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OLGA Optimum

Improving the economics of integrated biomass gasification plants by extension of the functionalities of the OLGA tar washer

Non-confidential version

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Preface

The work described in this report was carried out within the framework of the project "OLGA Optimum - Improvement of the economy of integrated biomass gasification plants by extension of the functionalities of the OLGA tar washer" that was partly financed by SenterNovem (formerly: Novem) within the framework of the DEN-programme under project number 2020-03-12-14-005. Project duration was from 1 January 2004 till 1 June 2006. Partners in the project were the Energy research Centre of the Netherlands (ECN), Dahlman Industrial Group, Foster Wheeler Energia (FWE), and Technical University of Vienna (TUV). Applicable ECN project number was 7.5264.

Justification

This report comprises the final report of the project focussing on the system and economic assessment results. The results of the experimental activities are described in a technical appendix that is published as a separate report titled "*OLGA Optimum - Combined tar and dust removal in OLGA unit*".

Abstract

The main technical challenge in the implementation of integrated biomass gasification plants is the removal of tar from the product gas, *e.g.* with the OLGA tar removal technology. OLGA requires cooling of the gas to below 320-340°C and de-dusting of the product gas. The economy of gasification plants can be improved when the OLGA is suitable for combined tar and dust removal (*i.e.* then the hot gas filter is superfluous) and when the risk of tar fouling is minimised. In this study the impact of OLGA extensions addressing the system efficiency and economic potential were assessed for eight selected systems, based on circulating fluidised bed (CFB) or fast internally circulating fluidised bed (FICFB) gasification. The use of an OLGA for combined dust and tar removal increases the system efficiency, while changing the cooler temperature has no efficiency effect (although it is desired for operational point of view). The total fuel utilisation in the CFB systems is more than 80%, while the standard FICFB case has a 4%-point lower utilisation. Installation of an OLGA unit to replace the standard FICFB tar scrubber has little effect. However, the OLGA allows the reduction of the steam consumption in the gasifier and this results in an increase of the fuel utilisation to 79%. The IRR of the FICFB reference case is 8.1%; the economic viability can be increased to 8.8% when an OLGA unit and an air-cooled fire-tube cooler are installed. The highest IRR of 9.8% is realised in the system based on a CFB gasifier with an OLGA (and fire-tube product gas cooler). The cost-effectiveness of integrated biomass gasification CHP plants can be improved when an OLGA unit for combined tar and dust removal is part of the system.

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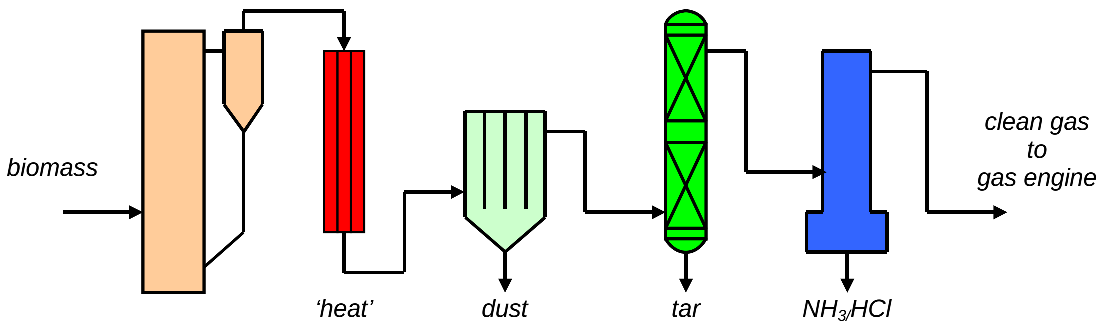
Summary

The main technical challenge in the implementation of integrated biomass gasification plants has been, and still is, the removal of tar from the product gas. “Tar” is equivalent to a major economic penalty in biomass gasification. Tar aerosols and deposits lead to more frequent maintenance and resultantly decrease of revenues, or alternatively, to higher investments. Furthermore, removal of tar components from the process wastewater requires considerable investments. Early 2001 the development of the “OLGA” was initiated at ECN. The patented OLGA is based on applying an organic scrubbing liquid (*i.e.* “OLGA” is the Dutch acronym for oil-based gas washer). The advantages of the OLGA tar removal technology, compared to alternative conventional tar removal approaches, can be summarised as:

- Tar dewpoint of clean product gas is below temperature of application, therefore there is no condensation of tars in the system;
- No fouling of the system resulting in increased system reliability and higher availability;
- Tars are removed prior to water condensation to prevent pollution of process water;
- Tars can be recycled to gasifier and destructed avoiding the handling of problematic (and expensive) tar waste streams;
- Technology is scalable allowing the application from lab to commercial scales.

It is assumed that the OLGA is operated downstream a high-efficient solids removal step (*e.g.* a hot gas filter). The OLGA gas inlet temperature has to be kept higher than the tar dewpoint, similarly the gas outlet temperature must be higher than the water dewpoint. In the OLGA the product gas is cooled, upon which the liquid tars are collected. Additionally, gaseous tars are absorbed in the scrubbing liquid at the resulting temperature. In the design of the OLGA the liquid tar collection and the gaseous tar absorption are performed in two separate scrubbing columns, *i.e.* the Collector and the Absorber. The cleaned product gas leaving the Absorber is “tar-free” (*i.e.* free of tar related problems) and can be treated further in the water-based gas cleaning, fired in a gas engine, or used for more advanced catalytic applications.

Gasification of biomass to convert the solid biomass in a combustible product gas is the key-step in integrated biomass gasification systems for the production of electricity and heat, in so-called bio-CHP plants. After cooling and cleaning the product gas can be applied in gas engine. See scheme below.



The cost-effectiveness of integrated biomass gasification plants can be improved when the capital and operational costs are reduced and the reliability and the number of operational hours are increased. Identified system improvements are:

1. Making the hot gas filter superfluous
2. Increase the outlet temperature of the cooler to minimise the risk of tar condensation and resulting fouling of the gas cooler.

Both system improvements can be realised by innovative extension of the OLGA functionalities. OLGA is made suitable for the removal of dust simultaneous, and combined with, tar removal, which is the primary task of OLGA. Furthermore, the OLGA gas inlet temperature is increased so that the gas cooling in the fouling-critical temperature range (<400°C) is achieved by direct contact of the gas and the washing liquid.

System definition

The integration possibilities of the OLGA gas cleaning process into various biomass gasification CHP systems were investigated. The focus in this project was upon 20 MW_{th} (total fuel input) bio-CHP systems with a gas engine. Two types of gasifiers were considered for the raw gas production. The base-case is an auto-thermal CFB gasifier, the alternative is the allothermal dual fluidized bed (fast internal circulating fluidized bed, FICFB) gasifier as in operation in Güssing. The following configurations were selected:

- **Case 1 - CFB reference**. Reference case for conventional OLGA: The product gas from the CFB is cooled to 320°C, de-dusted in a high temperature hot gas filter (HGF) with sinter metal candles, tars are removed using the OLGA gas cleaning system to the desired tar dew point, and the gas is further cooled. The cleaned gas is combusted in a gas engine. Additionally, a DENOX-system is used to remove the remaining NO_x from the flue gases of the gas engine.
- **Case 2 - CFB new filter**. The system is similar as Case 1, however, with (cheaper) filter candles from a new ceramic material as under development by Foster Wheeler Energia. The candles can operate at temperatures of 500-700°C, have a pressure drop of 10-15 mbar, and a price similar to Teflon filters.
- **Case 3 - CFB OLGA dust**. The HGF is replaced by a cyclone (lower investment and operational costs), which only removes coarse dust; OLGA removes the fine dust.
- **Case 4 - CFB OLGA dust & cooling**. In addition to the dust-load as in Case 3, the gas inlet temperature of OLGA is increased by decreasing the cooler capacity to 500°C. Cooler fouling by (tar) deposition will be avoided (less operational costs). An OLGA washing liquid has to be selected which is applicable at the higher temperatures.
- **Case 5 - FICFB reference**. The Güssing plant, with the existing gas cleaning (pre-coatised bag house filter and RME-scrubber), is assessed for 20 MW_{th}.
- **Case 6 - FICFB OLGA**. The Güssing gasifier under standard operational conditions, as in Case 5, is assessed with the OLGA gas cleaning as in Case 1: HGF-OLGA.
- **Case 7 - FICFB optimum**. It is assumed, that the installation of an OLGA gas cleaning gives the system more operational degrees of freedom compared to the reference Case 5. Therefore, the operational conditions of the gasifier will be optimised, by changing the steam to fuel ratio.
- **Case 8 - FICFB dust & cooling**. The optimised Güssing gasifier from Case 7 is equipped with OLGA-dust-cooling, as in Case 4.

