 903-909.

MCKENDRY, P., 2002. Energy Production from Biomass (Part 3): Gasification Technologies. *Bioresource Technology*, vol. 83, no. 1, pp. 55-63.

METSO PAPER., 2010. *Rotatory Screw Reclaimers.* [viewed 05 September 2013]. Available from: http://www.metso.com/MP/marketing/Vault2MP.nsf/BYWID/WID-030328-2256C-57B5E/$File/WHDU_B_2080_013-02.pdf?OpenElement.

MILNE, T.A., ABATZOGLOU, N. and EVANS, R.J., 1998. *Biomass Gasifier "Tars": Their Nature, Formation, and Conversion.* National Renewable Energy Laboratory Golden, CO.

MME., 2013. *Boletim Mensal De Acompanhamento Da Indústria De Gás Natural.* Ministério das Minas e Energia [viewed 07 November 2013]. Available from: http://www.mme.gov.br/sproduct gas/galerias/arquivos/publicacoes/boletim_mensal_acompanhamento_industria_gas_natural/Boletim_Gas_Natural_nr_74_mai_13.pdf.

MORAN, M.J., SHAPIRO, H.N., BOETTNER, D.D. and BAILEY, M., 2006. *Fundamentals of Engineering Thermodynamics.* Fifth ed. John Wiley & Sons, Inc.

NIGRO, J.C., 1982. *Firing an Iron Ore Pelletizing Kiln with Low-Btu Gas from Lignite.* US Department of the Interior, Bureau of Mines.

OLOFSSON, I., NORDIN, A. and SÖDERLIND, U., 2005. *Initial Review and Evaluation of Process Technologies and Systems Suitable for Cost-Efficient Medium-Scale Gasification for Biomass to Liquid Fuels.* Umeå Universitet.

PALONEN, J., OY, F.W.E., ANTTIKOSKI, T. and ERIKSSON, T., 2005. *The Foster Wheeler Gasification Technology for Biofuels: Refuse-Derived Fuel (RDF) Power Generation.*

PUIG-ARNAVAT, M., BRUNO, J.C. and CORONAS, A., 2010. Review and Analysis of Biomass Gasification Models. *Renewable and Sustainable Energy Reviews*, vol. 14, no. 9, pp. 2841-2851.

QUÉNO, L.M.R., et al, 2011. Custo De Produção Das Biomassas De Eucalipto e Capim Elefante Para Energia. *Cerne, Lavras*, vol. 17, no. 3, pp. 417-426.

RAYAPROLU, K., 2010. Fluidized Bed Combustion. *In: Boilers for power and process,* CRC Press, pp. 505-558.

REILLY, B.G., 2011. *International Silo Association.* [viewed 30 October 2013]. Available from: http://www.silo.org/costs.htm.

ROCKSTRÖM, J., et al, 2009. A Safe Operating Space for Humanity. *Nature*, vol. 461, no. 7263, pp. 472-475.

SALO, K. and MOJTAHEDI, W., 1998. Fate of Alkali and Trace Metals in Biomass Gasification. *Biomass and Bioenergy*, vol. 15, no. 3, pp. 263-267 ISSN 0961-9534.

SAMARCO., n.d. *Samarco Technical Information.* [viewed 28 March 2013]. Available from: http://www.samarco.com.br/uploads/xgccbf.pdf.

SZTABERT, Z.T. and KUDRA, T., 2006. Cost-Estimation Methods for Drying. A.S. MUJUMDAR ed., *In: Handbook of industrial drying,* CRC Press.

TOWLER, G.P. and SINNOTT, R.K., 2012. Chapter 07: Capital Costs Estimating. *In: Chemical Engineering Design: Principles, Practice, and Economics of Plant and Process Design,* Second Edition ed. Access Online via Elsevier, pp. 307-354.

TURN, S.Q., KINOSHITA, C.M., ISHIMURA, D.M. and ZHOU, J., 1998. The Fate of Inorganic Constituents of Biomass in Fluidized Bed Gasification. *Fuel*, vol. 77, no. 3, pp. 135-146.

United Nations Statistics Division., 2013. *2012 International Trade Statistics Yearbook.* New York: United Nations [viewed 05 July 2013] Available from: http://comtrade.un.org/pb/CountryPagesNew.aspx?y=2012.

VAKKILAINEN, E., KUPARINEN, K. and HEINIMÖ, J., 2013. *Industrial Users of Energy Biomass.* Lappeenranta: IEA Bioenergy - Task 40.

VAN DER MEIJDEN, C., VERINGA, H. and RABOU, L., 2010. The Production of Synthetic Natural Gas (SNG): A Comparison of Three Wood Gasification Systems for Energy Balance and overall Efficiency. *Biomass and Bioenergy,* vol. 34, no. 3, pp. 302-311.

VAN LOO, S. and KOPPEJAN, J., 2008. Biomass Fuel Supply and Pre-treatment. *In: The handbook of biomass combustion and co-firing,* Earthscan, pp. 44-108.

VITASARI, C.R., JURASCIK, M. and PTASINSKI, K.J., 2011. Exergy Analysis of Biomass-to-Synthetic Natural Gas (SNG) Process Via Indirect Gasification of various Biomass Feedstock. *Energy*, vol. 36, no. 6, pp. 3825-3837.

WORLEY, M. and YALE, J., 2012. *Biomass Gasification Technology Assessment: Consolidated Report.* Golden, CO.: National Renewable Energy Laboratory (NREL).

YIP, K., TIAN, F., HAYASHI, J. and WU, H., 2009. Effect of Alkali and Alkaline Earth Metallic Species on Biochar Reactivity and Syngas Compositions during Steam Gasification. *Energy & Fuels*, vol. 24, no. 1, pp. 173-181.

ZAHL, R. and NIGRO, J., 1979. *The use of Low-Btu Gas for Iron Oxide Pellet Induration: An Interim Report.* Symposium proceedings, Environmental Aspects of Fuel Conversion Technology, IV (April 1979, Hollywood, FL)US Environmental Protection Agency, Industrial Environmental Research Laboratory.

ZWART, R., et al., 2006. *Production of Synthetic Natural Gas (SNG) from Biomass.* Energy Research Centre of the Netherlands (ECN).

ZWART, R., et al, 2009. Oil-based Gas washing—Flexible Tar Removal for high-efficient Production of Clean Heat and Power as Well as Sustainable Fuels and Chemicals. *Environmental Progress & Sustainable Energy*, vol. 28, no. 3, pp. 324-335.

APPENDIX A: Shomate Equation's Parameters

Gas	**Temp. [K]**		**A**	**B**	**C**	**D**	**E**	**F**	**G**	**H**
	Min.	**Max.**								
CH_4	298	1300	-0.70	108.5	-42.52	5.863	0.679	-76.84	158.7	-74.87
	1300	6000	85.81	11.26	-2.114	0.138	-26.42	-153.5	224.4	-74.87
CO	298	1300	25.57	6.096	4.055	-2.671	0.131	-118.0	227.4	-110.5
	1300	6000	35.15	1.300	-0.206	0.014	-3.283	-127.8	231.7	-110.5
CO_2	298	1200	25.00	55.19	-33.69	7.948	-0.137	-403.6	228.2	-393.5
	1200	6000	58.17	2.720	-0.492	0.039	-6.447	-425.9	263.6	-393.5
H_2	298	1000	33.07	-11.36	11.43	-2.773	-0.159	-9.981	172.7	0.000
	1000	2500	18.56	12.26	-2.860	0.268	1.978	-1.147	156.3	0.000
N_2	100	500	28.99	1.854	-9.647	16.635	0.000	-8.672	226.4	0.000
	500	2000	19.51	19.89	-8.599	1.370	0.528	-4.935	212.4	0.000
O_2	100	700	31.32	-20.24	57.87	-36.51	-0.007	-8.903	246.8	0.000
	700	2000	30.03	8.773	-3.988	0.788	-0.742	-11.32	236.2	0.000
SO_2	298	1200	21.43	74.35	-57.75	16.36	0.087	-305.8	254.9	-296.8
	1200	6000	57.48	1.009	-0.076	0.005	-4.045	-324.4	302.8	-296.8
H_2O	500	1700	30.09	6.833	6.793	-2.534	0.082	-250.9	223.4	-241.8
	1700	6000	41.96	8.622	-1.500	0.098	-11.16	-272.2	219.8	-241.8

APPENDIX B: Enthalpy Calculation Results

1. Air and flue gas enthalpies for NG combustion:

Phase	**Air**		**Flue Gas**	
Temperature [K]	1 273		1 646	
Component	**%-mol (wet)**	**Enthalpy kJ/mol**	**%-mol (wet)**	**Enthalpy kJ/mol**
CO_2	0.00%	48.61	2.28%	70.31
N_2	77.5%	30.58	76.6%	43.53
O_2	20.6%	32.37	17.1%	45.97
SO_2	0.00%	49.55	0.03%	70.78
H_2O	1.88%	37.73	4.00%	55.15
TOTAL	**100%**	**31.08**	**100%**	**45.03**

2. Air, product gas and flue gas enthalpies for air gasification:

Phase	**Air**		**Product gas**		**Flue Gas**	
T [K]	1 273		873		1 646	
Component	**%-mol (wet)**	**Enthalpy kJ/mol**	**%-mol (wet)**	**Enthalpy kJ/mol**	**%-mol (wet)**	**Enthalpy kJ/mol**
CH_4	0.00%	59.11	4.14%	29.40	0.00%	91.04
CO	0.00%	30.94	18.0%	17.52	0.00%	44.03
CO_2	0.00%	48.61	11.5%	26.61	3.58%	70.31
H_2	0.00%	29.07	13.2%	16.87	0.00%	41.06
N_2	77.5%	30.58	41.4%	17.36	75.7%	43.53
O_2	20.6%	32.37	0.00%	18.32	16.3%	45.97
SO_2	0.00%	49.55	0.00%	27.58	0.03%	70.78
H_2O	1.88%	37.73	11.8%	20.86	4.48%	55.15
TOTAL	**100%**	**31.08**	**100%**	**19.30**	**100%**	**45.42**

3. Air, product gas and flue gas enthalpies for oxygen gasification:

Phase	**Air**		**Product gas**		**Flue Gas**	
T [K]	1 273		873		1 646	
Component	**%-mol (wet)**	**Enthalpy kJ/mol**	**%-mol (wet)**	**Enthalpy kJ/mol**	**%-mol (wet)**	**Enthalpy kJ/mol**
CH_4	0.00%	59.11	0.76%	32.77	0.00%	91.04
CO	0.00%	30.94	35.7%	19.15	0.00%	44.03
CO_2	0.00%	48.61	10.5%	29.25	3.26%	70.31
H_2	0.00%	29.07	27.9%	18.36	0.00%	41.06
N_2	77.5%	30.58	0.85%	18.96	75.2%	43.53
O_2	20.6%	32.37	0.00%	20.03	17.1%	45.97
SO_2	0.00%	49.55	0.00%	30.26	0.03%	70.78
H_2O	1.88%	37.73	24.2%	22.86	4.48%	55.15
TOTAL	**100%**	**31.08**	**100%**	**21.00**	**100%**	**45.35**

4. Air, product gas and flue gas enthalpies for steam gasification (wet gas):

Phase	**Air**		**Product gas**		**Flue Gas**	
T [K]	1 273		873		1 646	
Component	**%-mol (wet)**	**Enthalpy kJ/mol**	**%-mol (wet)**	**Enthalpy kJ/mol**	**%-mol (wet)**	**Enthalpy kJ/mol**
CH_4	0.00%	59.11	7.32%	29.40	0.00%	91.04
CO	0.00%	30.94	14.0%	17.52	0.00%	44.03
CO_2	0.00%	48.61	14.1%	26.61	2.85%	70.31
H_2	0.00%	29.07	31.1%	16.87	0.00%	41.06
N_2	77.5%	30.58	0.04%	17.36	74.8%	43.53
O_2	20.6%	32.37	0.00%	18.32	16.8%	45.97
SO_2	0.00%	49.55	0.00%	27.58	0.03%	70.78
H_2O	1.88%	37.73	33.5%	20.86	5.59%	55.15
TOTAL	**100%**	**31.08**	**100%**	**20.59**	**100%**	**45.36**

5. Air, product gas and flue gas enthalpies for steam gasification (dry gas):

Phase	**Air**		**Product Gas**		**Flue Gas**	
T [K]	1 273		348		1 646	
Component	**%-mol (wet)**	**Enthalpy kJ/mol**	**%-mol (wet)**	**Enthalpy kJ/mol**	**%-mol (wet)**	**Enthalpy kJ/mol**
CH_4	0.00%	59.11	9.7%	1.83	0.00%	91.04
CO	0.00%	30.94	18.5%	1.46	0.00%	44.03
CO_2	0.00%	48.61	18.6%	1.91	2.94%	70.31
H_2	0.00%	29.07	41.0%	1.45	0.00%	41.06
N_2	77.5%	30.58	0.06%	1.46	75.4%	43.53
O_2	20.6%	32.37	0.00%	1.48	16.8%	45.97
SO_2	0.00%	49.55	0.00%	2.04	0.03%	70.78
H_2O	1.88%	37.73	12.2%	1.69	4.74%	55.15
TOTAL	**100%**	**31.08**	**100%**	**1.60**	**100%**	**45.29**

6. Air and flue gas enthalpies for bio-SNG combustion:

Phase	**Air**		**Flue Gas**	
Temperature [K]	1 273		1 646	
Component	**%-mol (wet)**	**Enthalpy kJ/mol**	**%-mol (wet)**	**Enthalpy kJ/mol**
CH_4	0.00%	59.11	0.00%	91.04
CO	0.00%	30.94	0.00%	44.03
CO_2	0.00%	48.61	2.28%	70.31
H_2	0.00%	29.07	0.00%	41.06
N_2	77.5%	30.58	76.6%	43.53
O_2	20.6%	32.37	17.1%	45.97
SO_2	0.00%	49.55	0.03%	70.78
H_2O	1.88%	37.73	4.00%	55.15
TOTAL	**100%**	**31.08**	**100%**	**45.03**

APPENDIX C: Detailed Investment Costs Methodology

Item	Considerations	OBS	Reference
Vertical Silo	Fuel consumption 40 t/h (960 t/day) for all gasification technologies; 5 day storage (4,800 t)	150,000 USD for 3,000 tons vertical silo	(Reilly 2011)
Belt Conveyors	1 m wide 500 m long, for the two occurrences	c = 46,000 + 1,320*length	(Towler et al. 2012)
Belt Dryer	Total deck surface of 350 m^2 required for evaporating 25 t/h moisture	$c = 1,550*21*S^{0.59}$	(Sztabert et al. 2006)
Gasification	Includes all equipment described before	Equipment not considered in this study, (e.g. tar reformer) were not included	(Worley et al. 2012)
Air injection system	From NREL report	-	Id.
Steam boiler	Packaged, 15 to 40 bars; 23,000 kg of steam/h	c = 124,000 + 10*flow rate	(Towler et al. 2012)
Methanation	Includes all equipment described before	Lowest cost option	(Gassner et al. 2009)
Nitrogen system	From NREL report	-	(Worley et al. 2012)
Installation	10% over equipment costs	Adapted from NREL report	-
Civil & foundation	4.0%	From NREL report	(Worley et al. 2012)
Buildings	26%	From NREL report	Id.
Piping	1.0%	From NREL report	Id.
AEI	5.0%	From NREL report	Id.
Insulation & painting	0.5%	From NREL report	Id.
Engineering &	10%	From NREL report	Id.

Supervision			
Pre-project cost	0.5%	From NREL report	Id.
Outside engineering & construction	2.0%	From NREL report	Id.
Sales taxes	3.5%	From NREL report	Id.
Freight	5.0%	Adapted from NREL report	-
Training & start-up	2.0%	From NREL report	(Worley et al. 2012)
Owner engineering	2.0%	From NREL report	Id.
Environmental and legal expenses	1.0%	From NREL report	Id.
Capitalized spares	3.0%	From NREL report	Id.
Contingency	15%	From NREL report	Id.