System assessment

CHP systems based on a circulating fluidised bed (CFB) gasifier with an OLGA for combined dust and tar removal operated downstream a cyclone have on average a 1%-point higher chemical efficiency that the corresponding systems with a hot gas filter. This is caused by higher amounts of fly ash (*i.e.* mainly carbon) and tar that are removed in the OLGA and returned to the gasifier. In these systems the chemical efficiency reaches 78% based on the biomass input (*i.e.* corrected for the recycling of spent scrubbing oil to the gasifier). The extension of the OLGA functionalities improves the efficiency. The other system modification of increasing the inlet temperature to the OLGA gas cleaning system has no positive effect on the chemical efficiency of the system.

Systems based on the FICFB gasification technology have in general a 2.5%-point lower chemical efficiency than the corresponding CFB systems. The main reason for this difference is the high steam consumption (*i.e*. steam to fuel ratio) in the FICFB gasifier. A significant improvement of 2%-points in chemical efficiency of the FICFB system can be achieved if the steam to fuel ratio is lowered. Reducing the steam consumption is possible as the gas inlet temperature of the OLGA is much higher than of the standard RME scrubber (*i.e*. 400 versus 140°C).

The gross electrical efficiencies of the different cases are coupled to the chemical efficiencies and the same trends are observed. Maximum efficiencies are approximately 28.5%. The standard FICFB plant has an electric efficiency of about 27%; the same plant with the OLGA gas cleaning and lower steam consumption has an efficiency of more than 27.5%. The net electric efficiencies are approximately 1%-point lower than the gross electric efficiencies. The heat efficiencies for similar CFB and FICFB systems are similar, *i.e*. approximately 52%.

Summarising it can be concluded that the use of an OLGA for combined dust and tar removal increases the system efficiency, while changing the cooler temperature has no efficiency effect (although it is desired for operational point of view). The total fuel utilisation in the CFB systems is more than 80%, while the standard FICFB case has a 4%-point lower utilisation. Installation of an OLGA unit to replace the RME scrubber has little effect. However, the OLGA allows also, in theory, the reduction of the steam consumption in the gasifier and this results in an increase of the fuel utilisation to 79%.

Economic evaluation

The economic assessment revealed that all the 20 MW_{th} systems are economically not very attractive (*i.e*. IRR below 10%) under the conditions assessed. In the base-case assessment, the FICFB systems are more attractive than the CFB-based systems, in spite of the higher efficiencies of the CFB systems. This results partly from the higher costs for the downstream equipment due to the higher product gas flow, but mainly from the higher costs of both the CFB gasifier and the product gas cooler. The impact of the two latter effects is quite significant.

When a more competitive budget estimate for a CFB gasifier is used, the attractiveness of the CFB cases increases with 2.5%-points. In that systems based on CFB and FICFB are equally attractive, which means that the higher costs for the downstream equipment in the CFB systems are compensated by the higher efficiencies. In the bases-case a relative expensive product gas cooler is selected that is cooled with steam. An alternative is to use a much simpler and cheaper fire-tube cooler that is cooled with air (as in operation and tested for 700 hours downstream the CFB gasifier at ECN). Installing the alternative and cheaper cooler results in an increase of the IRR with 1.9%-point.

Summarising, it can be concluded that the IRR of the FICFB reference case is 8.1%, that the economic viability can be increased to 8.8% when an OLGA unit and an air-cooled fire-tube cooler is installed. However, the highest IRR of 9.8% is realised in the system based on a CFB gasifier with an OLGA (and fire-tube product gas cooler).

Concluding

The cost-effectiveness of integrated biomass gasification CHP plants can be improved when an OLGA unit for combined tar and dust removal is part of the system.

1. Introduction

1.1 Background

Gasification (of coal) is an old technology that today is the key chemical process in almost every major method of energy generation, used in the production of electricity, in refineries, and in a variety of other commercial uses. The statement about the position of coal gasification, however, does not apply for biomass gasification. The main technical challenge in the implementation of integrated biomass gasification plants has been, and still is, the removal of tar from the product gas. "Tar" is equivalent to a major economic penalty in biomass gasification. Tar aerosols and deposits lead to more frequent maintenance and resultantly decrease of revenues, or alternatively, to higher investments. Furthermore, removal of tar components from the process wastewater requires considerable investments. Several measures for tar removal have been studied or are under investigation.

Early 2001 the development of the "OLGA" was initiated at ECN. The patented OLGA is based on applying an organic scrubbing liquid (*i.e*. "OLGA" is the Dutch acronym for oil-based gas washer). In the development of the OLGA tar removal technology, ECN has chosen an approach that concentrates on the behaviour (*i.e*. the properties) of the tar and does not concentrate on the tar content. Hence, a "tar-free" product gas is synonymous to a gas "free of tar related problems". The advantages of the OLGA tar removal technology, compared to alternative conventional tar removal approaches, can be summarised as [1]:

- Tar dewpoint of clean product gas is below temperature of application, therefore there is no condensation of tars in system;
- No fouling of the system resulting in increased system reliability and higher availability;
- Tars are removed prior to water condensation to prevent pollution of process water;
- Tars can be recycled to gasifier and destructed avoiding the handling of problematic (and expensive) tar waste streams;
- Technology is scalable allowing the application from lab to commercial scales.

1.2 OLGA process description

A simplified process flow diagram of the OLGA process is shown in Figure 1.1. It is assumed that the OLGA is operated downstream a high-efficient solids removal step (*e.g*. a hot gas filter). The OLGA gas inlet temperature has to be kept higher than the tar dewpoint, similarly the gas outlet temperature must be higher than the water dewpoint. In the OLGA the product gas is cooled, upon which the liquid tars are collected. Additionally, gaseous tars are absorbed in the scrubbing liquid at the resulting temperature. In the design of the OLGA the liquid tar collection and the gaseous tar absorption are performed in two separate scrubbing columns, *i.e*. the Collector and the Absorber. Although, both processes could be performed in a single scrubber unit, separation in two sections is preferred because of process operation considerations.

The liquid tars are separated from the scrubbing liquid and returned to the gasifier; also a small amount of the scrubbing liquid is bleed and recycled to the gasifier. For the absorption step, scrubbing columns were selected that are interacting with each other in a classical absorption-regeneration mode. The scrubbing liquid from the Absorber with the dissolved tars is regenerated in the Stripper. In case of air-blown gasification, air is used to strip the tar. Subsequently, the air with the stripped tars is used as gasifying medium. The loss of scrubbing liquid in the Stripper by volatilisation is minimised by the use of a condenser.

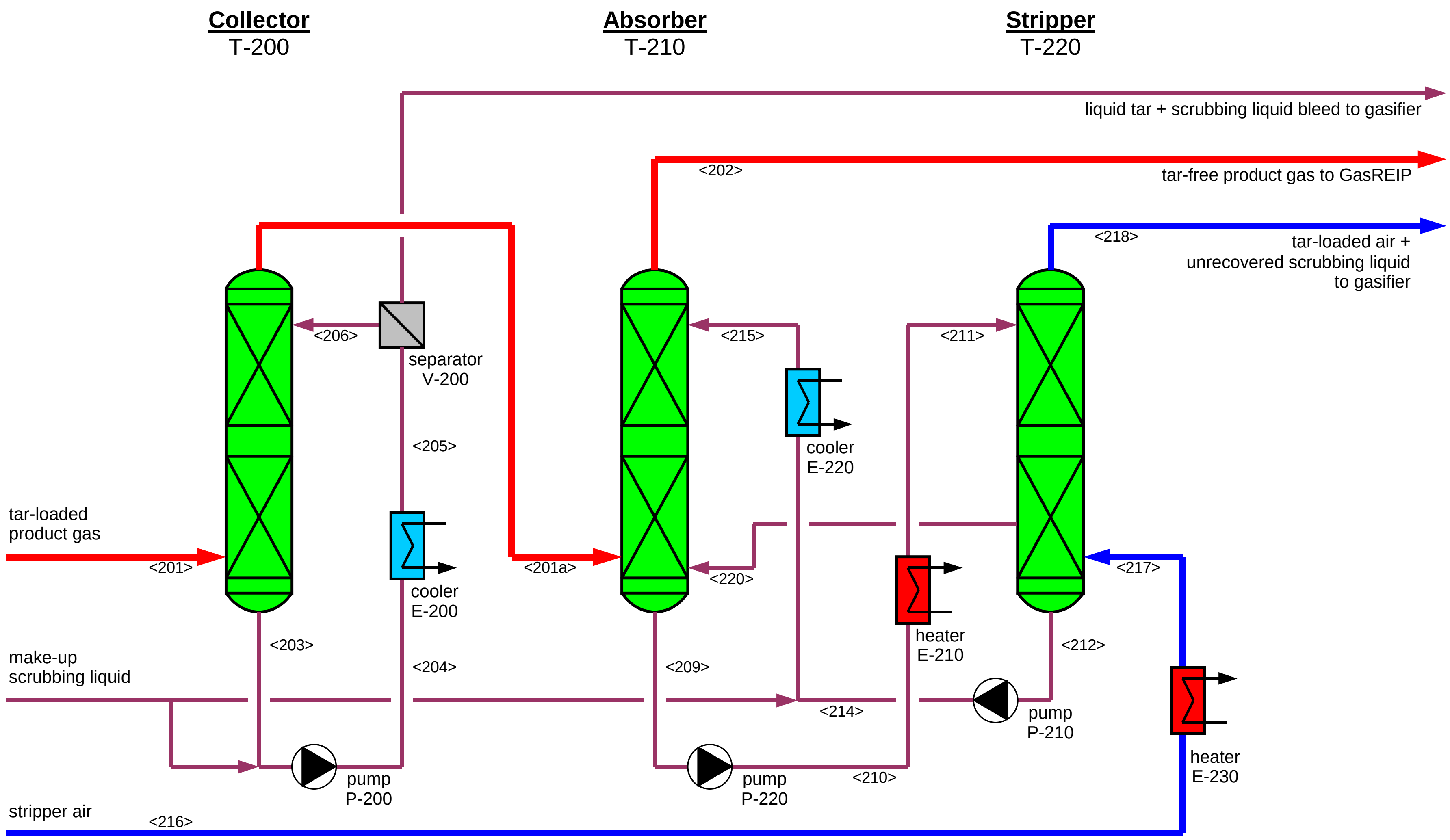


Figure 1.1. Simplified process flow diagram of pilot OLGA unit for tar removal from dust-free gas.

The cleaned product gas leaving the Absorber is "tar-free" (*i.e.* free of tar related problems) and can be treated further in the water-based gas cleaning, fired in a gas engine, or used for more advanced catalytic applications. Typically, tars are removed with >99% efficiency, affording tar dew points below -5°C. Under these conditions, only a part of the one-ring compounds toluene and xylenes remains in the gas. The scope of the OLGA technology is not limited to CHP application or to a specific type of gasifier, *i.e.* OLGA can be operated downstream fixed-bed, CFB, as well as BFB gasifiers and the technology has no fundamental scale limit. The performance is also independent of the tar load in the gas, which was shown by cleaning product gas from oxygen-blown BFB gasification containing ~30 g/m_n^3 of tars. The suitability of the OLGA to meet very stringent tar specifications was proven by successfully using this tar-free gas (after removal of inorganic impurities) as feed for a 500 hours Fischer-Tropsch synthesis run [1] and for the production of Synthetic Natural Gas (SNG).

1.3 Issue definition

Gasification of biomass to convert the solid biomass in a combustible product gas is the key-step in integrated biomass gasification systems for the production of electricity and heat, in so-called bio-CHP plants. After cooling and cleaning the product gas can be applied in a prime mover. A schematic example of such system is shown in Figure 1.2. To make the product gas suitable for a prime mover (*e.g.* gas engine or turbine) the gas must be cooled and all impurities dust, tars, NH_3 and HCl must be removed. This applies to all types of gasification systems: removal of dust, tars, NH_3, and HCl is generic a problem. An important lesson learned from experiences in demonstration and commercial plants (*i.e.* AMER, Güssing, and Lahti) is that gas cooling and removal of dust, tar, and NH_3 are no independent process steps. In an optimally operating and reliable installation all process steps should be carefully tuned.

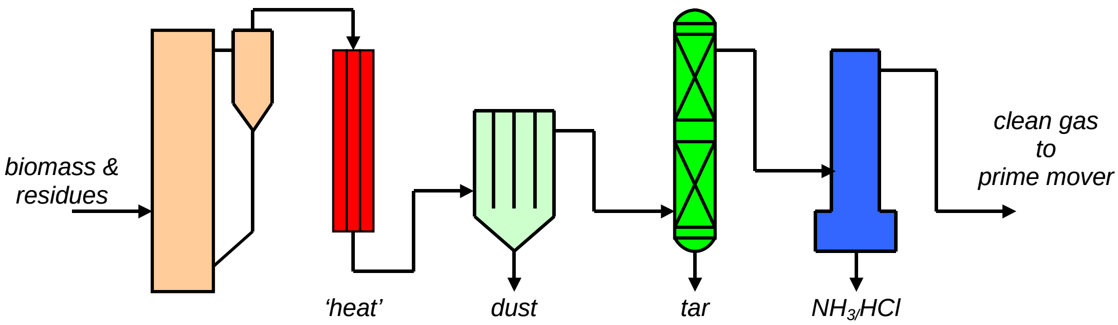


Figure 1.2. Schematic line-up of an integrated biomass gasification CHP plant.

At ECN an integrated pilot CHP system is operated based on a 500 kW_{th} circulating fluidised bed (CFB) gasifier, double pipe gas cooler, hot gas filter, OLGA unit, and a water-based ammonia scrubber. The 'original configuration' of the OLGA unit[a] requires cooling of the gas to below 320-340°C (due to stability of the applied washing liquid) and essentially complete de-dusting of the product gas with the upstream hot gas filter.

The cost-effectiveness of integrated biomass gasification plants can be improved when the capital and operational costs are reduced and the reliability and the number of operational hours are increased. Identified system improvements are:

1. Making the hot gas filter superfluous
2. Increase the outlet temperature of the cooler to minimise the risk of tar condensation and resulting fouling of the gas cooler.

a. The 'original configuration' refers to the pilot plant configuration that was demonstrated in May 2004 to deliver the proof-of-Concept of the OLGA technology [1].

Both system improvements can be realised by innovative extension of the OLGA functionalities.

1.3.1 Dust removal

Although, the system is well designed, and the process steps are well tuned, the relative high capital and operational costs of the hot gas filter are a disadvantage (*viz*. reduce the cost-effectiveness). For indication: cooling and gas cleaning comprise approximately 40% of the capital cost of the integrated system as shown in the Figure above, of which the hot gas filter represents 8 to 10 percent-points. The hot gas filter can be made superfluous when the OLGA can be made suitable for combined dust and tar removal. This results in significant lower investment costs.

1.3.2 Gas cooling

As illustrated above, fouling of the gas coolers and the loss of cooling capacity or blockage is a major problem in operational gasification plants like the AMER and Güssing (as well as it was in ARBRE). Replacement or cleaning of the coolers results in significant reductions of the operational hours and income. In the installations mentioned, fouling starts to significantly take place if the temperature decreases below ~400°C due to tar condensation on the surfaces in the heat exchanger that are (much) colder as they are cooled with water or steam. In all cooling technologies that are based on *indirect* cooling, colder surfaces are present on which condensation may occur. The risk of cooler fouling can be decreased by applying primary measures in the gasifier to reduce the tar content (dew point), as applied in the Güssing plant. Alternatively, the gas may be cooled with a method in which the gas is brought in *direct* contact with the cooling medium. This can be established in the OLGA when the gas inlet temperature can be increased. This results in an increased reliability of the plant resulting in lower operational costs. The OLGA operation will still be generic and applicable downstream all types of gasifiers.

1.4 Objective

Objective of this OLGA development is to extent the functionalities of the OLGA tar removal unit. OLGA will be made suitable for the removal of dust simultaneous, and combined with, tar removal, which is the primary task of OLGA. Furthermore, the OLGA gas inlet temperature will be increased so that the gas cooling in the fouling-critical temperature range (<400°C) is achieved by direct contact of the gas and the washing liquid.

Upon successful extension of the OLGA functionalities, the hot gas filter upstream of OLGA will become superfluous and the gas cooler can be smaller and will have no more fouling problems. This results in a drastic simplification of the line-up of integrated biomass gasification plants, as illustrated in Figure 1.3. Resultantly, this will lead to lower capital as well as operational costs, increased reliability, and resultantly to more cost-effective production of electricity, heat, and/or energy carriers from biomass. Crucial condition for the extension of the functionalities is that primary task of OLGA (*i.e.* tar removal) may not negatively be affected. As criterion is defined that the tar dew point must be below 5°C. Due to the lower capital and operational costs and increased number of operational hours, the cost-effectiveness of the production of green electricity, heat, and/or energy carriers from biomass increases.