APPENDIX D: Detailed Investment Costs Results

1. Air Gasification:

Size [MW]	**3**	**12**	**24**	**60**	**157**
Factor	**0.063**	**0.165**	**0.269**	**0.510**	**1.000**
Item	**k€**	**k€**	**k€**	**k€**	**k€**
DRYING	**170**	**450**	**720**	**1 380**	**2 700**
Storage Silo	10	30	50	100	200
Conveyor (wet)	40	100	160	310	600
Belt Dryer	90	250	400	770	1 500
Conveyor (dry)	30	70	110	200	400
GASIFIER	**810**	**2 170**	**3 510**	**6 690**	**13 100**
H&S Distribution	410	1 090	1 770	3 370	6 600
H&S Feeding	90	250	400	770	1 500
H&S Bed Mat.	10	30	50	100	200
CFB gasifier	290	780	1 260	2 400	4 700
H&S Gasifier ash	10	20	30	50	100
Air system	10	20	30	50	100
OTHER ISBL	**140**	**380**	**620**	**1 180**	**2 300**
Nitrogen system	40	120	190	360	700
Installation	100	260	430	820	1 600
EQUIPMENT	**1 130**	**3 020**	**4 880**	**9 300**	**18 200**
OTHER DIRECT	**410**	**1 100**	**1 770**	**3 370**	**6 600**
Civil & foundation	40	120	190	360	700
Buildings	290	780	1 260	2 400	4 700
Piping	10	30	50	100	200
AEI	60	150	240	460	900
Insulation & painting	10	20	30	50	100
TDC	**1 540**	**4 120**	**6 650**	**12 670**	**24 800**
INDIRECT (ISBL)	**360**	**930**	**1 490**	**2 860**	**5 600**
Engineering & Supervision	150	400	640	1 220	2 400
Pre-project cost	10	20	30	50	100

Outside eng. & construction	30	80	130	260	500
Sales taxes	60	150	240	460	900
Freight	80	200	320	610	1 200
Training & start-up	30	80	130	260	500
INDIRECT (OSBL)	**310**	**840**	**1 360**	**2 610**	**5 100**
Owner engineering	30	80	130	260	500
Environmental & Legal	10	30	50	100	200
Capitalized spares	40	120	190	360	700
Contingency	230	610	990	1 890	3 700
TIC	**670**	**1 770**	**2 850**	**5 470**	**10 700**
TD&IC	**2 210**	**5 890**	**9 500**	**18 140**	**35 500**

2. Oxygen Gasification:

Size [MW]	**3**	**12**	**24**	**60**	**132**
Factor	**0.071**	**0.187**	**0.303**	**0.576**	**1.000**
Item	**k€**	**k€**	**k€**	**k€**	**k€**
DRYING	**190**	**500**	**810**	**1 560**	**2 700**
Storage Silo	10	40	60	120	200
Conveyor (wet)	40	110	180	350	600
Belt Dryer	110	280	450	860	1 500
Conveyor (dry)	30	70	120	230	400
GASIFIER	**800**	**2 130**	**3 450**	**6 570**	**11 400**
H&S Distribution	410	1 080	1 760	3 340	5 800
H&S Feeding	90	240	390	750	1 300
H&S Bed Mat.	10	40	60	120	200
CFB gasifier	280	750	1 210	2 300	4 000
H&S Gasifier ash	10	20	30	60	100
Oxygen system	10	20	30	60	100
ASU	350	930	1 520	2 880	5 000
OTHER ISBL	**170**	**460**	**760**	**1 440**	**2 500**
Nitrogen system	40	110	180	350	600
Labor	130	350	580	1 090	1 900

EQUIPMENT	**1 520**	**4 040**	**6 570**	**12 510**	**21 700**
OTHER DIRECT	**560**	**1 490**	**2 390**	**4 550**	**7 900**
Civil & foundation	60	170	270	520	900
Buildings	400	1 050	1 700	3 220	5 600
Piping	10	40	60	120	200
AEI	80	210	330	630	1 100
Insulation & painting	10	20	30	60	100
TDC	**2 080**	**5 530**	**8 960**	**17 060**	**29 600**
INDIRECT (ISBL)	**480**	**1 250**	**2 020**	**3 870**	**6 700**
Engineering & Supervision	210	540	880	1 670	2 900
Pre-project cost	10	20	30	60	100
Outside eng. & construction	40	110	180	350	600
Sales taxes	70	190	300	580	1 000
Freight	110	280	450	860	1 500
Training & start-up	40	110	180	350	600
INDIRECT (OSBL)	**430**	**1 160**	**1 870**	**3 570**	**6 200**
Owner engineering	40	110	180	350	600
Environmental & Legal	20	60	90	170	300
Capitalized spares	60	170	270	520	900
Contingency	310	820	1 330	2 530	4 400
TIC	**910**	**2 410**	**3 890**	**7 440**	**12 900**
TD&IC	**2 990**	**7 940**	**12 850**	**24 500**	**42 500**

3. Steam Gasification:

Size [MW]	**3**	**12**	**24**	**60**	**136**
Factor	**0.069**	**0.183**	**0.297**	**0.564**	**1.000**
Item	**k€**	**k€**	**k€**	**k€**	**k€**
DRYING	**180**	**490**	**810**	**1 530**	**2 700**
Storage Silo	10	40	60	110	200
Conveyor (wet)	40	110	180	340	600
Belt Dryer	100	270	450	850	1 500
Conveyor (dry)	30	70	120	230	400

GASIFIER	**1 240**	**3 280**	**5 320**	**10 100**	**17 900**
H&S Distribution	440	1 170	1 900	3 610	6 400
H&S Feeding	100	260	420	790	1 400
H&S Bed Mat.	10	40	60	110	200
CFB gasifier	310	820	1 340	2 540	4 500
H&S Gasifier ash	10	20	30	60	100
Comb. Chamber	360	950	1 540	2 930	5 200
H&S Combustion ash	10	20	30	60	100
Air system	10	20	30	60	100
Steam system	10	20	30	60	100
Steam boiler	20	50	90	170	300
OTHER ISBL	**190**	**490**	**800**	**1 520**	**2 700**
Nitrogen system	40	110	180	340	600
Labor	150	380	620	1 180	2 100
EQUIPMENT	**1 650**	**4 350**	**7 080**	**13 440**	**23 800**
OTHER DIRECT	**580**	**1 560**	**2 530**	**4 800**	**8 500**
Civil & foundation	60	160	270	510	900
Buildings	420	1 120	1 810	3 440	6 100
Piping	10	40	60	110	200
AEI	80	220	360	680	1 200
Insulation & painting	10	20	30	60	100
TDC	**2 230**	**5 910**	**9 610**	**18 240**	**32 300**
INDIRECT (ISBL)	**500**	**1 330**	**2 180**	**4 110**	**7 300**
Engineering & Supervision	220	580	950	1 800	3 200
Pre-project cost	10	40	60	110	200
Outside eng. & construction	40	110	180	340	600
Sales taxes	80	200	330	620	1 100
Freight	110	290	480	900	1 600
Training & start-up	40	110	180	340	600
INDIRECT (OSBL)	**460**	**1 220**	**2 000**	**3 780**	**6 700**
Owner engineering	40	110	180	340	600
Environmental & Legal	20	50	90	170	300

Capitalized spares	70	180	300	560	1 000
Contingency	330	880	1 430	2 710	4 800
TIC	**960**	**2 550**	**4 180**	**7 890**	**14 000**
TD&IC	**3 190**	**8 460**	**13 790**	**26 130**	**46 300**

4. Steam Gasification with product gas Drying:

Size [MW]	**3**	**12**	**24**	**60**	**128**
Factor	**0.072**	**0.191**	**0.310**	**0.588**	**1.000**
Item	**k€**	**k€**	**k€**	**k€**	**k€**
DRYING	**190**	**520**	**830**	**1 590**	**2 700**
Storage Silo	10	40	60	120	200
Conveyor (wet)	40	110	190	350	600
Belt Dryer	110	290	460	880	1 500
Conveyor (wet)	30	80	120	240	400
GASIFIER	**1 332**	**3 530**	**5 700**	**10 850**	**18 430**
H&S Distribution	470	1 240	2 010	3 820	6 500
H&S Feeding	110	290	460	880	1 500
H&S Bed Mat.	10	40	60	120	200
CFB gasifier	330	880	1 430	2 710	4 600
H&S Gasifier ash	10	20	30	60	100
Comb. Chamber	390	1 030	1 670	3 180	5 400
H&S Combustion ash	10	20	30	60	100
Heat Exchangers	2.0	10.0	10.0	20.0	30.0
GASIFYING MEDIA	**40.0**	**100**	**150**	**300**	**500**
Air system	10	20	30	60	100
Steam system	10	20	30	60	100
Steam boiler	20	60	90	180	300
OTHER ISBL	**200**	**530**	**870**	**1 650**	**2 800**
Nitrogen system	50	130	220	410	700
Labor	150	400	650	1 240	2 100
EQUIPMENT	**1 762**	**4 680**	**7 550**	**14 390**	**24 430**
OTHER DIRECT	**640**	**1 680**	**2 720**	**5 190**	**8 800**

Civil & foundation	70	190	310	590	1 000
Buildings	460	1 200	1 950	3 710	6 300
Piping	10	40	60	120	200
AEI	90	230	370	710	1 200
Insulation & painting	10	20	30	60	100
TOTAL DIRECT	**2 402**	**6 360**	**10 270**	**19 580**	**33 230**
INDIRECT (ISBL)	**560**	**1 480**	**2 420**	**4 590**	**7 800**
Engineering & Supervision	240	630	1 020	1 940	3 300
Pre-project cost	10	40	60	120	200
Outside eng. & construction	50	130	220	410	700
Sales taxes	90	230	370	710	1 200
Freight	120	320	530	1 000	1 700
Training & start-up	50	130	220	410	700
INDIRECT (OSBL)	**500**	**1 330**	**2 170**	**4 120**	**7 000**
Owner engineering	50	130	220	410	700
Environm./legal expenses	20	60	90	180	300
Capitalized spares	70	190	310	590	1 000
Contingency	360	950	1 550	2 940	5 000
TOTAL INDIRECT	**1 060**	**2 810**	**4 590**	**8 710**	**14 800**
TOTAL INVESTMENT	**3 462**	**9 170**	**14 860**	**28 290**	**48 030**

5. Bio-SNG Production:

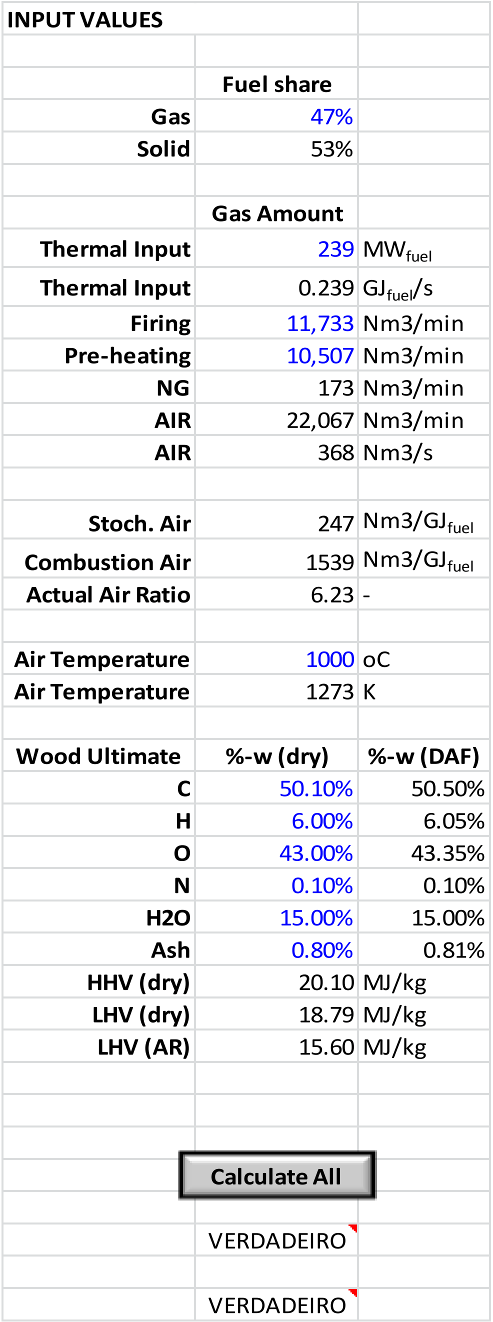
Size [MW]	**3**	**12**	**24**	**60**	**112**
Factor	**0.079**	**0.209**	**0.340**	**0.646**	**1.000**
Item	**k€**	**k€**	**k€**	**k€**	**k€**
DRYING	**220**	**560**	**920**	**1 750**	**2 700**
Storage Silo	20	40	70	130	200
Wet fuel conveyor	50	130	200	390	600
Belt Dryer	120	310	510	970	1 500
Dry fuel conveyor	30	80	140	260	400
GASIFIER	**1 490**	**3 900**	**6 360**	**12 070**	**18 700**
H&S Distribution	520	1 380	2 250	4 260	6 600

H&S Feeding	120	310	510	970	1 500
H&S Bed Material	20	40	70	130	200
CFB gasifier	370	980	1 600	3 040	4 700
H&S Gasifier ash	10	20	30	60	100
Combustion Chamber	440	1 150	1 870	3 550	5 500
H&S Combustion ash	10	20	30	60	100
Air system	10	20	30	60	100
Steam system	10	20	30	60	100
Steam boiler	20	60	100	190	300
METHANATION	**3 670.0**	**9 680**	**15 720**	**29 850**	**46 200**
Gas Conditioning	830	2 180	3 540	6 720	10 400
Reactor	1 310	3 460	5 610	10 660	16 500
SNG-Upgrading	1 530	4 040	6 570	12 470	19 300
OTHER ISBL	**600**	**1 570**	**2 550**	**4 840**	**7 500**
Nitrogen system	60	150	240	450	700
Labor	540	1 420	2 310	4 390	6 800
EQUIPMENT	**6 020**	**15 810**	**25 710**	**48 820**	**75 600**
OTHER DIRECT	**2 190**	**5 810**	**9 420**	**17 900**	**27 700**
Civil & foundation	240	630	1 020	1 940	3 000
Buildings	1 560	4 130	6 700	12 730	19 700
Piping	60	170	270	520	800
AEI	300	800	1 290	2 450	3 800
Insulation & painting	30	80	140	260	400
TDC	**8 210**	**21 620**	**35 130**	**66 720**	**103 300**
INDIRECT (ISBL)	**1 900**	**4 980**	**8 080**	**15 380**	**23 800**
Engineering & Supervision	820	2 160	3 500	6 650	10 300
Pre-project cost	40	100	170	320	500
Outside eng. & construction	170	440	710	1 360	2 100
Sales taxes	290	750	1 220	2 330	3 600
Freight	410	1 090	1 770	3 360	5 200
Training & start-up	170	440	710	1 360	2 100
INDIRECT (OSBL)	**1 730**	**4 550**	**7 370**	**14 020**	**21 700**

Owner engineering	170	440	710	1 360	2 100
Environmental & Legal	80	210	340	650	1 000
Capitalized spares	250	650	1 050	2 000	3 100
Contingency	1 230	3 250	5 270	10 010	15 500
TIC	**3 630**	**9 530**	**15 450**	**29 400**	**45 500**
TD&IC	**11 840**	**31 150**	**50 580**	**96 120**	**148 800**

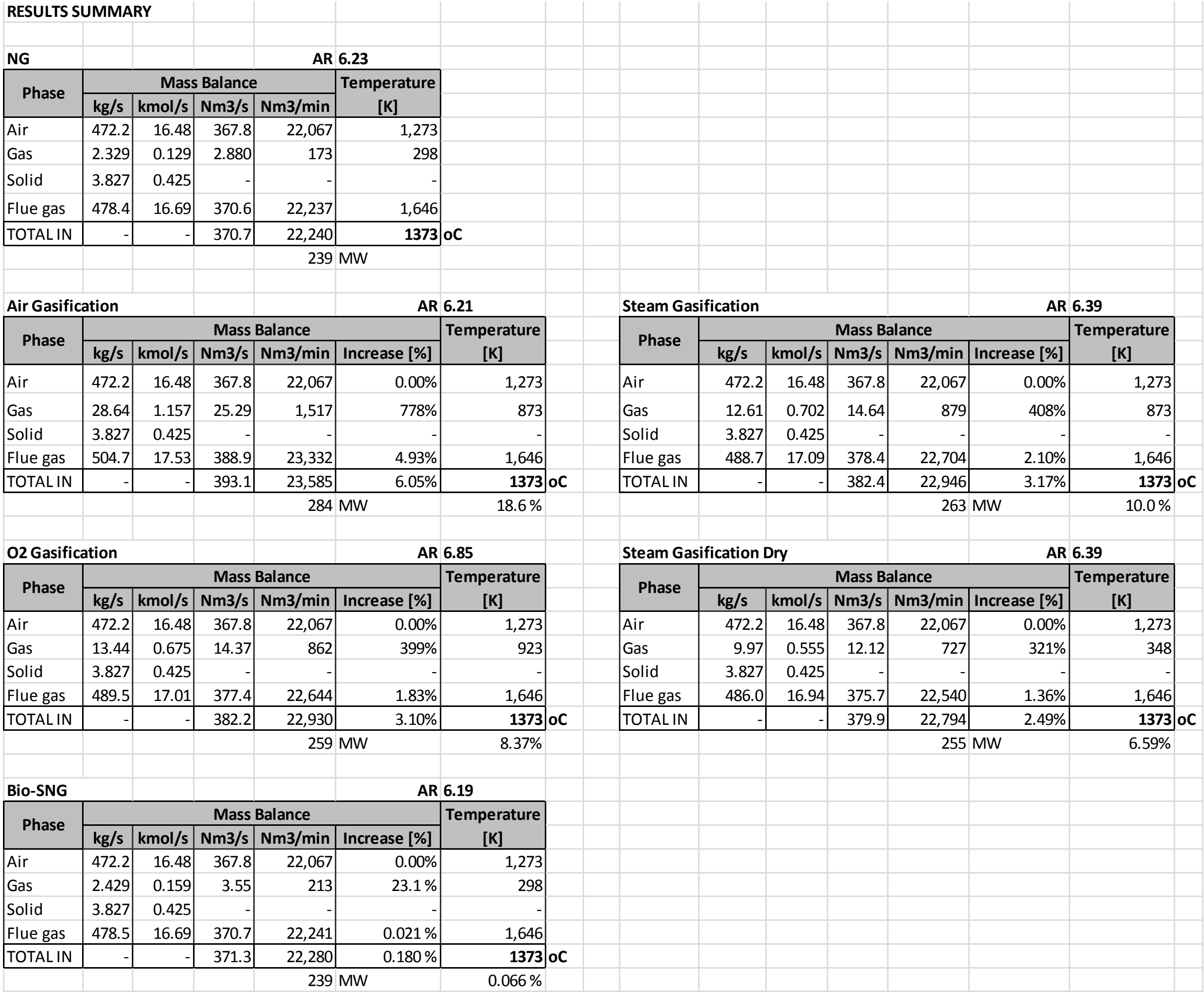
APPENDIX E: Excel Spreadsheet Designed for Calculation