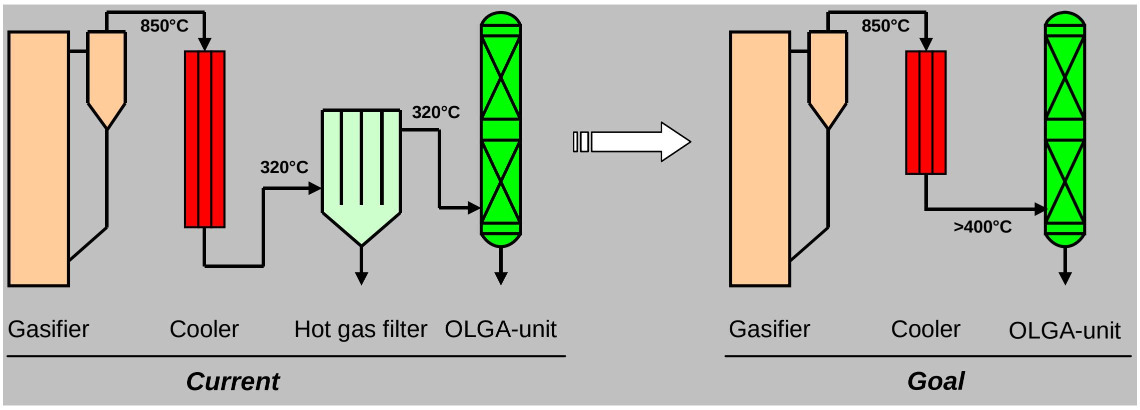


Figure 1.3. Schematic line-up of a part of an integrated biomass gasification CHP plant indicating the process simplification when the OLGA functionalities are extended to include dust removal and partly gas cooling.

1.5 This report

The results of the experimental activities in the OLGA development to extent the functionalities of the OLGA tar removal unit are described in a separate report, *i.e*. “OLGA Optimum - Combined tar and dust removal in OLGA unit”. In this underlying report the results are described of the system and economic assessment to quantify the impact of the OLGA modifications on the cost-effectiveness of selected bio-CHP systems. In Chapter 2 the selected systems are defined and discussed. Chapter 3 describes the results of the system assessment and presents the plant efficiencies. The economic potential of the systems is evaluated in Chapter 4 and in Chapter 5 the conclusions are presented.

2. System definition

2.1 Introduction

The project addresses two innovative extensions of the OLGA functionalities, *i.e.* combined dust and tar removal and increased the OLGA inlet temperature. This could result in a drastic simplification of the line-up of integrated biomass gasification plants. Due to the lower capital and operational costs and increased number of operational hours, the cost-effectiveness of the production of green electricity, heat, and/or energy carriers from biomass could increase.

The focus of the project is to investigate the integration possibilities of the OLGA gas cleaning process into various biomass gasification CHP systems. The so obtained system assessment results from Chapter 3 (*i.e.* mass and energy balances and efficiencies) will be used as basis for the economic assessment in Chapter 4.

2.2 Evaluated systems

In the following the different systems will be defined which will be assessed aimed at the determination of the complete mass and energy balances. The focus in this project is laid upon bio-CHP systems with a gas engine as prime mover. The scale was fixed to 20 MW_{th} (total fuel input). Two types of gasifiers were considered for the raw gas production. The base-case is an auto-thermal CFB gasifier, the alternative is the allothermal dual fluidized bed (fast internal circulating fluidized bed, FICFB) gasifier as in operation in Güssing. The following configurations were selected (Table 2.1).

Table 2.1. Overview about the system configurations.

#	Gasifier	Cooler	Dust	Tar	Remarks
1	CFB	320°C	HGF (candles)	OLGA	CFB reference
2	CFB	320°C	HGF (new type)	OLGA	CFB new filter
3	CFB	320°C	cyclone	OLGA (dust)	CFB OLGA dust
4	CFB	500°C	cyclone	OLGA (dust, T)	CFB OLGA dust & cooling
5	FICFB	180°C	pre-coat filter	RME-scrubber	FICFB reference
6	FICFB	320°C	HGF (new type)	OLGA	FICFB OLGA
7	FICFB	320°C	HGF (new type)	OLGA	FICFB optimum
8	FICFB	500°C	cyclone	OLGA (dust, T)	FICFB dust & cooling

<u>Case 1 - CFB reference</u>
Reference case for conventional OLGA: The product gas from the CFB is cooled to 320°C, de-dusted in a high temperature hot gas filter (HGF) with sinter metal candles, tars are removed using the OLGA gas cleaning system to the desired tar dew point, and the gas is further cooled. The cleaned gas is combusted in a gas engine. Additionally, a DENOX-system is used to remove the remaining NO_x from the flue gases of the gas engine.

<u>Case 2 - CFB new filter</u>
The system is similar as Case 1, however, with (cheaper) filter candles from a new ceramic material as under development by Foster Wheeler Energia. The candles can operate at temperatures of 500-700°C, have a pressure drop of 10-15 mbar, and a price similar to Teflon filters.

Case 3 - CFB OLGA dust

The HGF is replaced by a cyclone (lower investment and operational costs), which only removes coarse dust; OLGA removes the fine dust.

Case 4 - CFB OLGA dust & cooling

In addition to the dust-load as in Case 3, the gas inlet temperature of OLGA is increased by decreasing the cooler capacity to 500°C. Cooler fouling by (tar) deposition will be avoided (less operational costs). An OLGA washing liquid has to be selected which is applicable at the higher temperatures.

Case 5 - FICFB reference

The Güssing plant, with the existing gas cleaning (pre-coatised bag house filter and RME-scrubber), is assessed for 20 MW_{th}.

Case 6 - FICFB OLGA

The Güssing gasifier under standard operational conditions, as in Case 5, is assessed with the OLGA gas cleaning as in Case 1: HGF-OLGA.

Case 7 - FICFB optimum

It is assumed, that the installation of an OLGA gas cleaning gives the system more operational degrees of freedom compared to the reference Case 5. Therefore, the operational conditions of the gasifier will be optimised, by changing the steam to fuel ratio.

Case 8 - FICFB dust & cooling.

The optimised Güssing gasifier from Case 7 is equipped with OLGA-dust-cooling, as in Case 4.

2.3 Biomass composition

For the different simulations a typical clean forest wood biomass with the characteristics given in Table 2.2 is used.

Table 2.2. Biomass feedstock characteristics.

component		Forest wood
ash wf	[wt%]	1.60
C waf	[wt%]	50.80
H waf	[wt%]	6.10
O waf	[wt%]	42.70
N waf	[wt%]	0.30
S waf	[wt%]	0.06
Cl waf	[wt%]	0.04
water content	[wt%]	15
hhv wf	[kJ/kg]	19,845
lhv wf	[kJ/kg]	18,534

3. System assessment

3.1 Methodology

To assess the described concepts technically, it is important to define objective and comparable criteria. Therefore, efficiencies are set up to be able to compare simulation results among the concepts and to other results. To obtain mass and energy balances for the different concepts the process simulation tool IPSEpro was used. Furthermore, it is essential to base a technical assessment on the same parameters to achieve comparable results.

3.1.1 Efficiency definitions

For evaluation several efficiencies are defined, which provide a possibility for the comparison of the different process configurations. However, for a proper comparison these characteristic efficiencies have to be defined precisely. For the evaluation in this work the following characteristic efficiencies are defined:

Definition of electrical efficiencies

The ***gross electrical efficiency*** is calculated as the ratio of the produced electrical power to the fuel power of the feedstock entering the plant (e.g. before a possible fuel preparation).

$$\eta_{el,gross} = \frac{P_{elGE}}{\sum \dot{m}_{Fuel} \cdot lhv_{Fuel}} \qquad \text{Equation 3-1}$$

The net electrical efficiency is calculated as the ratio of the produced electrical power reduced by the electrical consumption of the apparatus to the fuel power of the feedstock entering the plant.

$$\eta_{el,net} = \frac{P_{elGE} - P_{elCons}}{\sum \dot{m}_{Fuel} \cdot lhv_{Fuel}} \qquad \text{Equation 3-2}$$

Definition of the thermal efficiency

The thermal efficiency of a plant is calculated as the ratio of the produced heat (district or/and process heat) to the fuel power of the feedstock entering the plant.

$$\eta_{Q} = \frac{\dot{Q}}{\sum \dot{m}_{Fuel} \cdot lhv_{Fuel}} \qquad \text{Equation 3-3}$$

Definition of the fuel utilisation

The gross fuel utilisation is calculated as the ratio of the produced electrical power and heat (district or process heat) to the fuel power of the feedstock entering the plant.

$$\eta_{fuel,gross} = \frac{P_{elGE} + \dot{Q}}{\sum \dot{m}_{Fuel} \cdot lhv_{Fuel}} \qquad \text{Equation 3-4}$$

The net fuel utilisation is calculated as the ratio of the produced electrical power and heat (district or process heat) reduced by the electrical consumption of the apparatus to the fuel power of the feedstock entering the plant.

$$\eta_{fuel,net} = \frac{P_{elGE} - P_{elCons} + \dot{Q}}{\sum \dot{m}_{Fuel} \cdot lhv_{Fuel}} \qquad \text{Equation 3-5}$$

Definition of the chemical efficiency

Especially for the characterisation of gasification processes the chemical efficiency is a widely used parameter. The chemical efficiency is generally defined as the amount of chemical energy, which can be transferred from fuel into the product gas of a thermo-chemical conversion process expressed as ratio of energy streams:

$$\eta_{chem} = \frac{\dot{m}_{PG} \cdot lhv_{PG}}{\sum \dot{m}_{Fuel} \cdot lhv_{Fuel}} \qquad \text{Equation 3-6}$$

where $\dot{m}_{PG}$ represents the mass flow of the product gas, lhv_{PG} the lower heating value of the product gas, $\dot{m}_{Fuel}$ the mass flow of the fuel into the gasification reactor and lhv_{Fuel} the lower heating value of the feedstock.

Definition of the chemical efficiency of the gasifier

This definition is very well suited for the characterisation of a gasifier. The fuel power of the product gas includes the combustible compounds including tars and higher hydrocarbons, which leave the gasifier. However, this efficiency does not indicate the quality of the gasification process, but only expresses the ratio between the lower heating value of the gaseous compounds at the exit of the gasifier and the fuel power of the feedstock. Therefore, the chemical efficiency of the gasifier does not indicate its usability in the gas utilisation. The term $\dot{m}_{PG} \cdot lhv_{PG}$ in Equation 3-6 refers to the conditions of the product gas at the exit of the gasifier, however, at that stage the quality of the gas is not suitable for direct gas utilisation in most cases. The term $\sum \dot{m}_{Fuel} \cdot lhv_{Fuel}$ refers to the total fuel input into the gasifier. Since the system boundary is set around the gasifier, any additional fuels recycled from the gas cleaning e.g. in form of tars or char have to be considered as external streams too. Figure 3.1 demonstrates the system boundary for the calculation of the chemical efficiency for the gasifier.

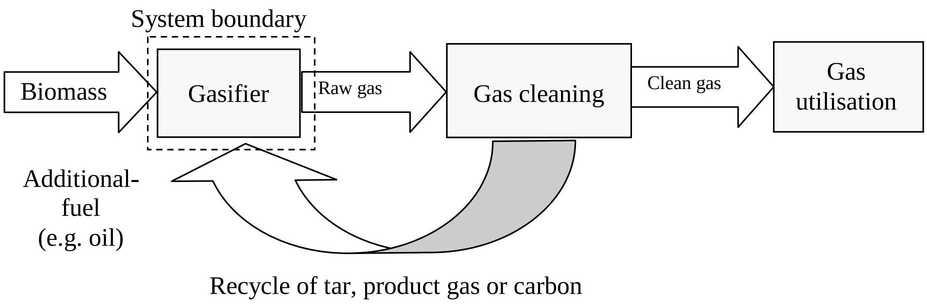


Figure 3.1. *System boundary for the calculation of the chemical efficiency for the gasifier*

Definition of the chemical efficiency of a gasification plant

For the evaluation of gasification plants the definition of the chemical efficiency for the gasifier is not very well suited since the quality of the product gas is neglected. By including the gas cleaning into the system boundary the quality of the product gas can be taken into account in the analysis. In this case the mass flow and the lower heating value of the product gas at the inlet to the gas utilisation are taken. Therefore, only usable combustible compounds are assessed, tars or char are not included anymore. Possible recycle streams from the gas cleaning into the gasifier are treated as internal streams, since they do not leave the system boundary. The system boundary for the calculation of the chemical efficiency of a plant is given in Figure 3.2.

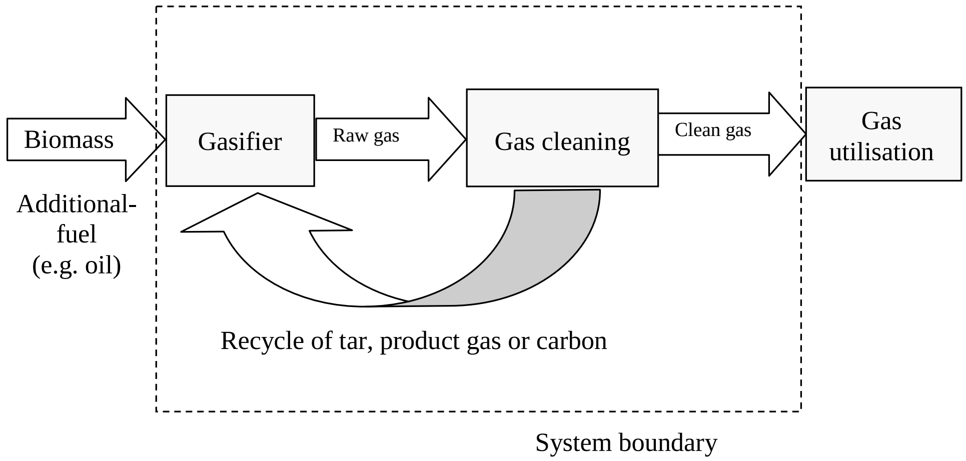


Figure 3.2. System boundary for the calculation of the chemical efficiency for a plant.

3.1.2 Simulation tool IPSEpro

IPSEpro is an equation oriented process simulation environment with a modular structure to offer flexible handling of process units. This process simulation tool solves the modelled process by forming a non-linear equation system, which is solved by a Newton-Raphson-algorithm. An essential advantage of this tool is the modular set up, shown in Figure 3.3.

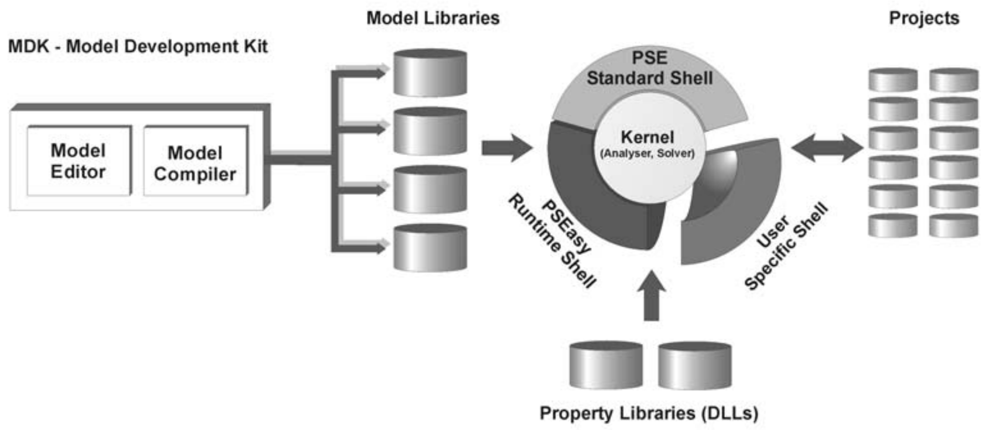


Figure 3.3. *Structure of the simulation environment*

The process simulation environment (PSE) with the equation solver (Kernel) refers to a model library, with the information about the utilised apparatus. This model library can be edited with a special editor called model developer kit (MDK), which allows the implementation of user-defined models. The thermodynamic and physical data for the calculations are provided by external property libraries (DLLs). The standard software package IPSEpro® [2], which is designed to model standard power plant processes, has been greatly enlarged to model and describe gasification processes; dryers, gasifiers and gas cleaning equipment have been implemented, by mass- and energy balances, including possible chemical reactions and empirical correlations from measurements of real gasification plants. A detailed description of the models can be found in references 3 and 4.

3.1.3 General conditions

In the following the ambient conditions, the efficiencies of the specific apparatus and the biomass feedstock are given. All simulation work is based on these conditions. The specific ambient conditions and the general set-up can be found in Table 3.1. The implemented efficiencies for the specific apparatus can be seen in Table 3.2.

Table 3.1. Ambient conditions and general plant data.

Ambient conditions	
temperature	15°C
relative humidity	60 %
ambient pressure	1.013 bar
General set-up	
fuel power (including additional fuels)	20 MW
exit temperature gasifier	830°C
stack temperature	120°C
district heating feed	70°C
district heating drain	100°C
Δp heat exchangers (PG-, flue gas-, air side)	10 mbar

Table 3.2. Efficiencies of the specific apparatus

	η_s	η_m
compressors	0.75	0.99
pumps	0.75	0.99
	η_{el}	η_m
motors, generators	0.98	0.98

The gas engine is not listed in the table since it is a more complex model, which was developed together with Jenbacher. It consists of a gas mixer, a turbo charger, a high and low temperature combustion gas cooling, a throttle, and a combustion chamber.