1. Input Data



INPUT VALUES		
	Fuel share	
Gas	47%	
Solid	53%	
	Gas Amount	
Thermal Input	239	MW_{fuel}
Thermal Input	0.239	GJ_{fuel}/s
Firing	11,733	Nm3/min
Pre-heating	10,507	Nm3/min
NG	173	Nm3/min
AIR	22,067	Nm3/min
AIR	368	Nm3/s
Stoch. Air	247	$Nm3/GJ_{fuel}$
Combustion Air	1539	$Nm3/GJ_{fuel}$
Actual Air Ratio	6.23	-
Air Temperature	1000	oC
Air Temperature	1273	K
Wood Ultimate	%-w (dry)	%-w (DAF)
C	50.10%	50.50%
H	6.00%	6.05%
O	43.00%	43.35%
N	0.10%	0.10%
H2O	15.00%	15.00%
Ash	0.80%	0.81%
HHV (dry)	20.10	MJ/kg
LHV (dry)	18.79	MJ/kg
LHV (AR)	15.60	MJ/kg
	Calculate All	
	VERDADEIRO	
	VERDADEIRO	

2. Summary



RESULTS SUMMARY

NG AR **6.23**

Phase	Mass Balance				Temperature
	kg/s	kmol/s	Nm3/s	Nm3/min	[K]
Air	472.2	16.48	367.8	22,067	1,273
Gas	2.329	0.129	2.880	173	298
Solid	3.827	0.425	-	-	-
Flue gas	478.4	16.69	370.6	22,237	1,646
TOTAL IN	-	-	370.7	22,240	**1373** oC
				239	MW

Air Gasification AR **6.21**

Phase	Mass Balance					Temperature
	kg/s	kmol/s	Nm3/s	Nm3/min	Increase [%]	[K]
Air	472.2	16.48	367.8	22,067	0.00%	1,273
Gas	28.64	1.157	25.29	1,517	778%	873
Solid	3.827	0.425	-	-	-	-
Flue gas	504.7	17.53	388.9	23,332	4.93%	1,646
TOTAL IN	-	-	393.1	23,585	6.05%	**1373** oC
				284	MW	18.6 %

Steam Gasification AR **6.39**

Phase	Mass Balance					Temperature
	kg/s	kmol/s	Nm3/s	Nm3/min	Increase [%]	[K]
Air	472.2	16.48	367.8	22,067	0.00%	1,273
Gas	12.61	0.702	14.64	879	408%	873
Solid	3.827	0.425	-	-	-	-
Flue gas	488.7	17.09	378.4	22,704	2.10%	1,646
TOTAL IN	-	-	382.4	22,946	3.17%	**1373** oC
				263	MW	10.0 %

O2 Gasification AR **6.85**

Phase	Mass Balance					Temperature
	kg/s	kmol/s	Nm3/s	Nm3/min	Increase [%]	[K]
Air	472.2	16.48	367.8	22,067	0.00%	1,273
Gas	13.44	0.675	14.37	862	399%	923
Solid	3.827	0.425	-	-	-	-
Flue gas	489.5	17.01	377.4	22,644	1.83%	1,646
TOTAL IN	-	-	382.2	22,930	3.10%	**1373** oC
				259	MW	8.37%

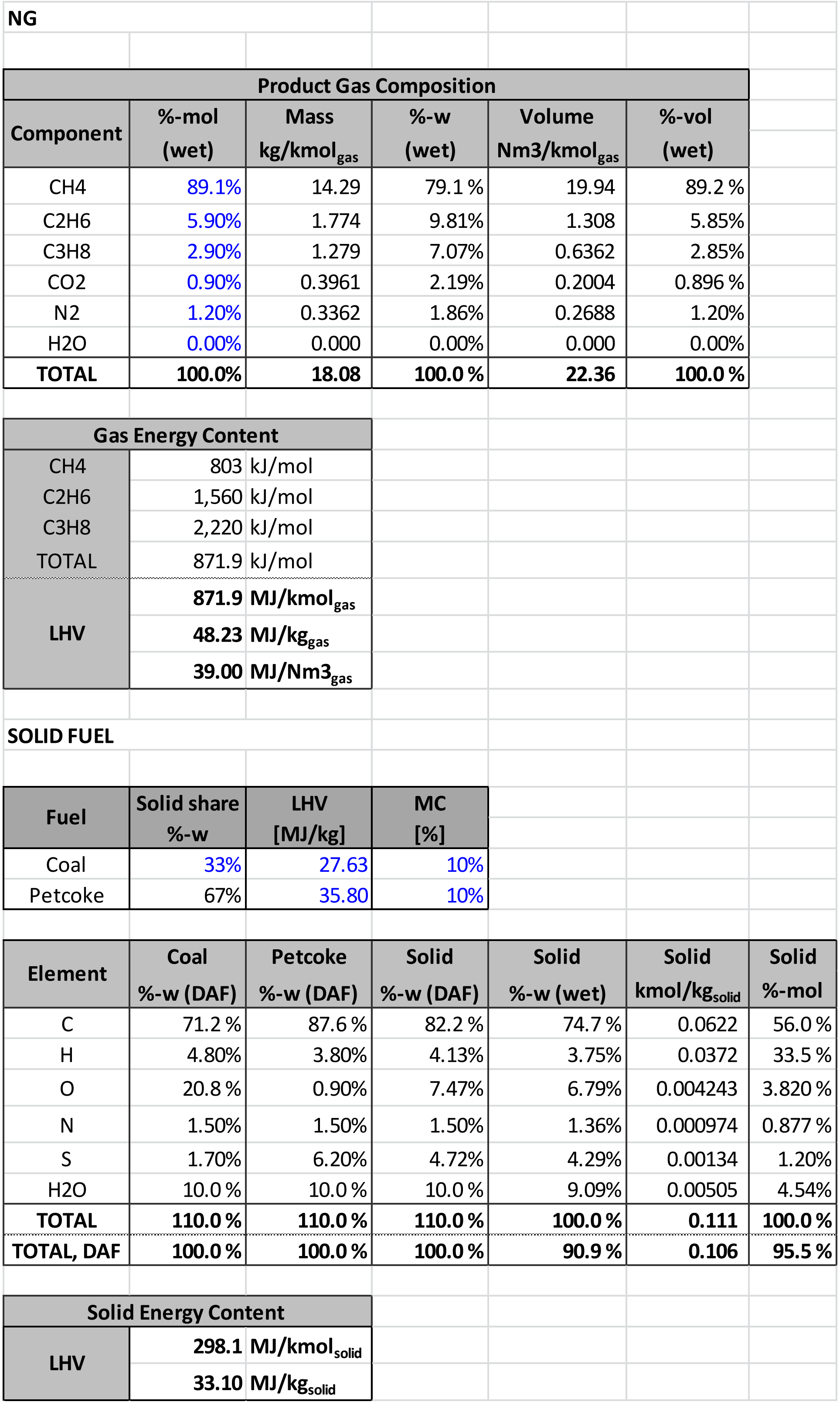
Steam Gasification Dry AR **6.39**

Phase	Mass Balance					Temperature
	kg/s	kmol/s	Nm3/s	Nm3/min	Increase [%]	[K]
Air	472.2	16.48	367.8	22,067	0.00%	1,273
Gas	9.97	0.555	12.12	727	321%	348
Solid	3.827	0.425	-	-	-	-
Flue gas	486.0	16.94	375.7	22,540	1.36%	1,646
TOTAL IN	-	-	379.9	22,794	2.49%	**1373** oC
				255	MW	6.59%

Bio-SNG AR **6.19**

Phase	Mass Balance					Temperature
	kg/s	kmol/s	Nm3/s	Nm3/min	Increase [%]	[K]
Air	472.2	16.48	367.8	22,067	0.00%	1,273
Gas	2.429	0.159	3.55	213	23.1 %	298
Solid	3.827	0.425	-	-	-	-
Flue gas	478.5	16.69	370.7	22,241	0.021 %	1,646
TOTAL IN	-	-	371.3	22,280	0.180 %	**1373** oC
				239	MW	0.066 %

3. NG and solid fuel composition:



NG

Product Gas Composition					
Component	**%-mol (wet)**	**Mass kg/kmol$_{gas}$**	**%-w (wet)**	**Volume Nm3/kmol$_{gas}$**	**%-vol (wet)**
CH4	89.1%	14.29	79.1 %	19.94	89.2 %
C2H6	5.90%	1.774	9.81%	1.308	5.85%
C3H8	2.90%	1.279	7.07%	0.6362	2.85%
CO2	0.90%	0.3961	2.19%	0.2004	0.896 %
N2	1.20%	0.3362	1.86%	0.2688	1.20%
H2O	0.00%	0.000	0.00%	0.000	0.00%
TOTAL	**100.0%**	**18.08**	**100.0 %**	**22.36**	**100.0 %**

Gas Energy Content		
CH4	803	kJ/mol
C2H6	1,560	kJ/mol
C3H8	2,220	kJ/mol
TOTAL	871.9	kJ/mol
LHV	**871.9**	**MJ/kmol$_{gas}$**
	48.23	**MJ/kg$_{gas}$**
	39.00	**MJ/Nm3$_{gas}$**

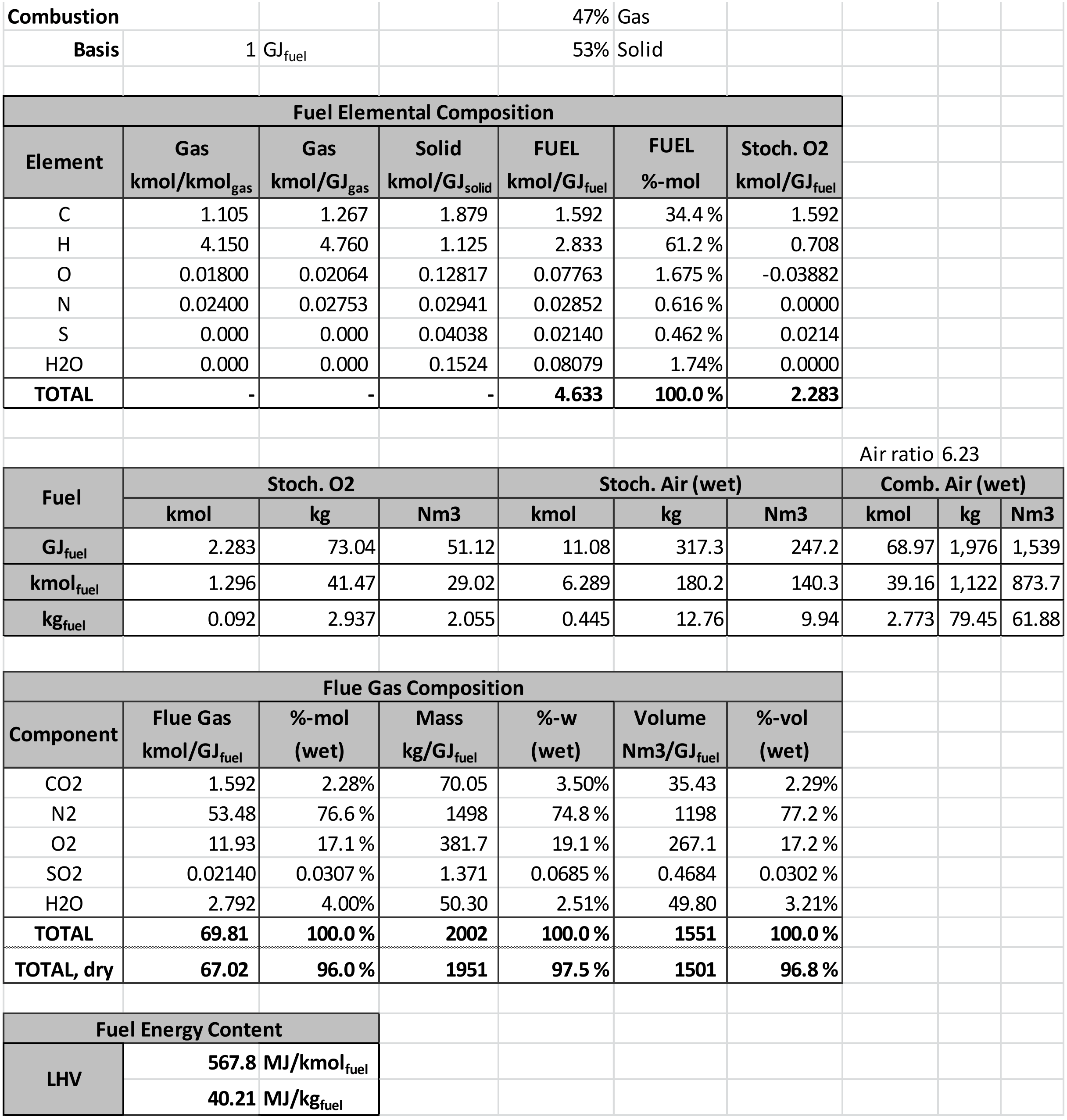
SOLID FUEL

Fuel	**Solid share %-w**	**LHV [MJ/kg]**	**MC [%]**
Coal	33%	27.63	10%
Petcoke	67%	35.80	10%

Element	**Coal %-w (DAF)**	**Petcoke %-w (DAF)**	**Solid %-w (DAF)**	**Solid %-w (wet)**	**Solid kmol/kg$_{solid}$**	**Solid %-mol**
C	71.2 %	87.6 %	82.2 %	74.7 %	0.0622	56.0 %
H	4.80%	3.80%	4.13%	3.75%	0.0372	33.5 %
O	20.8 %	0.90%	7.47%	6.79%	0.004243	3.820 %
N	1.50%	1.50%	1.50%	1.36%	0.000974	0.877 %
S	1.70%	6.20%	4.72%	4.29%	0.00134	1.20%
H2O	10.0 %	10.0 %	10.0 %	9.09%	0.00505	4.54%
TOTAL	**110.0 %**	**110.0 %**	**110.0 %**	**100.0 %**	**0.111**	**100.0 %**
TOTAL, DAF	**100.0 %**	**100.0 %**	**100.0 %**	**90.9 %**	**0.106**	**95.5 %**

Solid Energy Content		
LHV	**298.1**	**MJ/kmol$_{solid}$**
	33.10	**MJ/kg$_{solid}$**

4. NG and solid fuel combustion:



Combustion				47%	Gas
Basis	1	GJ_{fuel}		53%	Solid

Fuel Elemental Composition						
Element	**Gas** **$kmol/kmol_{gas}$**	**Gas** **$kmol/GJ_{gas}$**	**Solid** **$kmol/GJ_{solid}$**	**FUEL** **$kmol/GJ_{fuel}$**	**FUEL** **%-mol**	**Stoch. O2** **$kmol/GJ_{fuel}$**
C	1.105	1.267	1.879	1.592	34.4 %	1.592
H	4.150	4.760	1.125	2.833	61.2 %	0.708
O	0.01800	0.02064	0.12817	0.07763	1.675 %	-0.03882
N	0.02400	0.02753	0.02941	0.02852	0.616 %	0.0000
S	0.000	0.000	0.04038	0.02140	0.462 %	0.0214
H2O	0.000	0.000	0.1524	0.08079	1.74%	0.0000
TOTAL	**-**	**-**	**-**	**4.633**	**100.0 %**	**2.283**

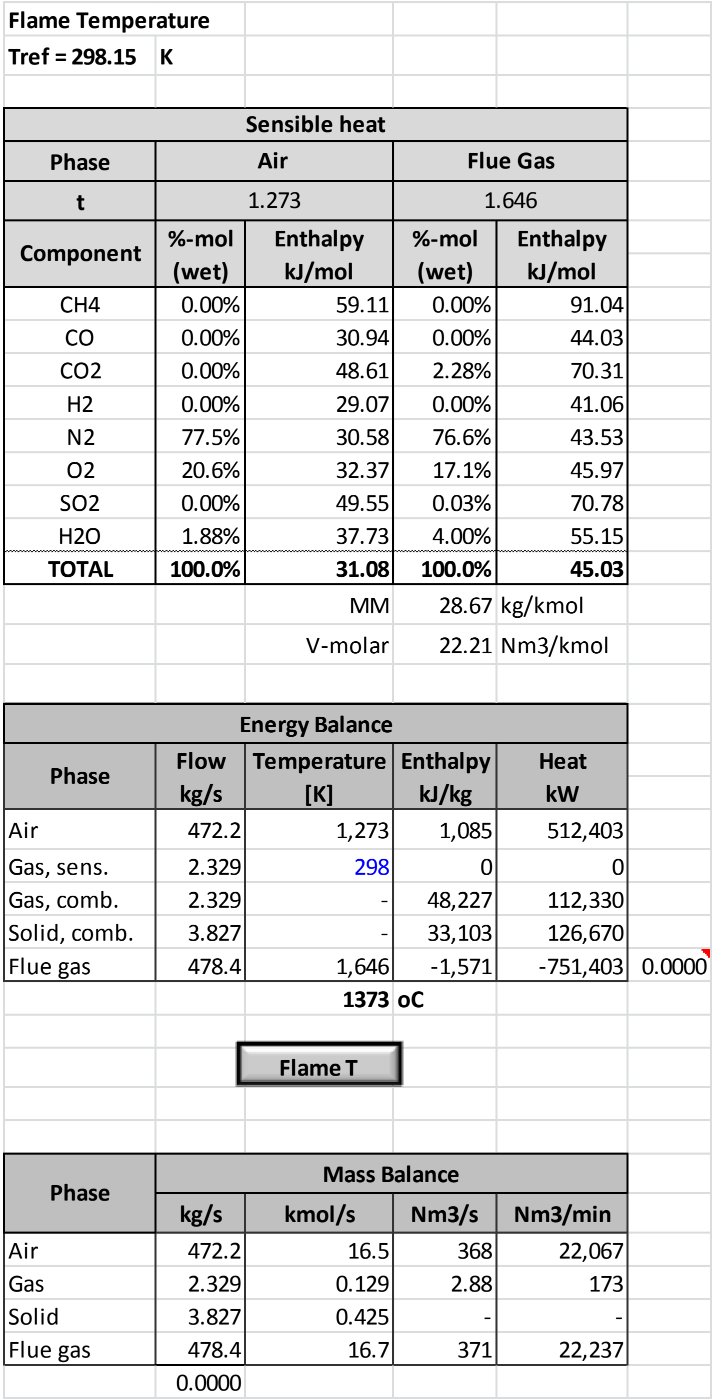
Air ratio 6.23

Fuel	**Stoch. O2**			**Stoch. Air (wet)**			**Comb. Air (wet)**		
	kmol	**kg**	**Nm3**	**kmol**	**kg**	**Nm3**	**kmol**	**kg**	**Nm3**
GJ_{fuel}	2.283	73.04	51.12	11.08	317.3	247.2	68.97	1,976	1,539
$kmol_{fuel}$	1.296	41.47	29.02	6.289	180.2	140.3	39.16	1,122	873.7
kg_{fuel}	0.092	2.937	2.055	0.445	12.76	9.94	2.773	79.45	61.88

Flue Gas Composition						
Component	**Flue Gas** **$kmol/GJ_{fuel}$**	**%-mol** **(wet)**	**Mass** **kg/GJ_{fuel}**	**%-w** **(wet)**	**Volume** **$Nm3/GJ_{fuel}$**	**%-vol** **(wet)**
CO2	1.592	2.28%	70.05	3.50%	35.43	2.29%
N2	53.48	76.6 %	1498	74.8 %	1198	77.2 %
O2	11.93	17.1 %	381.7	19.1 %	267.1	17.2 %
SO2	0.02140	0.0307 %	1.371	0.0685 %	0.4684	0.0302 %
H2O	2.792	4.00%	50.30	2.51%	49.80	3.21%
TOTAL	**69.81**	**100.0 %**	**2002**	**100.0 %**	**1551**	**100.0 %**
TOTAL, dry	**67.02**	**96.0 %**	**1951**	**97.5 %**	**1501**	**96.8 %**

Fuel Energy Content		
LHV	**567.8**	**$MJ/kmol_{fuel}$**
	40.21	**MJ/kg_{fuel}**

5. NG and solid fuel flame temperature:



Flame Temperature

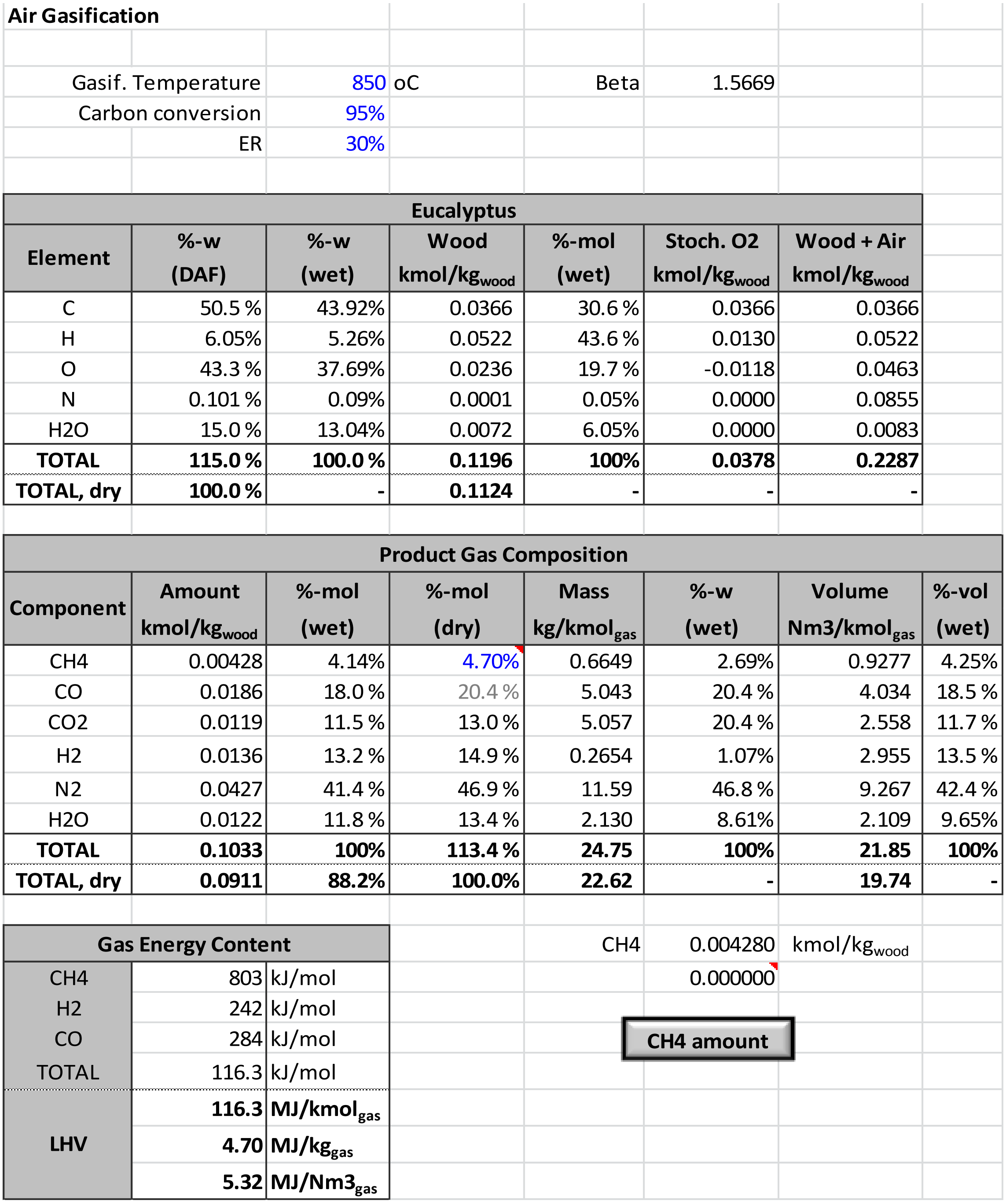
Tref = 298.15 K

Sensible heat				
Phase	**Air**		**Flue Gas**	
t	1.273		1.646	
Component	**%-mol (wet)**	**Enthalpy kJ/mol**	**%-mol (wet)**	**Enthalpy kJ/mol**
CH4	0.00%	59.11	0.00%	91.04
CO	0.00%	30.94	0.00%	44.03
CO2	0.00%	48.61	2.28%	70.31
H2	0.00%	29.07	0.00%	41.06
N2	77.5%	30.58	76.6%	43.53
O2	20.6%	32.37	17.1%	45.97
SO2	0.00%	49.55	0.03%	70.78
H2O	1.88%	37.73	4.00%	55.15
TOTAL	**100.0%**	**31.08**	**100.0%**	**45.03**
		MM	28.67	kg/kmol
		V-molar	22.21	Nm3/kmol

Energy Balance					
Phase	**Flow kg/s**	**Temperature [K]**	**Enthalpy kJ/kg**	**Heat kW**	
Air	472.2	1,273	1,085	512,403	
Gas, sens.	2.329	298	0	0	
Gas, comb.	2.329	-	48,227	112,330	
Solid, comb.	3.827	-	33,103	126,670	
Flue gas	478.4	1,646	-1,571	-751,403	0.0000
		1373	**oC**		

Phase	**Mass Balance**			
	kg/s	**kmol/s**	**Nm3/s**	**Nm3/min**
Air	472.2	16.5	368	22,067
Gas	2.329	0.129	2.88	173
Solid	3.827	0.425	-	-
Flue gas	478.4	16.7	371	22,237
	0.0000			

6. Air gasification product gas composition:



Air Gasification

Gasif. Temperature	850	oC	Beta	1.5669
Carbon conversion	95%			
ER	30%			

Eucalyptus						
Element	**%-w (DAF)**	**%-w (wet)**	**Wood kmol/kg_{wood}**	**%-mol (wet)**	**Stoch. O2 kmol/kg_{wood}**	**Wood + Air kmol/kg_{wood}**
C	50.5 %	43.92%	0.0366	30.6 %	0.0366	0.0366
H	6.05%	5.26%	0.0522	43.6 %	0.0130	0.0522
O	43.3 %	37.69%	0.0236	19.7 %	-0.0118	0.0463
N	0.101 %	0.09%	0.0001	0.05%	0.0000	0.0855
H2O	15.0 %	13.04%	0.0072	6.05%	0.0000	0.0083
TOTAL	**115.0 %**	**100.0 %**	**0.1196**	**100%**	**0.0378**	**0.2287**
TOTAL, dry	**100.0 %**	**-**	**0.1124**	**-**	**-**	**-**

Product Gas Composition							
Component	**Amount kmol/kg_{wood}**	**%-mol (wet)**	**%-mol (dry)**	**Mass kg/kmol_{gas}**	**%-w (wet)**	**Volume Nm3/kmol_{gas}**	**%-vol (wet)**
CH4	0.00428	4.14%	4.70%	0.6649	2.69%	0.9277	4.25%
CO	0.0186	18.0 %	20.4 %	5.043	20.4 %	4.034	18.5 %
CO2	0.0119	11.5 %	13.0 %	5.057	20.4 %	2.558	11.7 %
H2	0.0136	13.2 %	14.9 %	0.2654	1.07%	2.955	13.5 %
N2	0.0427	41.4 %	46.9 %	11.59	46.8 %	9.267	42.4 %
H2O	0.0122	11.8 %	13.4 %	2.130	8.61%	2.109	9.65%
TOTAL	**0.1033**	**100%**	**113.4 %**	**24.75**	**100%**	**21.85**	**100%**
TOTAL, dry	**0.0911**	**88.2%**	**100.0%**	**22.62**	**-**	**19.74**	**-**

Gas Energy Content		
CH4	803	kJ/mol
H2	242	kJ/mol
CO	284	kJ/mol
TOTAL	116.3	kJ/mol
LHV	**116.3**	**MJ/kmol_{gas}**
	4.70	**MJ/kg_{gas}**
	5.32	**MJ/Nm3_{gas}**

CH4	0.004280	kmol/kg_{wood}
	0.000000	

7. Air gasification fuel combustion:

Combustion				52%	Gas
Basis	1	GJ$_{fuel}$		48%	Solid

Fuel Elemental Composition						
Element	**Gas kmol/kmol$_{gas}$**	**Gas kmol/GJ$_{gas}$**	**Solid kmol/GJ$_{solid}$**	**FUEL kmol/GJ$_{fuel}$**	**FUEL %-mol**	**Stoch. O2 kmol/GJ$_{fuel}$**
C	0.3364	2.893	1.879	2.401	21.8 %	2.401
H	0.4291	3.690	1.125	2.447	22.2 %	0.612
O	0.4098	3.525	0.128	1.878	17.0 %	-0.939
N	0.8275	7.117	0.029	3.680	33.4 %	0.000
S	0.0000	0.000	0.040	0.020	0.18%	0.020
H2O	0.1182	1.017	0.152	0.598	5.42%	0.000
TOTAL	**-**	**-**	**-**	**11.02**	**100%**	**2.094**

Air ratio 6.21

Fuel Gas	**Stoch. O2**			**Stoch. Air (wet)**			**Comb. Air (wet)**		
	kmol	**kg**	**Nm3**	**kmol**	**kg**	**Nm3**	**kmol**	**kg**	**Nm3**
GJ$_{fuel}$	2.094	67.00	46.89	10.16	291.1	226.7	63.10	1,808	1,408
kmol$_{fuel}$	0.428	13.70	9.59	2.077	59.50	46.35	12.90	369.5	287.8
kg$_{fuel}$	0.039	1.238	0.866	0.188	5.377	4.188	1.166	33.39	26.01

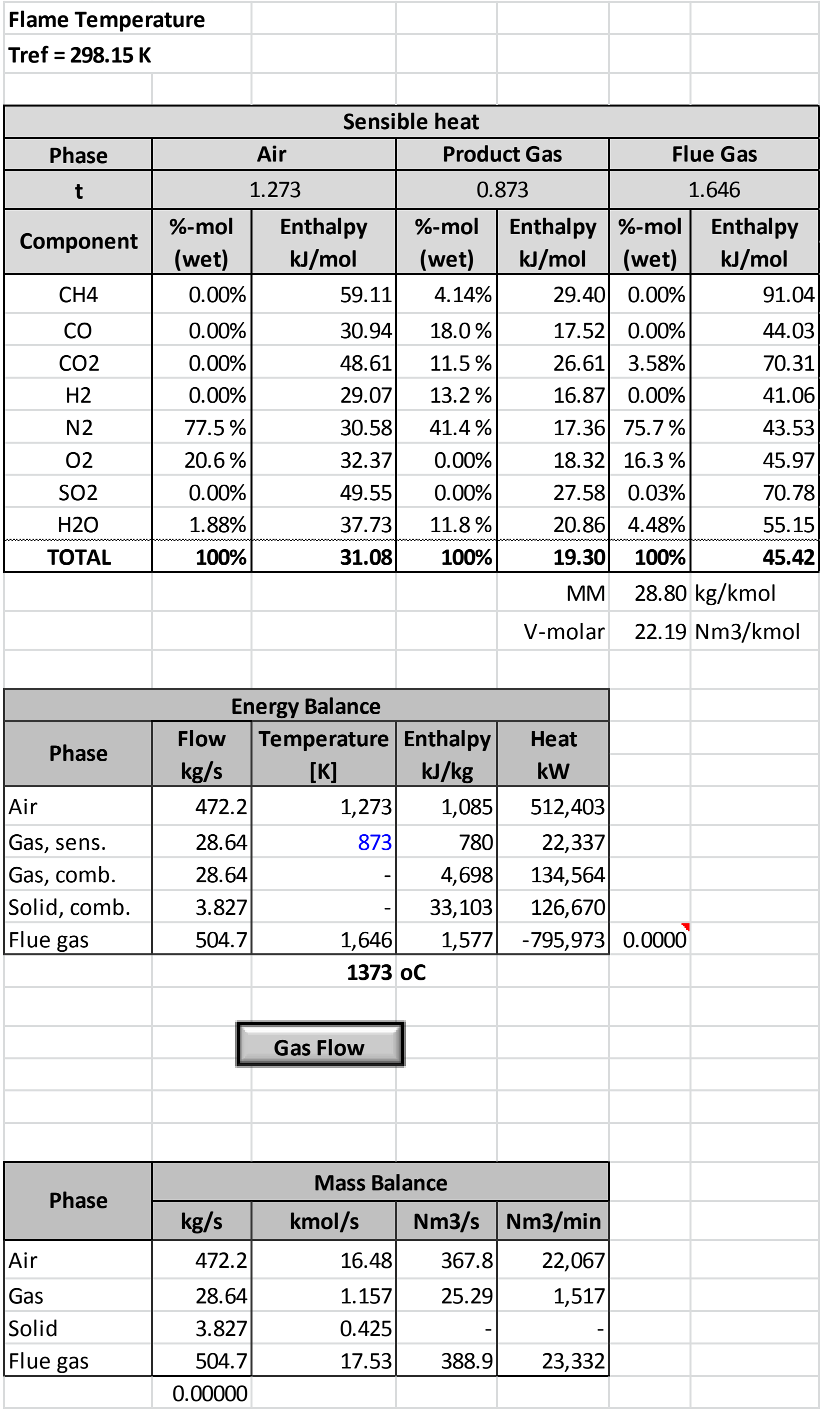
Flue Gas Composition						
Component	**Flue Gas kmol/GJ$_{fuel}$**	**%-mol (wet)**	**Mass kg/GJ$_{fuel}$**	**%-w (wet)**	**Volume Nm3/GJ$_{fuel}$**	**%-vol (wet)**
CO2	2.401	3.58%	105.69	5.47%	53.46	3.58%
N2	50.75	75.7 %	1,422	73.6 %	1,130	75.7 %
O2	10.91	16.3 %	349.1	18.1 %	242.9	16.3 %
SO2	0.020	0.03%	1.254	0.06%	0.436	0.03%
H2O	3.005	4.48%	54.14	2.80%	66.91	4.48%
TOTAL	**67.09**	**100%**	**1,932**	**100%**	**1,494**	**100%**
TOTAL, dry	**64.08**	**95.5 %**	**1878**	**97.2 %**	**1427**	**95.5 %**