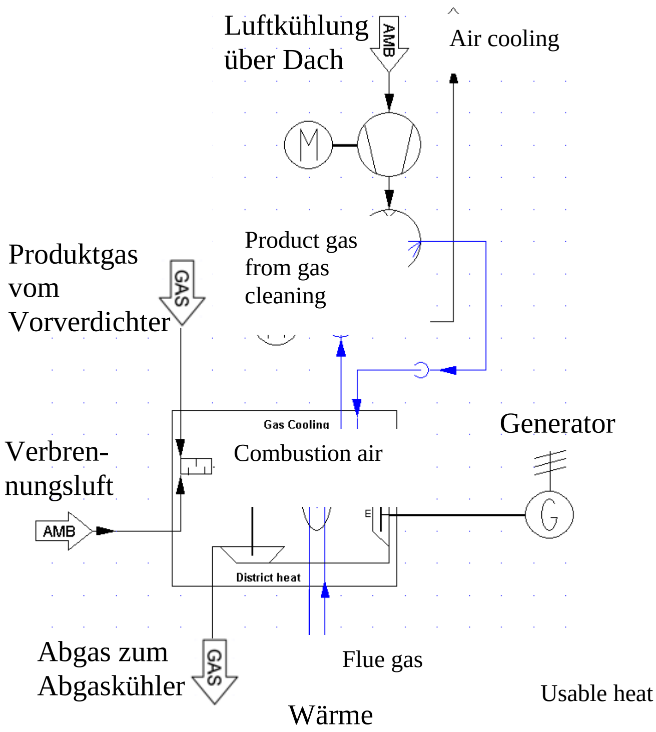


Figure 3.4. Gas engine model.

The gas engine achieves depending on the cylinder pressure and the combustion gas inlet temperature electrical efficiencies of 33 to 37%. In all simulations the gas engines were operated at a lambda of 1.3 to achieve comparable results. As fuel for the different simulations, biomass with the characteristics given in Table 2.2 is used. For the simulation of the CFB cases the carbon content in the ash was set to 42 wt% of the total ash according to laboratory measurements by ECN.

3.2 Implementation of OLGA into IPSEpro

In the following the implementation of the OLGA gas cleaning is explained in detail and how it was included into the existing simulation tool for gasification processes. Figure 3.5 shows the flow sheet of the OLGA gas cleaning implemented into IPSEpro. Product gas enters from the left hand side into the first column in standard operation conditions at about 320°C and is cooled and cleaned from tars by the OLGA liquid to about 85°C. The loaded scrubbing liquid is first cooled to 80°C before it is cleaned from dust and coarse particles in a filter. The clean liquid is then heated up to 180°C and stripped by hot air from the dissolved tars in the second column.

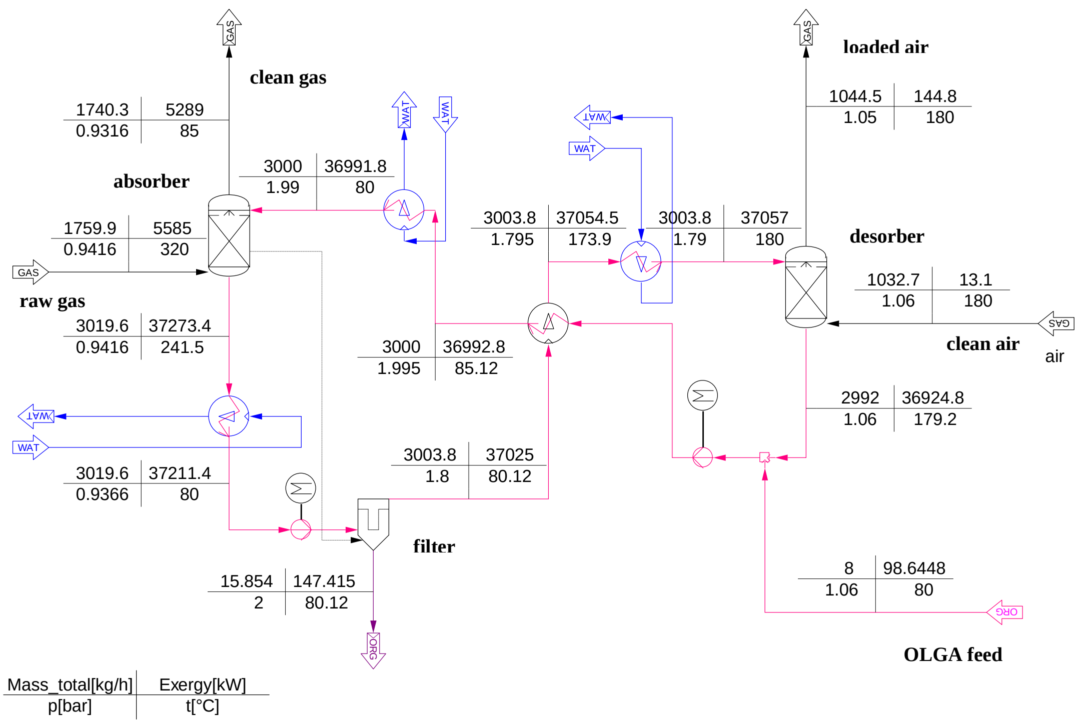


Figure 3.5. Flow sheet of the OLGA gas cleaning in IPSEpro (blue: water/steam; pink: OLGA liquid; black: gasification gas).

The loaded air can be used as combustion air in the gasifier and hence the stripped of tars can be recycled. Measures should be taken to keep the air temperature above the tar dew point. The cleaned liquid leaves the column at the bottom. Possible losses of scrubbing liquid are substituted before the liquid is cooled to 80°C entering again the absorption column. The relative nitrogen consumption of the filter was specified by ECN to 0.250 vol% of the raw gas flow, the specific oil consumption of the process to 4.00 gram of new scrubbing liquid per m_n^3 of raw product gas.

3.3 Results

In the following the results of the technical assessment will be presented, in the order of the different cases 1-8. Finally, a comparison of the different results will be given. Detailed mass and energy balances can be found in the appendix.

3.3.1 Case 1 - CFB reference

Represents a fluidised bed gasification plant with a fuel input of 20 MW thermal; including additional fuels, like the residues of the filter of the OLGA gas cleaning system. Biomass enters the plant (Figure 3.6) with the conditions specified in Table 2.2. Additional to this the residues of the OLGA gas cleaning are added as additional fuel to the gasifier. In the fluidised bed gasifier the fuel is gasified and leaves the reactor with a temperature of 830°C. The product gas is first cooled to 334°C for hot gas filtering of the fly ash and dust. The pre-coated filter will reach a dust separation rate of over 99% and will remove approximately 15% of the tar from the product gas, too. The cooled and dedusted product gas enters the OLGA gas cleaning stage at 320°C, where the remaining tar is removed. The cleaned product gas leaves the absorber column with 85°C and is cooled further to 40°C to reduce the water content before the gas engine. The produced condensate (about 700 kg/h) has to be disposed. The product gas is combusted in three gas engines for heat and power production. The flue gases are cleaned by an oxidation catalyst from remaining CO loads and are used for air preheating of the gasification process.

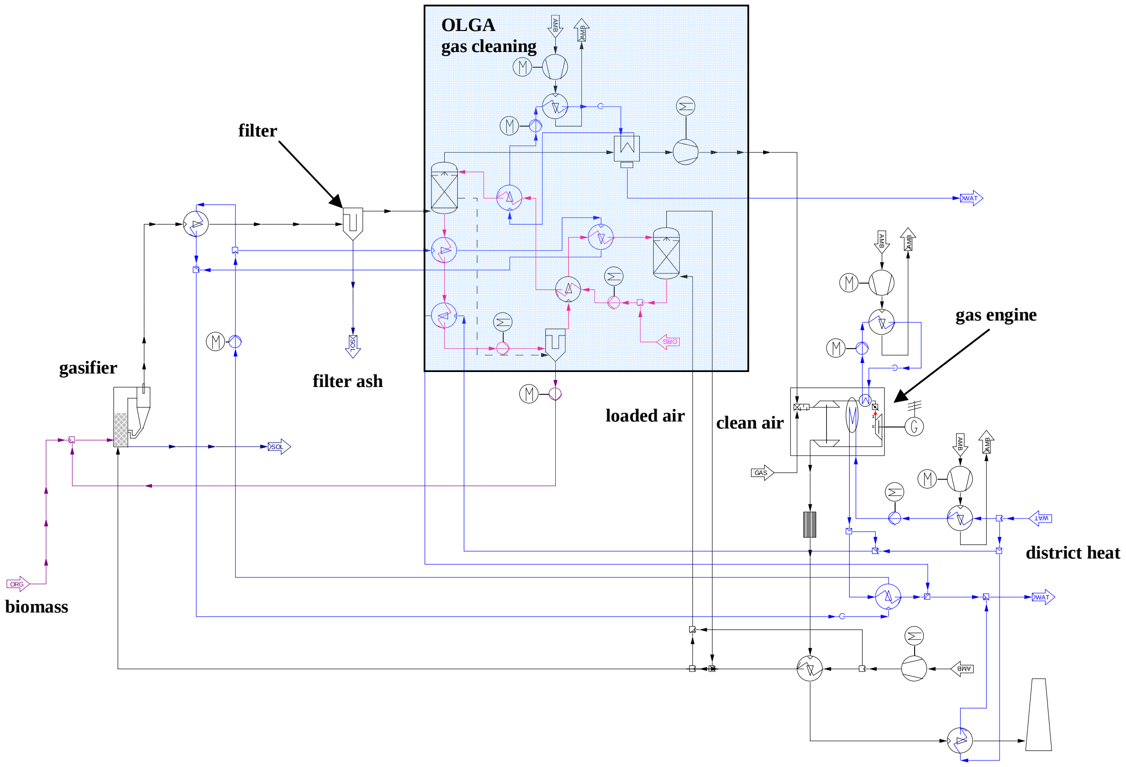


Figure 3.6. Flow chart case 1&2: CFB gasification without ash combustion (blue: water/steam; pink: OLGA liquid; black: gas streams).