Fuel Energy Content		
LHV	**204.4**	**MJ/kmol$_{fuel}$**
	18.47	**MJ/kg$_{fuel}$**

Hot Gas Efficiency		
Input	**Value**	**Unit**
Wood	11.21	kg/s
Carbon	0.25	kg/s
Ash	0.09	kg/s
Loss	0.23	kg/s
Wood	183,743	kW
Gas	156,900	kW
Efficiency	85.4 %	-

SUMMARY					
	Value	**Unit**			
Wood in	11.8	kg/s	42.4	t/h	dry
Planted	27,028	ha	72.1	t/h	wet
Required	54,055	ha			
Gasif. Size	157	MW			
TOTAL	284	MW			
Gas	55%				
Cold gas	261	MW			

8. Air gasification flame temperature:



Flame Temperature

Tref = 298.15 K

Sensible heat						
Phase	**Air**		**Product Gas**		**Flue Gas**	
t	1.273		0.873		1.646	
Component	**%-mol (wet)**	**Enthalpy kJ/mol**	**%-mol (wet)**	**Enthalpy kJ/mol**	**%-mol (wet)**	**Enthalpy kJ/mol**
CH4	0.00%	59.11	4.14%	29.40	0.00%	91.04
CO	0.00%	30.94	18.0 %	17.52	0.00%	44.03
CO2	0.00%	48.61	11.5 %	26.61	3.58%	70.31
H2	0.00%	29.07	13.2 %	16.87	0.00%	41.06
N2	77.5 %	30.58	41.4 %	17.36	75.7 %	43.53
O2	20.6 %	32.37	0.00%	18.32	16.3 %	45.97
SO2	0.00%	49.55	0.00%	27.58	0.03%	70.78
H2O	1.88%	37.73	11.8 %	20.86	4.48%	55.15
TOTAL	**100%**	**31.08**	**100%**	**19.30**	**100%**	**45.42**

MM 28.80 kg/kmol

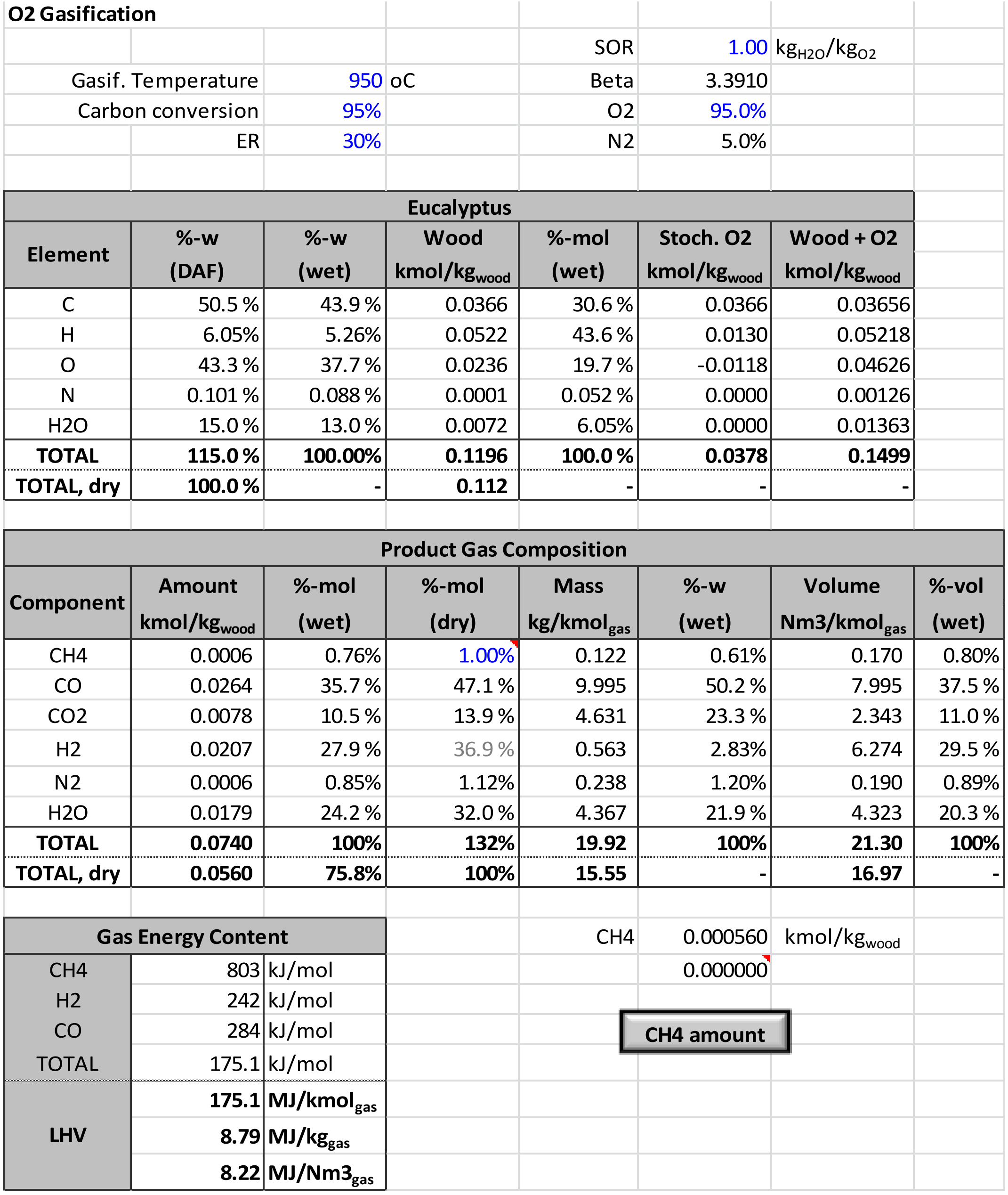
V-molar 22.19 Nm3/kmol

Energy Balance					
Phase	**Flow kg/s**	**Temperature [K]**	**Enthalpy kJ/kg**	**Heat kW**	
Air	472.2	1,273	1,085	512,403	
Gas, sens.	28.64	873	780	22,337	
Gas, comb.	28.64	-	4,698	134,564	
Solid, comb.	3.827	-	33,103	126,670	
Flue gas	504.7	1,646	1,577	-795,973	0.0000
		1373	**oC**		

Gas Flow

Phase	**Mass Balance**			
	kg/s	**kmol/s**	**Nm3/s**	**Nm3/min**
Air	472.2	16.48	367.8	22,067
Gas	28.64	1.157	25.29	1,517
Solid	3.827	0.425	-	-
Flue gas	504.7	17.53	388.9	23,332
	0.00000			

9. Oxygen gasification product gas composition:



O2 Gasification

			SOR	1.00	kg$_{H2O}$/kg$_{O2}$
Gasif. Temperature	950	oC	Beta	3.3910	
Carbon conversion	95%		O2	95.0%	
ER	30%		N2	5.0%	

Eucalyptus

Element	%-w (DAF)	%-w (wet)	Wood kmol/kg$_{wood}$	%-mol (wet)	Stoch. O2 kmol/kg$_{wood}$	Wood + O2 kmol/kg$_{wood}$
C	50.5 %	43.9 %	0.0366	30.6 %	0.0366	0.03656
H	6.05%	5.26%	0.0522	43.6 %	0.0130	0.05218
O	43.3 %	37.7 %	0.0236	19.7 %	-0.0118	0.04626
N	0.101 %	0.088 %	0.0001	0.052 %	0.0000	0.00126
H2O	15.0 %	13.0 %	0.0072	6.05%	0.0000	0.01363
TOTAL	**115.0 %**	**100.00%**	**0.1196**	**100.0 %**	**0.0378**	**0.1499**
TOTAL, dry	**100.0 %**	**-**	**0.112**	**-**	**-**	**-**

Product Gas Composition

Component	Amount kmol/kg$_{wood}$	%-mol (wet)	%-mol (dry)	Mass kg/kmol$_{gas}$	%-w (wet)	Volume Nm3/kmol$_{gas}$	%-vol (wet)
CH4	0.0006	0.76%	1.00%	0.122	0.61%	0.170	0.80%
CO	0.0264	35.7 %	47.1 %	9.995	50.2 %	7.995	37.5 %
CO2	0.0078	10.5 %	13.9 %	4.631	23.3 %	2.343	11.0 %
H2	0.0207	27.9 %	36.9 %	0.563	2.83%	6.274	29.5 %
N2	0.0006	0.85%	1.12%	0.238	1.20%	0.190	0.89%
H2O	0.0179	24.2 %	32.0 %	4.367	21.9 %	4.323	20.3 %
TOTAL	**0.0740**	**100%**	**132%**	**19.92**	**100%**	**21.30**	**100%**
TOTAL, dry	**0.0560**	**75.8%**	**100%**	**15.55**	**-**	**16.97**	**-**

Gas Energy Content

CH4	803	kJ/mol
H2	242	kJ/mol
CO	284	kJ/mol
TOTAL	175.1	kJ/mol
LHV	**175.1**	**MJ/kmol$_{gas}$**
	8.79	**MJ/kg$_{gas}$**
	8.22	**MJ/Nm3$_{gas}$**

CH4	0.000560	kmol/kg$_{wood}$
	0.000000	

10. Oxygen gasification fuel combustion:

Combustion			48%	Gas
Basis	1	GJ_{fuel}	52%	Solid

Fuel Elemental Composition						
Element	**Gas** $kmol/kmol_{gas}$	**Gas** $kmol/GJ_{gas}$	**Solid** $kmol/GJ_{solid}$	**FUEL** $kmol/GJ_{fuel}$	**FUEL** %-mol	**Stoch. O2** $kmol/GJ_{fuel}$
C	0.4696	2.683	1.879	2.267	32.7 %	2.267
H	0.5893	3.366	1.125	2.207	31.8 %	0.552
O	0.5673	3.241	0.128	1.630	23.5 %	-0.815
N	0.0170	0.097	0.029	0.062	0.90%	0.000
S	0.0000	0.000	0.040	0.021	0.30%	0.021
H2O	0.2424	1.385	0.152	0.747	10.8 %	0.000
TOTAL	-	-	-	**6.93**	**100%**	**2.024**

Air ratio 6.85

Fuel Gas	**Stoch. O2**			**Stoch. Air (wet)**			**Comb. Air (wet)**		
	kmol	**kg**	**Nm3**	**kmol**	**kg**	**Nm3**	**kmol**	**kg**	**Nm3**
GJ_{fuel}	2.024	64.78	45.34	9.82	281.4	219.2	67.33	1,929	1,502
$kmol_{fuel}$	0.483	15.46	10.82	2.345	67.18	52.33	16.07	460.4	358.6
kg_{fuel}	0.043	1.384	0.969	0.210	6.014	4.684	1.439	41.22	32.10

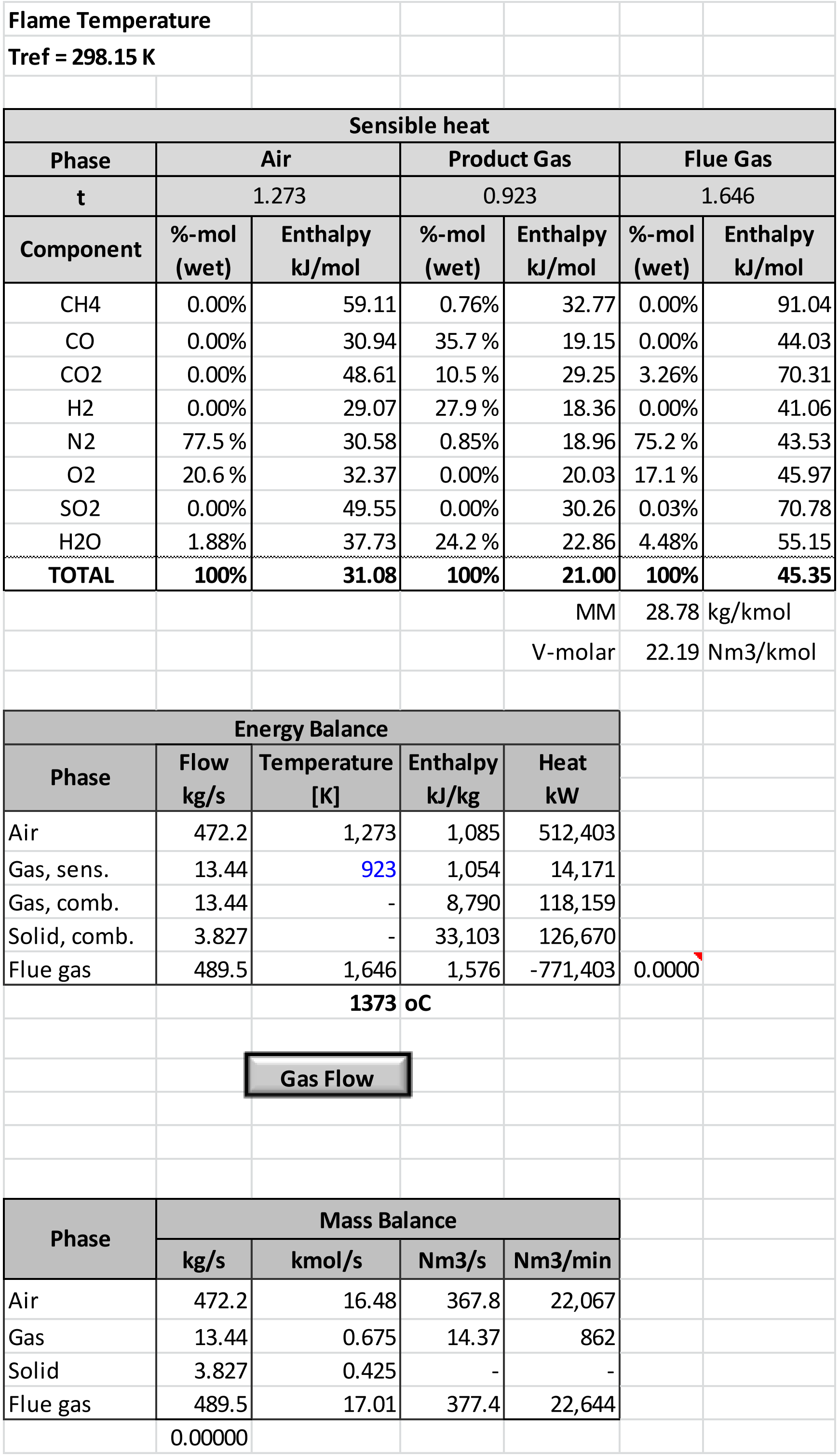
Flue Gas Composition						
Component	**Flue Gas** $kmol/GJ_{fuel}$	**%-mol (wet)**	**Mass** kg/GJ_{fuel}	**%-w (wet)**	**Volume** $Nm3/GJ_{fuel}$	**%-vol (wet)**
CO2	2.267	3.26%	99.77	4.99%	50.47	3.26%
N2	52.22	75.2 %	1,463	73.2 %	1,163	75.2 %
O2	11.85	17.1 %	379.2	19.0 %	263.8	17.1 %
SO2	0.021	0.03%	1.338	0.07%	0.465	0.03%
H2O	3.114	4.48%	56.10	2.81%	69.33	4.48%
TOTAL	**69.47**	**100%**	**1,999**	**100%**	**1,547**	**100%**
TOTAL, dry	**66.36**	**95.5 %**	**1943**	**97.2 %**	**1477**	**95.5 %**

Fuel Energy Content		
LHV	**238.7**	**$MJ/kmol_{fuel}$**
	21.37	**MJ/kg_{fuel}**

Hot Gas Efficiency		
Input	**Value**	**Unit**
Wood	9.13	kg/s
Carbon	0.20	kg/s
Ash	0.08	kg/s
Loss	0.19	kg/s
Wood	149,613	kW
Oxygen	3,761	kW
Gas	132,330	kW
HG eff.	88.4 %	-
Net eff.	86.3 %	-

SUMMARY				
	Value	**Unit**		
Wood in	9.59	kg/s	34.5	t/h
Planted	22,007	ha		
Required	44,015	ha		
Gasif. Size	132	MW		
TOTAL	259	MW		
Gas	51%	-		
Cold gas	245	MW		

11. Oxygen gasification flame temperature:



Flame Temperature

Tref = 298.15 K

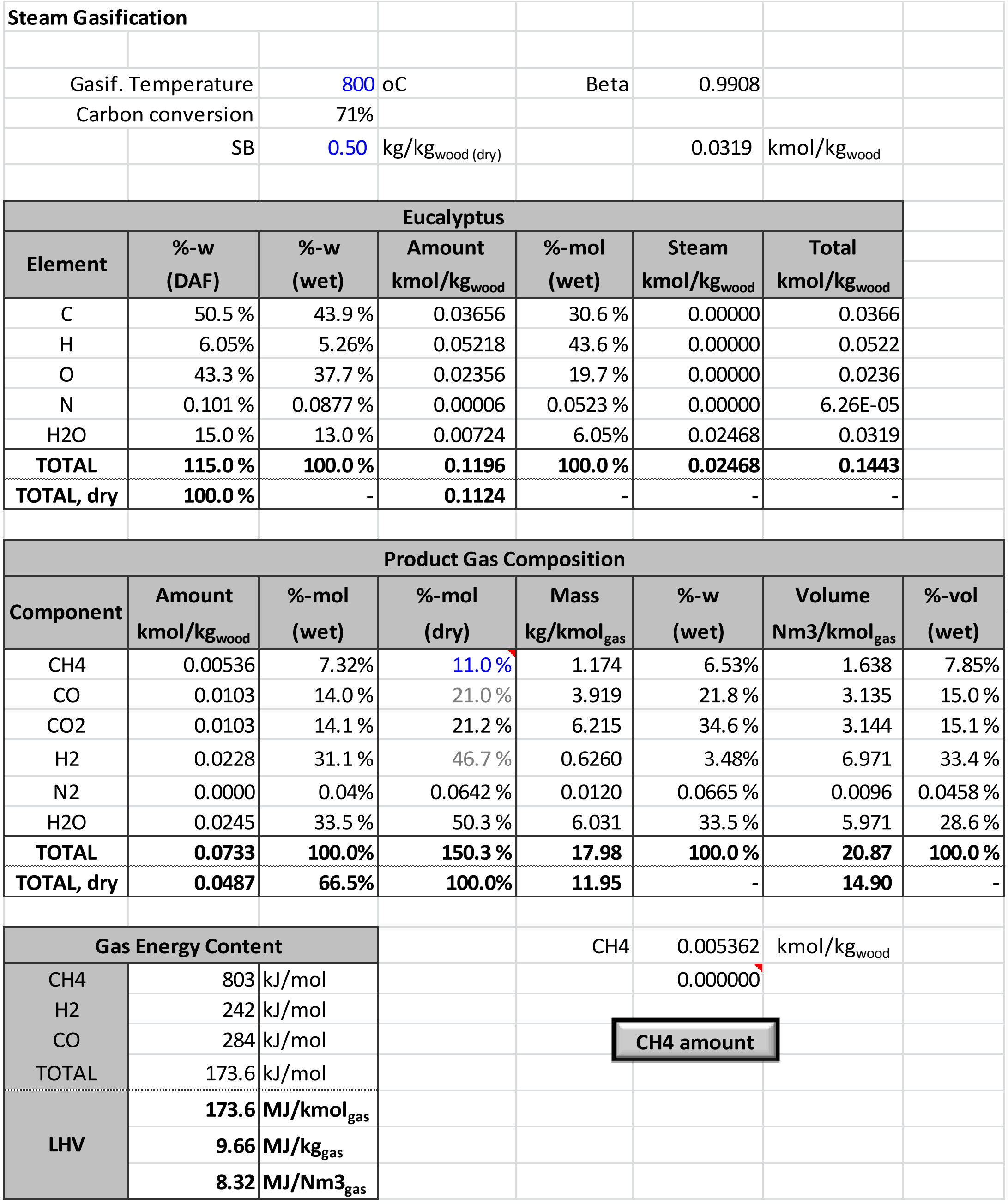
Sensible heat						
Phase	**Air**		**Product Gas**		**Flue Gas**	
t	1.273		0.923		1.646	
Component	**%-mol (wet)**	**Enthalpy kJ/mol**	**%-mol (wet)**	**Enthalpy kJ/mol**	**%-mol (wet)**	**Enthalpy kJ/mol**
CH4	0.00%	59.11	0.76%	32.77	0.00%	91.04
CO	0.00%	30.94	35.7 %	19.15	0.00%	44.03
CO2	0.00%	48.61	10.5 %	29.25	3.26%	70.31
H2	0.00%	29.07	27.9 %	18.36	0.00%	41.06
N2	77.5 %	30.58	0.85%	18.96	75.2 %	43.53
O2	20.6 %	32.37	0.00%	20.03	17.1 %	45.97
SO2	0.00%	49.55	0.00%	30.26	0.03%	70.78
H2O	1.88%	37.73	24.2 %	22.86	4.48%	55.15
TOTAL	**100%**	**31.08**	**100%**	**21.00**	**100%**	**45.35**
				MM	28.78	kg/kmol
				V-molar	22.19	Nm3/kmol

Energy Balance					
Phase	**Flow kg/s**	**Temperature [K]**	**Enthalpy kJ/kg**	**Heat kW**	
Air	472.2	1,273	1,085	512,403	
Gas, sens.	13.44	923	1,054	14,171	
Gas, comb.	13.44	-	8,790	118,159	
Solid, comb.	3.827	-	33,103	126,670	
Flue gas	489.5	1,646	1,576	-771,403	0.0000
		1373	**oC**		

Gas Flow

Phase	**Mass Balance**			
	kg/s	**kmol/s**	**Nm3/s**	**Nm3/min**
Air	472.2	16.48	367.8	22,067
Gas	13.44	0.675	14.37	862
Solid	3.827	0.425	-	-
Flue gas	489.5	17.01	377.4	22,644
	0.00000			

12. Steam gasification product gas composition:



Steam Gasification

Gasif. Temperature	800	oC	Beta	0.9908
Carbon conversion	71%			
SB	0.50	kg/kg$_{wood\ (dry)}$	0.0319	kmol/kg$_{wood}$

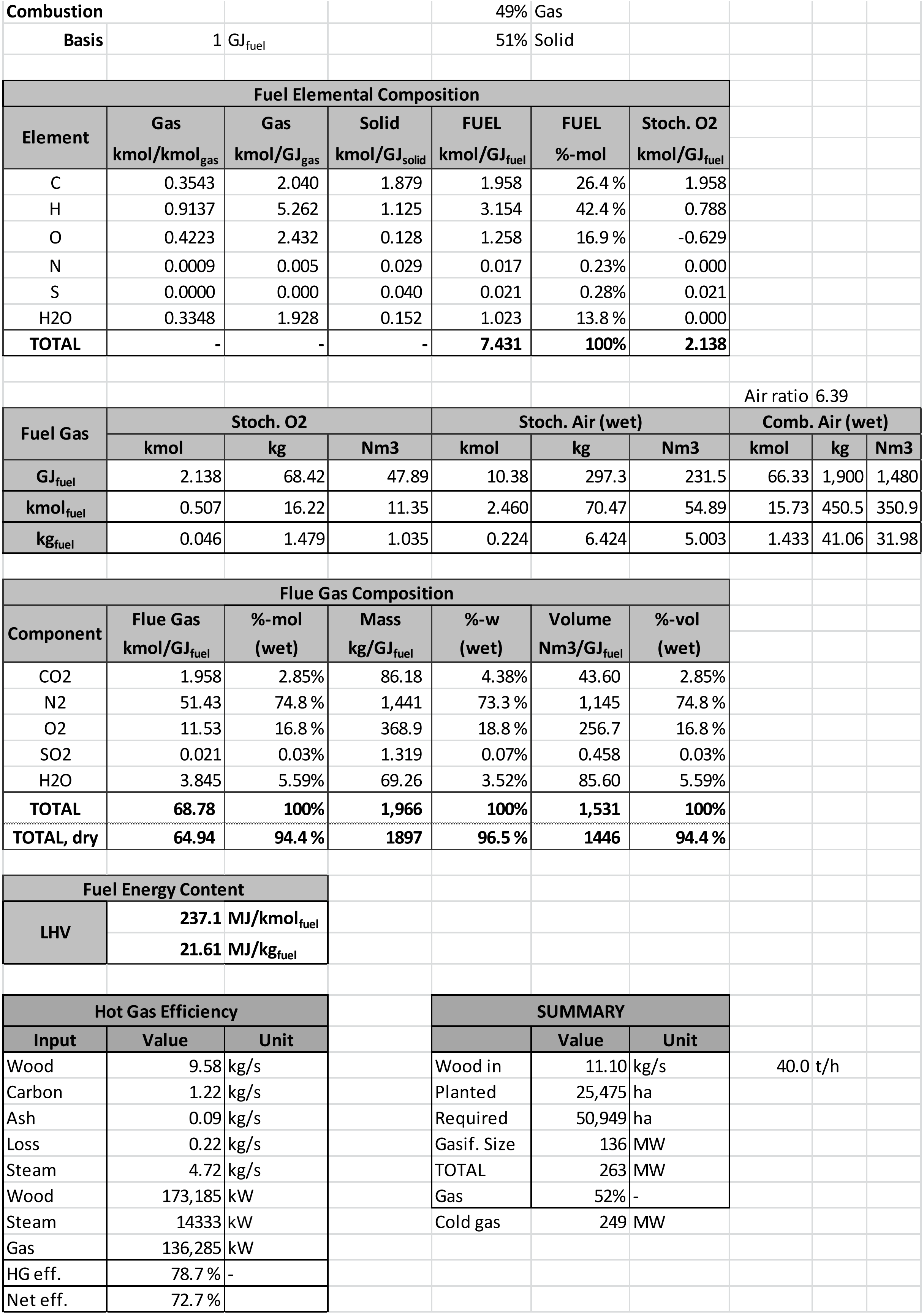
Eucalyptus						
Element	**%-w (DAF)**	**%-w (wet)**	**Amount kmol/kg$_{wood}$**	**%-mol (wet)**	**Steam kmol/kg$_{wood}$**	**Total kmol/kg$_{wood}$**
C	50.5 %	43.9 %	0.03656	30.6 %	0.00000	0.0366
H	6.05%	5.26%	0.05218	43.6 %	0.00000	0.0522
O	43.3 %	37.7 %	0.02356	19.7 %	0.00000	0.0236
N	0.101 %	0.0877 %	0.00006	0.0523 %	0.00000	6.26E-05
H2O	15.0 %	13.0 %	0.00724	6.05%	0.02468	0.0319
TOTAL	**115.0 %**	**100.0 %**	**0.1196**	**100.0 %**	**0.02468**	**0.1443**
TOTAL, dry	**100.0 %**	-	**0.1124**	-	-	-

Product Gas Composition							
Component	**Amount kmol/kg$_{wood}$**	**%-mol (wet)**	**%-mol (dry)**	**Mass kg/kmol$_{gas}$**	**%-w (wet)**	**Volume Nm3/kmol$_{gas}$**	**%-vol (wet)**
CH4	0.00536	7.32%	11.0 %	1.174	6.53%	1.638	7.85%
CO	0.0103	14.0 %	21.0 %	3.919	21.8 %	3.135	15.0 %
CO2	0.0103	14.1 %	21.2 %	6.215	34.6 %	3.144	15.1 %
H2	0.0228	31.1 %	46.7 %	0.6260	3.48%	6.971	33.4 %
N2	0.0000	0.04%	0.0642 %	0.0120	0.0665 %	0.0096	0.0458 %
H2O	0.0245	33.5 %	50.3 %	6.031	33.5 %	5.971	28.6 %
TOTAL	**0.0733**	**100.0%**	**150.3 %**	**17.98**	**100.0 %**	**20.87**	**100.0 %**
TOTAL, dry	**0.0487**	**66.5%**	**100.0%**	**11.95**	-	**14.90**	-

Gas Energy Content		
CH4	803	kJ/mol
H2	242	kJ/mol
CO	284	kJ/mol
TOTAL	173.6	kJ/mol
LHV	**173.6**	**MJ/kmol$_{gas}$**
	9.66	**MJ/kg$_{gas}$**
	8.32	**MJ/Nm3$_{gas}$**

CH4	0.005362	kmol/kg$_{wood}$
	0.000000	

13. Steam gasification fuel combustion:



Combustion				49%	Gas
Basis	1	GJ$_{fuel}$		51%	Solid

Fuel Elemental Composition

Element	Gas kmol/kmol$_{gas}$	Gas kmol/GJ$_{gas}$	Solid kmol/GJ$_{solid}$	FUEL kmol/GJ$_{fuel}$	FUEL %-mol	Stoch. O2 kmol/GJ$_{fuel}$
C	0.3543	2.040	1.879	1.958	26.4 %	1.958
H	0.9137	5.262	1.125	3.154	42.4 %	0.788
O	0.4223	2.432	0.128	1.258	16.9 %	-0.629
N	0.0009	0.005	0.029	0.017	0.23%	0.000
S	0.0000	0.000	0.040	0.021	0.28%	0.021
H2O	0.3348	1.928	0.152	1.023	13.8 %	0.000
TOTAL	**-**	**-**	**-**	**7.431**	**100%**	**2.138**

Air ratio 6.39

Fuel Gas	Stoch. O2			Stoch. Air (wet)			Comb. Air (wet)		
	kmol	kg	Nm3	kmol	kg	Nm3	kmol	kg	Nm3
GJ$_{fuel}$	2.138	68.42	47.89	10.38	297.3	231.5	66.33	1,900	1,480
kmol$_{fuel}$	0.507	16.22	11.35	2.460	70.47	54.89	15.73	450.5	350.9
kg$_{fuel}$	0.046	1.479	1.035	0.224	6.424	5.003	1.433	41.06	31.98

Flue Gas Composition

Component	Flue Gas kmol/GJ$_{fuel}$	%-mol (wet)	Mass kg/GJ$_{fuel}$	%-w (wet)	Volume Nm3/GJ$_{fuel}$	%-vol (wet)
CO2	1.958	2.85%	86.18	4.38%	43.60	2.85%
N2	51.43	74.8 %	1,441	73.3 %	1,145	74.8 %
O2	11.53	16.8 %	368.9	18.8 %	256.7	16.8 %
SO2	0.021	0.03%	1.319	0.07%	0.458	0.03%
H2O	3.845	5.59%	69.26	3.52%	85.60	5.59%
TOTAL	**68.78**	**100%**	**1,966**	**100%**	**1,531**	**100%**
TOTAL, dry	**64.94**	**94.4 %**	**1897**	**96.5 %**	**1446**	**94.4 %**