The air used for stripping in the OLGA gas cleaning process is heated up to 180°C. The tar-loaded air from the tar stripper is mixed with the remaining air and feed together to the gasifier as fluidisation air. District heat is extracted from the product gas itself, from the high temperature cooling in the OLGA liquid circuit, and from the remaining heat of the flue gases after the air preheating. The cooled flue gas leaves the stack with 120°C. The ash from the gasifier has according to measurements at ECN high carbon loadings of up to 44% of the biomass ash. Table 3.3 shows the energy balance of the overall plant.

Table 3.3. Energy balance overall plant case 1*.

Fuel power without OLGA [kW]	19,524
Fuel power with OLGA [kW]	20,000
Chemical power gasification gas [kW]	15,464
District heating power [kW]	10,217
Electrical power generator [kW]	5,654
Electrical consumption plant [kW]	147
Electrical net power production [kW]	5,507

The fly ash and tar mixture from the filter still have a high energy content of about 500 kW. Therefore, it was decided to utilise this energy by post combusting this ash. This option is shown in Figure 3.7, where additionally the ash from the filter is combusted and the flue gases are used for additional district heating production. The design of the “ash combustor” is left open at that stage.

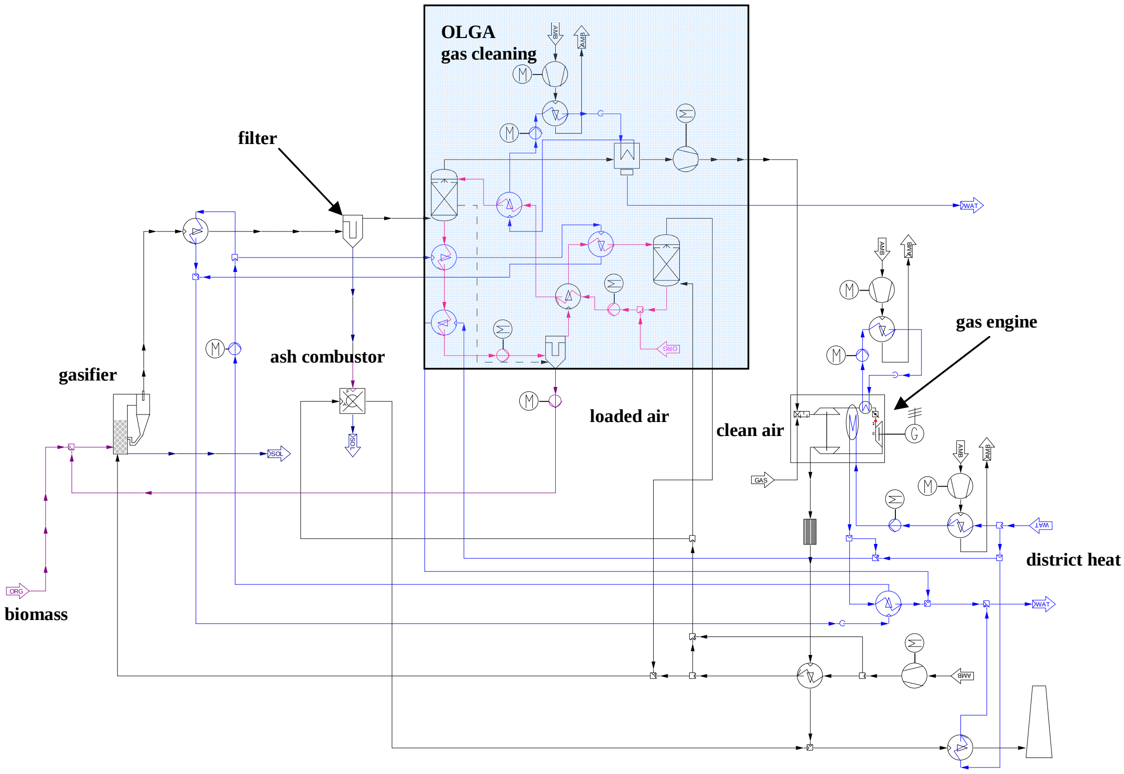


Figure 3.7. Flow chart Case 1&2 with integrated ash combustion.

Table 3.4 shows the energy balance of the overall plant with an additional ash combustion. The district heat production can be raised by about 400 kW.

Table 3.4. Energy balance overall plant Case 1.

Fuel power without OLGA [kW]	19,524
Fuel power with OLGA [kW]	20,000
Chemical power gasification gas [kW]	15,464
District heating power [kW]	10,614
Electrical power generator [kW]	5,654
Electrical consumption plant [kW]	169
Electrical net power production [kW]	5,485

3.3.2 Case 2 - CFB new filter

Case 2 is identical to Case 1 from the mass and energy balance side. The difference of the new hot gas filter does not influence the process, since the pressure drop of the filter is equivalent to the standard hot gas filter applied in Case 1. Therefore, the same performance as specified in Case 1 applies.

3.3.3 Case 3 - CFB OLGA dust

In this case the pre-coated filter is substituted by a cyclone since the effect on the system shall be investigated, if the OLGA gas cleaning system will remove tar and dust simultaneously. It is assumed that the cyclone removes about 90% of the dust from the product gas but only minor fractions of the tar.

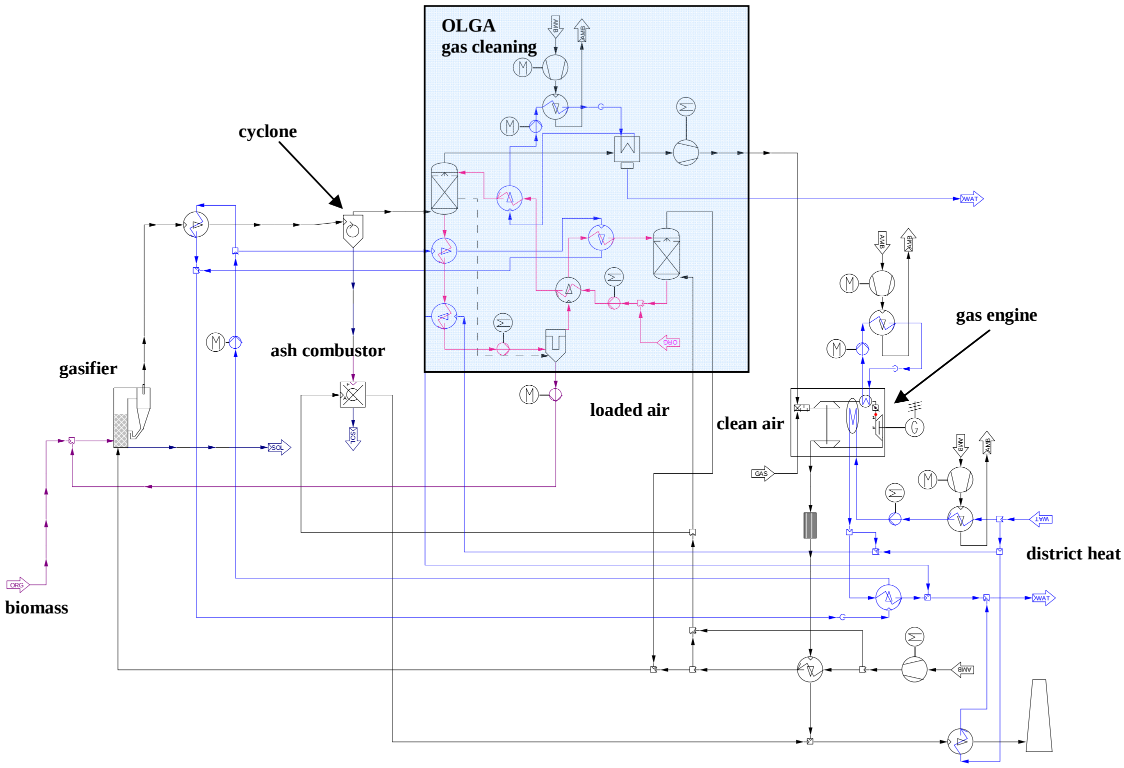


Figure 3.8. Flow chart Cases 3&4.

Advantageous is that no purge gas is needed for the cyclone and also the pressure drop is lower. The inlet temperature to the OLGA gas cleaning is kept at 320°C as well as all the other conditions of Cases 1&2. Table 3.5 shows the energy balance of the overall plant.