Fuel Energy Content

LHV	**237.1**	**MJ/kmol$_{fuel}$**
	21.61	**MJ/kg$_{fuel}$**

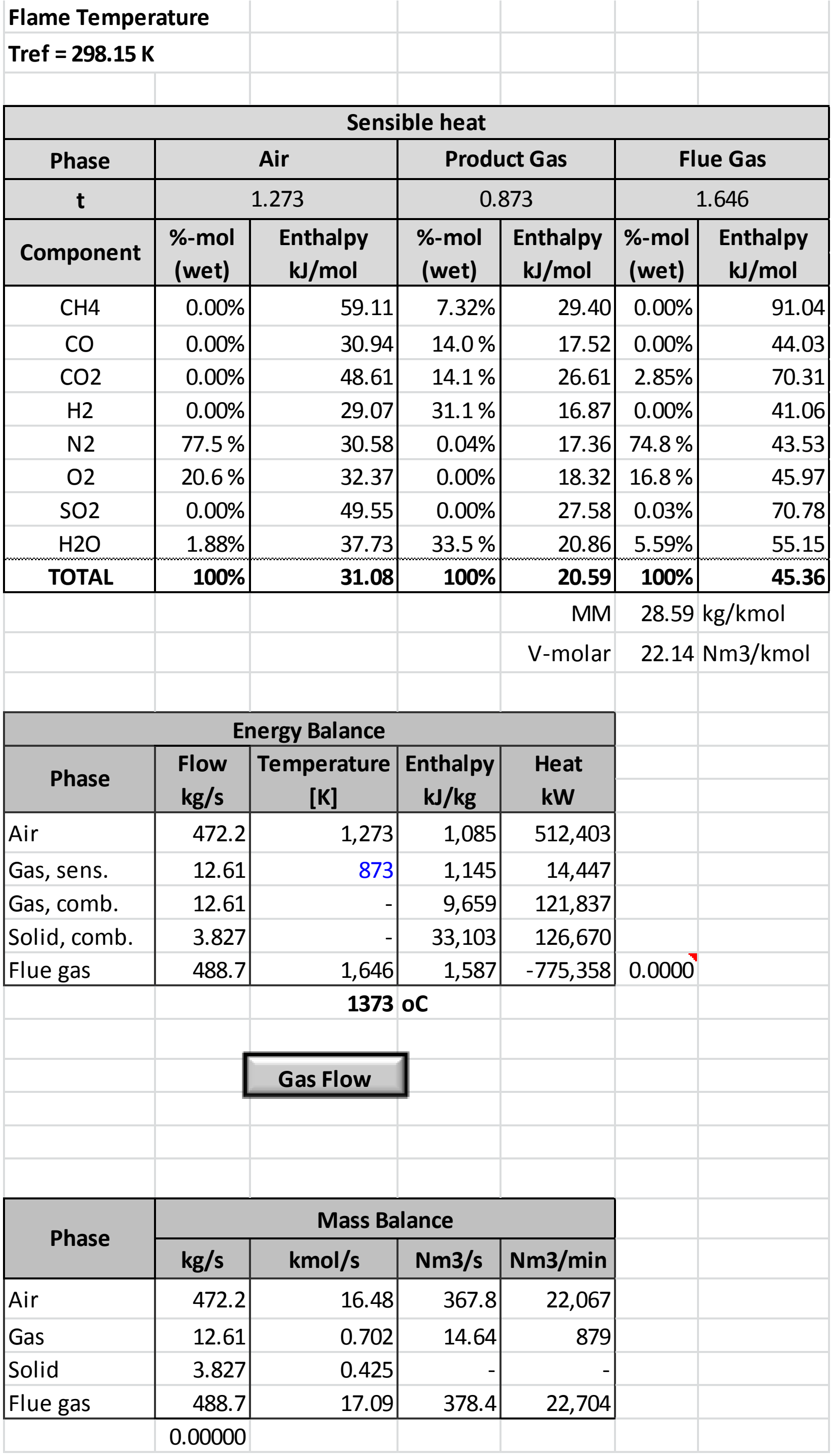
Hot Gas Efficiency

Input	Value	Unit
Wood	9.58	kg/s
Carbon	1.22	kg/s
Ash	0.09	kg/s
Loss	0.22	kg/s
Steam	4.72	kg/s
Wood	173,185	kW
Steam	14333	kW
Gas	136,285	kW
HG eff.	78.7 %	-
Net eff.	72.7 %	

SUMMARY

	Value	Unit		
Wood in	11.10	kg/s	40.0	t/h
Planted	25,475	ha		
Required	50,949	ha		
Gasif. Size	136	MW		
TOTAL	263	MW		
Gas	52%	-		
Cold gas	249	MW		

14. Steam gasification flame temperature:



Flame Temperature

Tref = 298.15 K

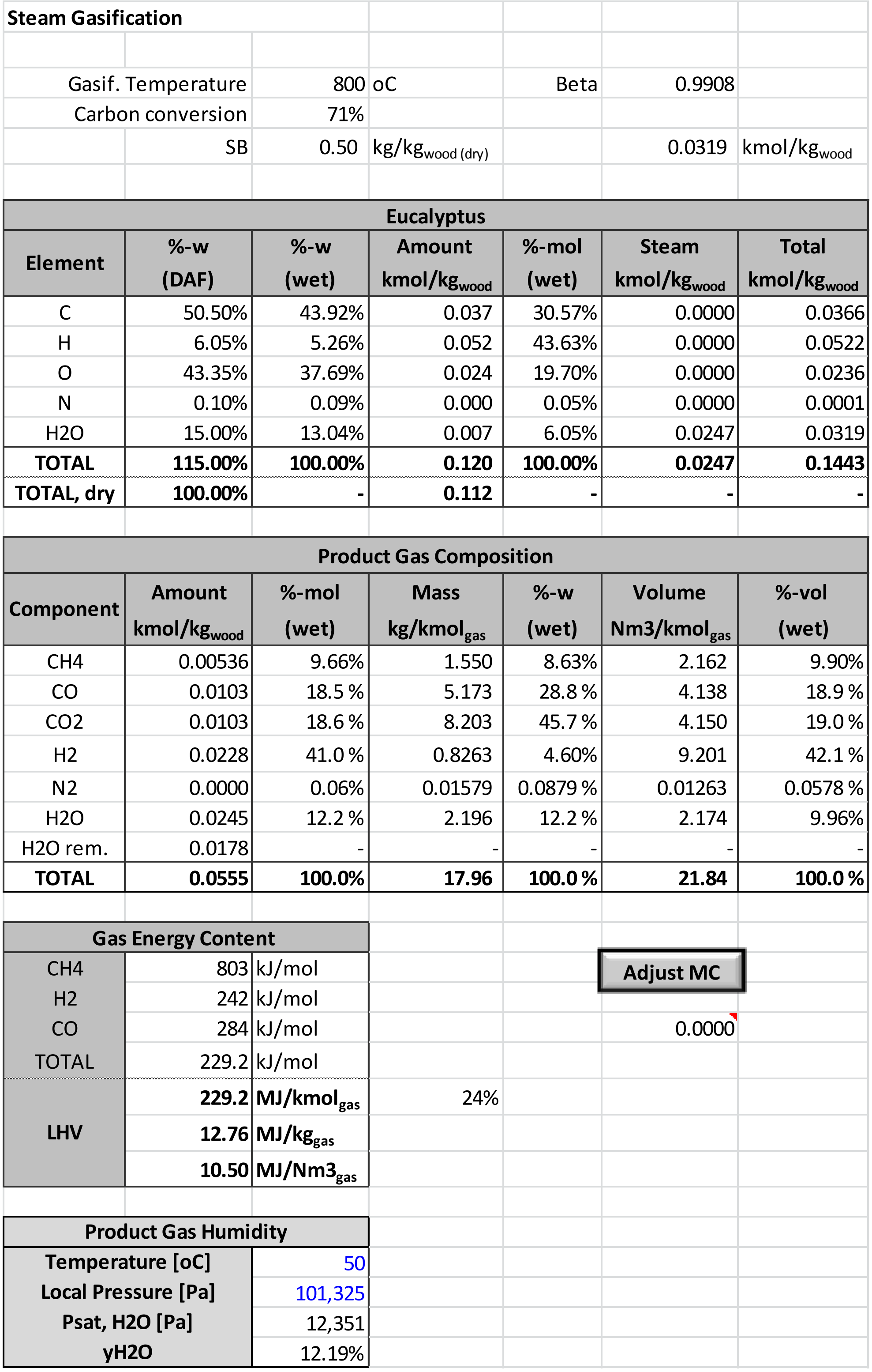
Sensible heat						
Phase	**Air**		**Product Gas**		**Flue Gas**	
t	1.273		0.873		1.646	
Component	**%-mol (wet)**	**Enthalpy kJ/mol**	**%-mol (wet)**	**Enthalpy kJ/mol**	**%-mol (wet)**	**Enthalpy kJ/mol**
CH4	0.00%	59.11	7.32%	29.40	0.00%	91.04
CO	0.00%	30.94	14.0 %	17.52	0.00%	44.03
CO2	0.00%	48.61	14.1 %	26.61	2.85%	70.31
H2	0.00%	29.07	31.1 %	16.87	0.00%	41.06
N2	77.5 %	30.58	0.04%	17.36	74.8 %	43.53
O2	20.6 %	32.37	0.00%	18.32	16.8 %	45.97
SO2	0.00%	49.55	0.00%	27.58	0.03%	70.78
H2O	1.88%	37.73	33.5 %	20.86	5.59%	55.15
TOTAL	**100%**	**31.08**	**100%**	**20.59**	**100%**	**45.36**
				MM	28.59	kg/kmol
				V-molar	22.14	Nm3/kmol

Energy Balance					
Phase	**Flow kg/s**	**Temperature [K]**	**Enthalpy kJ/kg**	**Heat kW**	
Air	472.2	1,273	1,085	512,403	
Gas, sens.	12.61	873	1,145	14,447	
Gas, comb.	12.61	-	9,659	121,837	
Solid, comb.	3.827	-	33,103	126,670	
Flue gas	488.7	1,646	1,587	-775,358	0.0000
		1373	**oC**		

Gas Flow

Phase	**Mass Balance**			
	kg/s	**kmol/s**	**Nm3/s**	**Nm3/min**
Air	472.2	16.48	367.8	22,067
Gas	12.61	0.702	14.64	879
Solid	3.827	0.425	-	-
Flue gas	488.7	17.09	378.4	22,704
	0.00000			

15. Steam gasification dry product gas composition:



Steam Gasification

Gasif. Temperature	800	oC	Beta	0.9908
Carbon conversion	71%			
SB	0.50	kg/$kg_{wood\ (dry)}$		0.0319 kmol/kg_{wood}

Eucalyptus

Element	%-w (DAF)	%-w (wet)	Amount kmol/kg_{wood}	%-mol (wet)	Steam kmol/kg_{wood}	Total kmol/kg_{wood}
C	50.50%	43.92%	0.037	30.57%	0.0000	0.0366
H	6.05%	5.26%	0.052	43.63%	0.0000	0.0522
O	43.35%	37.69%	0.024	19.70%	0.0000	0.0236
N	0.10%	0.09%	0.000	0.05%	0.0000	0.0001
H2O	15.00%	13.04%	0.007	6.05%	0.0247	0.0319
TOTAL	**115.00%**	**100.00%**	**0.120**	**100.00%**	**0.0247**	**0.1443**
TOTAL, dry	**100.00%**	**-**	**0.112**	**-**	**-**	**-**

Product Gas Composition

Component	Amount kmol/kg_{wood}	%-mol (wet)	Mass kg/$kmol_{gas}$	%-w (wet)	Volume Nm3/$kmol_{gas}$	%-vol (wet)
CH4	0.00536	9.66%	1.550	8.63%	2.162	9.90%
CO	0.0103	18.5 %	5.173	28.8 %	4.138	18.9 %
CO2	0.0103	18.6 %	8.203	45.7 %	4.150	19.0 %
H2	0.0228	41.0 %	0.8263	4.60%	9.201	42.1 %
N2	0.0000	0.06%	0.01579	0.0879 %	0.01263	0.0578 %
H2O	0.0245	12.2 %	2.196	12.2 %	2.174	9.96%
H2O rem.	0.0178	-	-	-	-	-
TOTAL	**0.0555**	**100.0%**	**17.96**	**100.0 %**	**21.84**	**100.0 %**

Gas Energy Content

CH4	803	kJ/mol	
H2	242	kJ/mol	
CO	284	kJ/mol	
TOTAL	229.2	kJ/mol	
LHV	**229.2**	**MJ/$kmol_{gas}$**	24%
	12.76	**MJ/kg_{gas}**	
	10.50	**MJ/$Nm3_{gas}$**	

0.0000

Product Gas Humidity

Temperature [oC]	50
Local Pressure [Pa]	101,325
Psat, H2O [Pa]	12,351
yH2O	12.19%

16. Steam gasification dry fuel combustion:

Combustion				50%	Gas
Basis	1	GJ$_{fuel}$		50%	Solid

Fuel Elemental Composition						
Element	**Gas kmol/kmol$_{gas}$**	**Gas kmol/GJ$_{gas}$**	**Solid kmol/GJ$_{solid}$**	**FUEL kmol/GJ$_{fuel}$**	**FUEL %-mol**	**Stoch. O2 kmol/GJ$_{fuel}$**
C	0.4677	2.040	1.879	1.960	28.7 %	1.960
H	1.2061	5.262	1.125	3.198	46.9 %	0.7995
O	0.5575	2.432	0.1282	1.283	18.8 %	-0.6413
N	0.0011	0.005	0.0294	0.017	0.3 %	0.0000
S	0.0000	0.000	0.0404	0.020	0.295 %	0.0201
H2O	0.1219	0.532	0.1524	0.343	5.02%	0.0000
TOTAL	**-**	**-**	**-**	**6.82**	**100%**	**2.138**

Air ratio 6.26

Fuel Gas	**Stoch. O2**			**Stoch. Air (wet)**			**Comb. Air (wet)**		
	kmol	**kg**	**Nm3**	**kmol**	**kg**	**Nm3**	**kmol**	**kg**	**Nm3**
GJ$_{fuel}$	2.138	68.42	47.89	10.38	297.3	231.5	64.93	1,860	1,449
kmol$_{fuel}$	0.564	18.03	12.62	2.735	78.35	61.03	17.11	490.3	381.8
kg$_{fuel}$	0.049	1.568	1.097	0.238	6.811	5.305	1.488	42.62	33.19

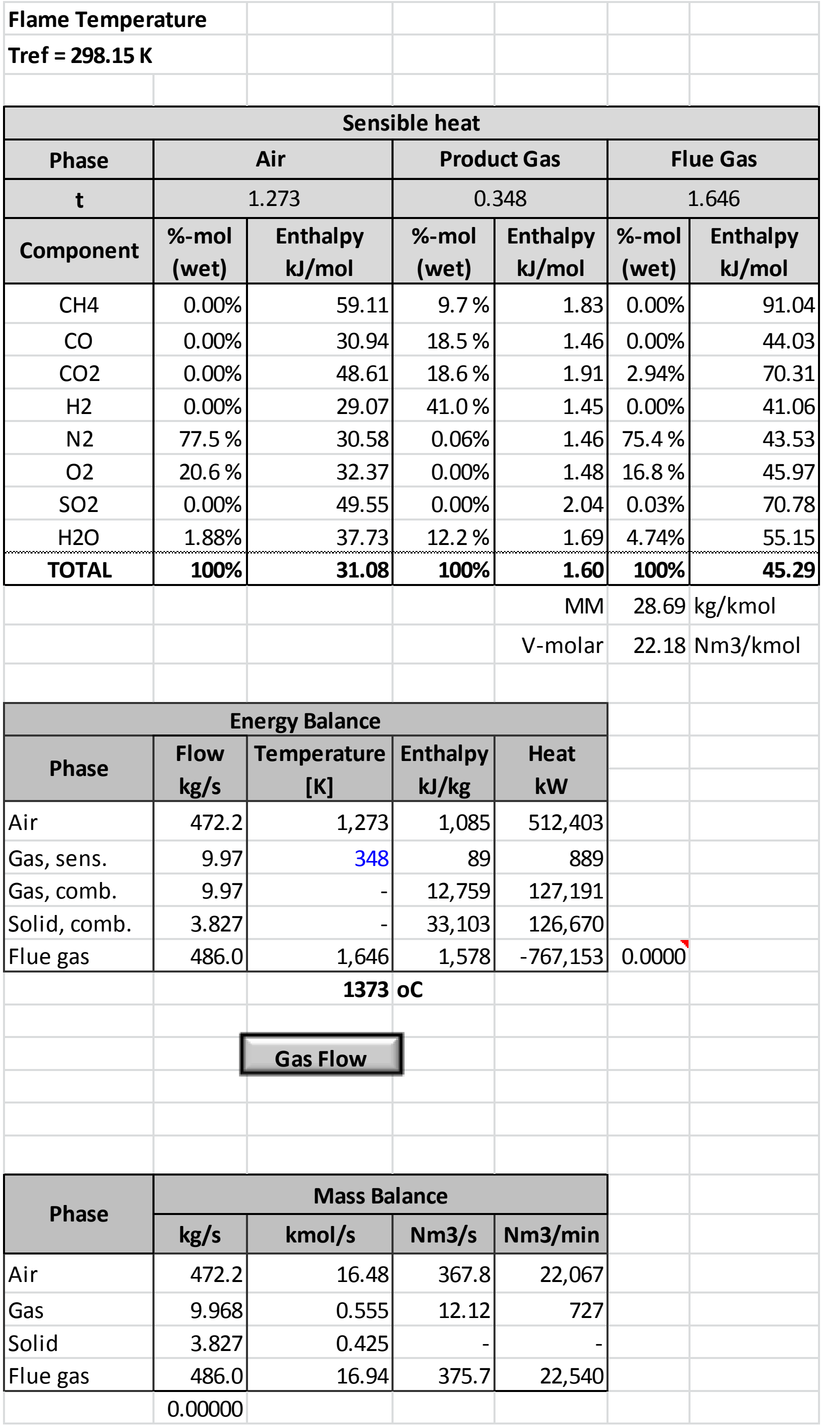
Flue Gas Composition						
Component	**Flue Gas kmol/GJ$_{fuel}$**	**%-mol (wet)**	**Mass kg/GJ$_{fuel}$**	**%-w (wet)**	**Volume Nm3/GJ$_{fuel}$**	**%-vol (wet)**
CO2	1.960	2.94%	86.26	4.51%	43.64	2.94%
N2	50.34	75.4 %	1,410	73.7 %	1,121	75.4 %
O2	11.24	16.8 %	359.7	18.8 %	250.3	16.8 %
SO2	0.020	0.03%	1.291	0.07%	0.449	0.03%
H2O	3.160	4.74%	56.93	2.97%	70.36	4.74%
TOTAL	**66.73**	**100%**	**1,914**	**100%**	**1,486**	**100%**
TOTAL, dry	**63.56**	**95.3 %**	**1858**	**97.0 %**	**1415**	**95.3 %**

Fuel Energy Content		
LHV	**263.6**	**MJ/kmol$_{fuel}$**
	22.91	**MJ/kg$_{fuel}$**

Hot Gas Efficiency		
Input	**Value**	**Unit**
Wood	10.00	kg/s
Carbon	1.27	kg/s
Ash	0.09	kg/s
Loss	0.23	kg/s
Steam	4.92	kg/s
Wood	180,795	kW
Steam	14963	kW
Gas	128,080	kW
HG eff.	70.8 %	-
Net eff.	65.4 %	

SUMMARY				
	Value	**Unit**		
Wood in	11.59	kg/s	41.7	t/h
Planted	26,594	ha		
Required	53,188	ha		
Gasif. Size	128	MW		
TOTAL	255	MW		
Gas	50%	-		
Cold gas	254	MW		

17. Steam gasification (dry) flame temperature:



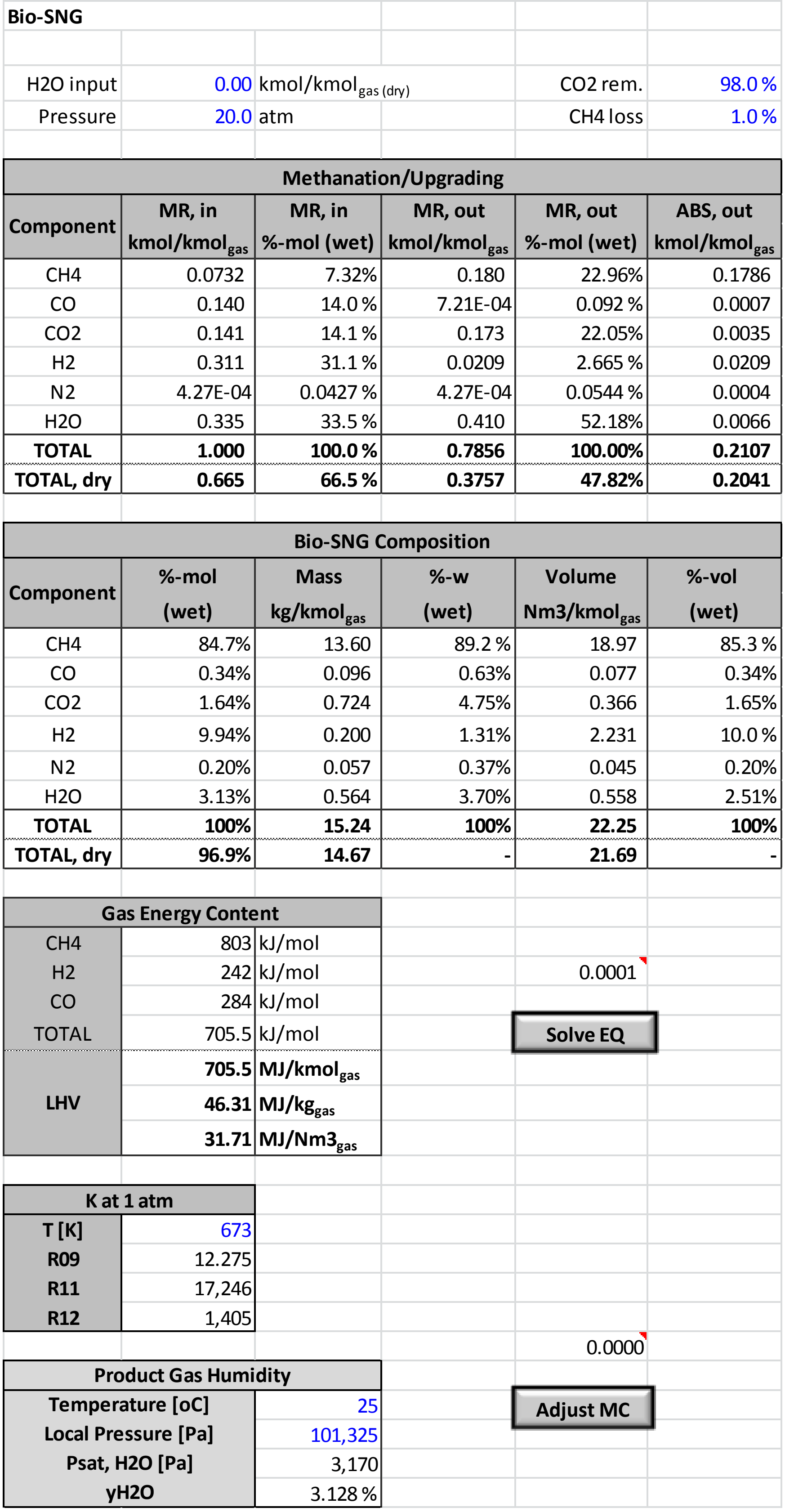
Flame Temperature
Tref = 298.15 K

Sensible heat						
Phase	**Air**		**Product Gas**		**Flue Gas**	
t	1.273		0.348		1.646	
Component	**%-mol (wet)**	**Enthalpy kJ/mol**	**%-mol (wet)**	**Enthalpy kJ/mol**	**%-mol (wet)**	**Enthalpy kJ/mol**
CH4	0.00%	59.11	9.7 %	1.83	0.00%	91.04
CO	0.00%	30.94	18.5 %	1.46	0.00%	44.03
CO2	0.00%	48.61	18.6 %	1.91	2.94%	70.31
H2	0.00%	29.07	41.0 %	1.45	0.00%	41.06
N2	77.5 %	30.58	0.06%	1.46	75.4 %	43.53
O2	20.6 %	32.37	0.00%	1.48	16.8 %	45.97
SO2	0.00%	49.55	0.00%	2.04	0.03%	70.78
H2O	1.88%	37.73	12.2 %	1.69	4.74%	55.15
TOTAL	**100%**	**31.08**	**100%**	**1.60**	**100%**	**45.29**
				MM	28.69	kg/kmol
				V-molar	22.18	Nm3/kmol

Energy Balance					
Phase	**Flow kg/s**	**Temperature [K]**	**Enthalpy kJ/kg**	**Heat kW**	
Air	472.2	1,273	1,085	512,403	
Gas, sens.	9.97	348	89	889	
Gas, comb.	9.97	-	12,759	127,191	
Solid, comb.	3.827	-	33,103	126,670	
Flue gas	486.0	1,646	1,578	-767,153	0.0000
		1373	**oC**		

Phase	**Mass Balance**			
	kg/s	**kmol/s**	**Nm3/s**	**Nm3/min**
Air	472.2	16.48	367.8	22,067
Gas	9.968	0.555	12.12	727
Solid	3.827	0.425	-	-
Flue gas	486.0	16.94	375.7	22,540
	0.00000			

18. Bio-SNG gasification product gas composition:



Bio-SNG

H2O input	0.00	kmol/kmol$_{gas\ (dry)}$	CO2 rem.	98.0 %
Pressure	20.0	atm	CH4 loss	1.0 %

Methanation/Upgrading

Component	MR, in kmol/kmol$_{gas}$	MR, in %-mol (wet)	MR, out kmol/kmol$_{gas}$	MR, out %-mol (wet)	ABS, out kmol/kmol$_{gas}$
CH4	0.0732	7.32%	0.180	22.96%	0.1786
CO	0.140	14.0 %	7.21E-04	0.092 %	0.0007
CO2	0.141	14.1 %	0.173	22.05%	0.0035
H2	0.311	31.1 %	0.0209	2.665 %	0.0209
N2	4.27E-04	0.0427 %	4.27E-04	0.0544 %	0.0004
H2O	0.335	33.5 %	0.410	52.18%	0.0066
TOTAL	**1.000**	**100.0 %**	**0.7856**	**100.00%**	**0.2107**
TOTAL, dry	**0.665**	**66.5 %**	**0.3757**	**47.82%**	**0.2041**

Bio-SNG Composition

Component	%-mol (wet)	Mass kg/kmol$_{gas}$	%-w (wet)	Volume Nm3/kmol$_{gas}$	%-vol (wet)
CH4	84.7%	13.60	89.2 %	18.97	85.3 %
CO	0.34%	0.096	0.63%	0.077	0.34%
CO2	1.64%	0.724	4.75%	0.366	1.65%
H2	9.94%	0.200	1.31%	2.231	10.0 %
N2	0.20%	0.057	0.37%	0.045	0.20%
H2O	3.13%	0.564	3.70%	0.558	2.51%
TOTAL	**100%**	**15.24**	**100%**	**22.25**	**100%**
TOTAL, dry	**96.9%**	**14.67**	**-**	**21.69**	**-**

Gas Energy Content

CH4	803	kJ/mol
H2	242	kJ/mol
CO	284	kJ/mol
TOTAL	705.5	kJ/mol
LHV	**705.5**	**MJ/kmol$_{gas}$**
	46.31	**MJ/kg$_{gas}$**
	31.71	**MJ/Nm3$_{gas}$**

0.0001

K at 1 atm

T [K]	673
R09	12.275
R11	17,246
R12	1,405

0.0000

Product Gas Humidity

Temperature [oC]	25
Local Pressure [Pa]	101,325
Psat, H2O [Pa]	3,170
yH2O	3.128 %

19. Bio-SNG gasification fuel combustion:

Combustion			47%	Gas
Basis	1	GJ$_{fuel}$	53%	Solid

Fuel Elemental Composition						
Element	**Gas kmol/kmol$_{gas}$**	**Gas kmol/GJ$_{gas}$**	**Solid kmol/GJ$_{solid}$**	**FUEL kmol/GJ$_{fuel}$**	**FUEL %-mol**	**Stoch. O2 kmol/GJ$_{fuel}$**
C	0.867	1.229	1.879	1.574	32.8 %	1.574
H	3.589	5.086	1.125	2.988	62.3 %	0.747
O	0.036	0.051	0.128	0.092	1.92%	-0.046
N	0.004	0.006	0.029	0.018	0.38%	0.000
S	0.000	0.000	0.040	0.021	0.45%	0.021
H2O	0.031	0.044	0.152	0.102	2.12%	0.000
TOTAL	**-**	**-**	**-**	**4.795**	**100%**	**2.296**

Air ratio 6.19

Fuel Gas	**Stoch. O2**			**Stoch. Air (wet)**			**Comb. Air (wet)**		
	kmol	**kg**	**Nm3**	**kmol**	**kg**	**Nm3**	**kmol**	**kg**	**Nm3**
GJ$_{fuel}$	2.296	73.47	51.42	11.14	319.2	248.6	68.93	1,975	1,538
kmol$_{fuel}$	1.124	35.98	25.18	5.457	156.3	121.7	33.75	966.9	753.1
kg$_{fuel}$	0.090	2.888	2.021	0.438	12.55	9.77	2.710	77.63	60.46

Flue Gas Composition						
Component	**Flue Gas kmol/GJ$_{fuel}$**	**%-mol (wet)**	**Mass kg/GJ$_{fuel}$**	**%-w (wet)**	**Volume Nm3/GJ$_{fuel}$**	**%-vol (wet)**
CO2	1.574	2.25%	69.25	3.46%	35.03	2.26%
N2	53.44	76.5 %	1,497	74.8 %	1,197	77.2 %
O2	11.91	17.1 %	381.0	19.0 %	266.6	17.2 %
SO2	0.021	0.03%	1.370	0.07%	0.468	0.03%
H2O	2.889	4.14%	52.06	2.60%	51.54	3.32%
TOTAL	**69.83**	**100%**	**2,001**	**100%**	**1,551**	**100%**
TOTAL, dry	**66.94**	**95.9 %**	**1949**	**97.4 %**	**1499**	**96.7 %**

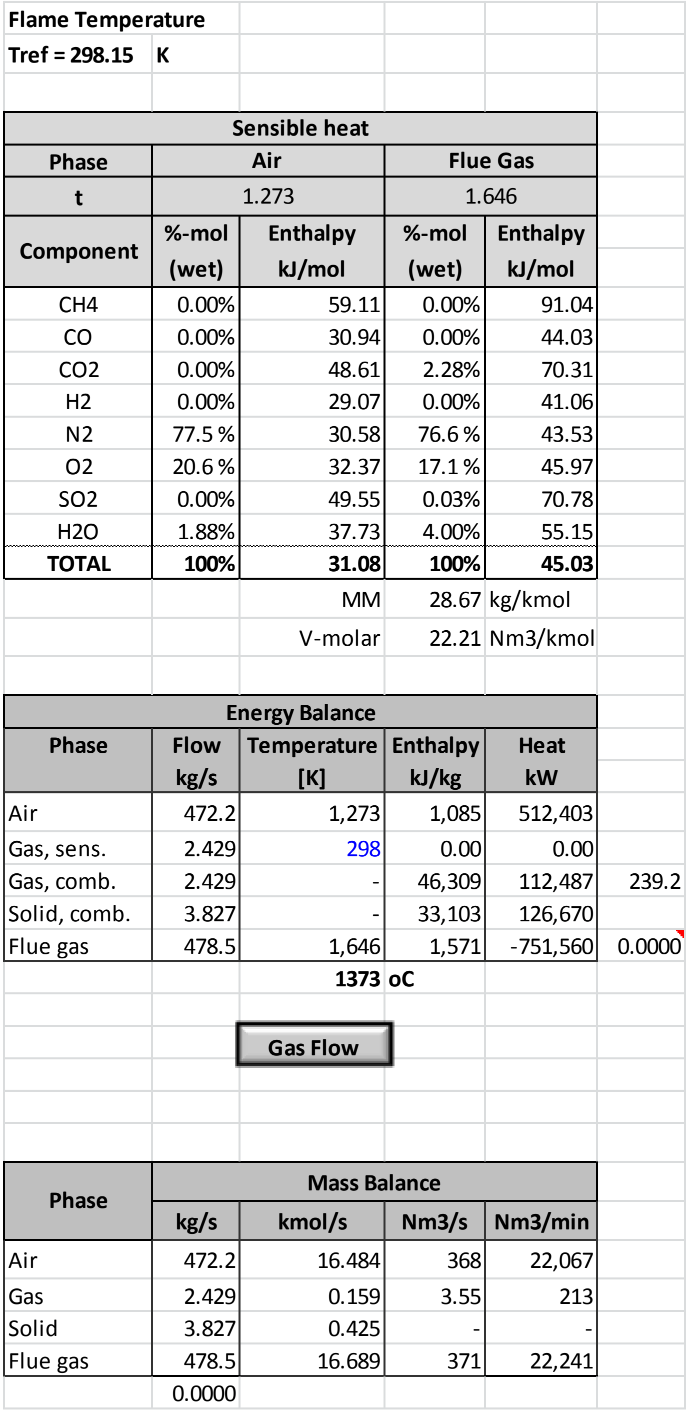
Fuel Energy Content		
LHV	**489.7**	**MJ/kmol$_{fuel}$**
	39.31	**MJ/kg$_{fuel}$**

Hot Gas Efficiency		
Input	**Value**	**Unit**
PG	0.757	kmol PG/s
Wood	10.33	kg/s
Carbon	1.32	kg/s
Ash	0.09	kg/s
Loss	0.23	kg/s
Steam	5.09	kg/s
Wood	186,763	kW
Steam	15457	kW
Gas	112,487	kW
HG eff.	60.2 %	-
Net eff.	55.6 %	

SUMMARY				
	Value	**Unit**		
Wood in	12.0	kg/s	43.1	t/h
Planted	27,472	ha		
Required	54,944	ha		
Gasif. Size	112	MW		
TOTAL	239	MW		
Gas	47%	-		
Cold gas	239	MW		

131 MW

20. Bio-SNG gasification flame temperature:



Flame Temperature

Tref = 298.15 K

Sensible heat				
Phase	**Air**		**Flue Gas**	
t	1.273		1.646	
Component	**%-mol (wet)**	**Enthalpy kJ/mol**	**%-mol (wet)**	**Enthalpy kJ/mol**
CH4	0.00%	59.11	0.00%	91.04
CO	0.00%	30.94	0.00%	44.03
CO2	0.00%	48.61	2.28%	70.31
H2	0.00%	29.07	0.00%	41.06
N2	77.5 %	30.58	76.6 %	43.53
O2	20.6 %	32.37	17.1 %	45.97
SO2	0.00%	49.55	0.03%	70.78
H2O	1.88%	37.73	4.00%	55.15
TOTAL	**100%**	**31.08**	**100%**	**45.03**

MM 28.67 kg/kmol

V-molar 22.21 Nm3/kmol

Energy Balance					
Phase	**Flow kg/s**	**Temperature [K]**	**Enthalpy kJ/kg**	**Heat kW**	
Air	472.2	1,273	1,085	512,403	
Gas, sens.	2.429	298	0.00	0.00	
Gas, comb.	2.429	-	46,309	112,487	239.2
Solid, comb.	3.827	-	33,103	126,670	
Flue gas	478.5	1,646	1,571	-751,560	0.0000
		1373	**oC**		

Gas Flow

Phase	**Mass Balance**			
	kg/s	**kmol/s**	**Nm3/s**	**Nm3/min**
Air	472.2	16.484	368	22,067
Gas	2.429	0.159	3.55	213
Solid	3.827	0.425	-	-
Flue gas	478.5	16.689	371	22,241
	0.0000			