Table 3.5. Energy balance overall plant Case 3.

Fuel power without OLGA [kW]	19,521
Fuel power with OLGA [kW]	20,000
Chemical power gasification gas [kW]	15,635
District heating power [kW]	10,518
Electrical power generator [kW]	5,727
Electrical consumption plant [kW]	160
Electrical net power production [kW]	5,567

3.3.4 Case 4 - CFB dust & cooling

Case 4 is equivalent to case 3. However, instead of the standard inlet temperature of 320°C to the OLGA gas cleaning a higher temperature of 500°C is considered. The OLGA gas cleaning process will in this case not only remove tar and particles from the product gas, but will also act as cooling device of the product gas. The flow chart can be found in Figure 3.8 and the overall mass balance in Table 3.6. A little effect on the overall mass and energy balances compared to Case 3 can be seen.

Table 3.6. Energy balance overall plant Case 4.

Fuel power without OLGA [kW]	19,521
Fuel power with OLGA [kW]	20,000
Chemical power gasification gas [kW]	15,634
District heating power [kW]	10,459
Electrical power generator [kW]	5,727
Electrical consumption plant [kW]	160
Electrical net power production [kW]	5,567

3.3.5 Case 5 - FICFB reference

Case 5 represents the Güssing biomass gasification plant [5]. The plant is an example for allothermal dual fluidised bed steam gasification. The fundamental idea of this gasification system is to physically separate the gasification reaction and the combustion reaction in order to gain a largely nitrogen-free product gas. The principle is shown graphically in Figure 3.9. The endothermic gasification of the fuel takes place in a stationary fluidised bed. This is connected via an inclined chute with the combustion section, which is operated as a circulating fluidised bed. Here, transported along with the bed material, any non-gasified fuel particles are fully combusted. The heated bed material delivered there is then separated and brought back into the gasification section. The heat required for the gasification reaction is produced by burning carbon brought along with the bed material into the combustion section. The gasification section is fluidised with steam, the combustion section with air and the gas flows are separately streamed off. Thus a nearly nitrogen-free product gas with heating values of over 12,000 kJ/m_n^3 (dry) is produced.

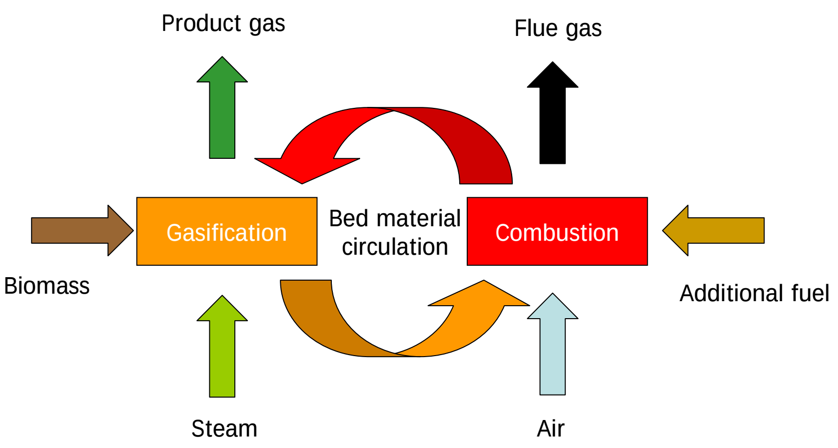


Figure 3.9. FICFB-concept.

Biomass enters the plant on the left (Figure 3.10) and is fed to the gasifier. Steam is used to gasify the biomass using the heat from the bed material for the endothermic gasification reaction. The yielded product gas is first cooled to about 180°C. Then it is filtered using a pre-coated fabric filter where the dust and about 15% of the tar is removed from the product gas. The filter ash is returned to the combustor to use the remaining energy content.

For tar removal a RME (rapeseed methyl ester) scrubber is used. Tars dissolve physically in the RME and are removed from the product gas. Condensed water is separated from the scrubbing liquid and evaporated and returned into the process. The scrubbing liquid is recycled with a small bleed of loaded liquid, which is energetically used in the combustor. The cleaned product gas is compressed and fed to the gas engines. Depending on the biomass water content a small part of the product gas is returned to the combustor to yield the necessary energy for the gasification section. The flue gases from the gas engine are cleaned by a catalyst from remaining CO-loads and the fed together with the flue gases from the combustor to the stack.

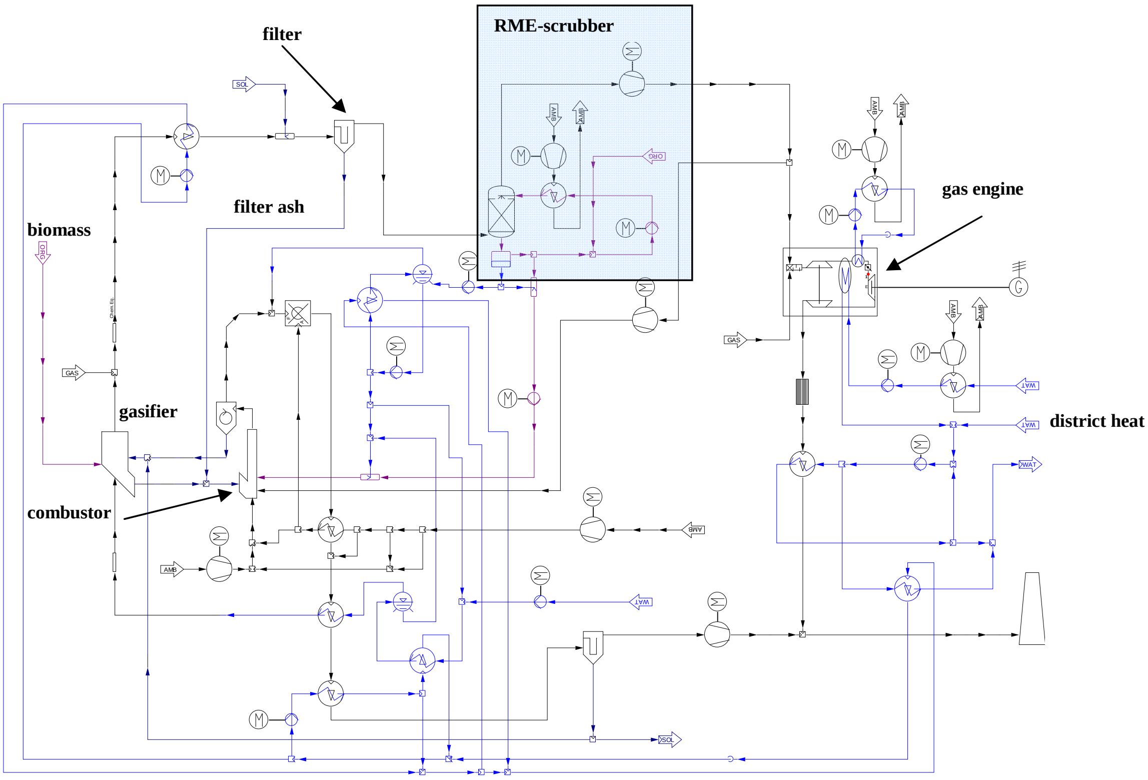


Figure 3.10. Flow chart Case 5.

In the combustor the remaining char from the gasification zone is combusted providing the necessary heat for the endothermic gasification reactions. The heated up bed material is separated from the flue gases by a cyclone and returned to the gasification section via a siphon. Any remaining combustibles in the flue gas are combusted in the post combustion chamber, before the hot flue gases are used for air preheating and district heating production. Table 3.7 shows the overall mass and energy balance for a 20 MW_{th} dual fluidised bed gasification plant.

Table 3.7. Energy balance overall plant Case 5.

Fuel power without OLGA [kW]	19,478
Fuel power with OLGA [kW]	20,000
Chemical power gasification gas [kW]	14,824
District heating power [kW]	10,144
Electrical power generator [kW]	5,409
Electrical consumption plant [kW]	254
Electrical net power production [kW]	5,155

3.3.6 Case 6 - FICFB OLGA

In case 6 the standard Güssing process is adapted for the integration of the OLGA gas cleaning process. The RME-scrubber is substituted by the OLGA gas cleaning process (Figure 3.11). Instead of the standard fabric filter the new hot gas filter is used to remove dust and some tar from the 320°C hot product gas. The dust free gas is then fed to the OLGA gas cleaning for complete tar removal. The cleaned product gas leaves the absorber column with 80°C and is further cooled. The produced condensate is returned to the process and evaporated. The bleed stream of from the OLGA gas cleaning circuit is fed together with the filter residues to the combustor. For stripping of the loaded OLGA liquid hot air from the air preheating system is used. The loaded air from the stripper is used as additional combustion air in the combustor.

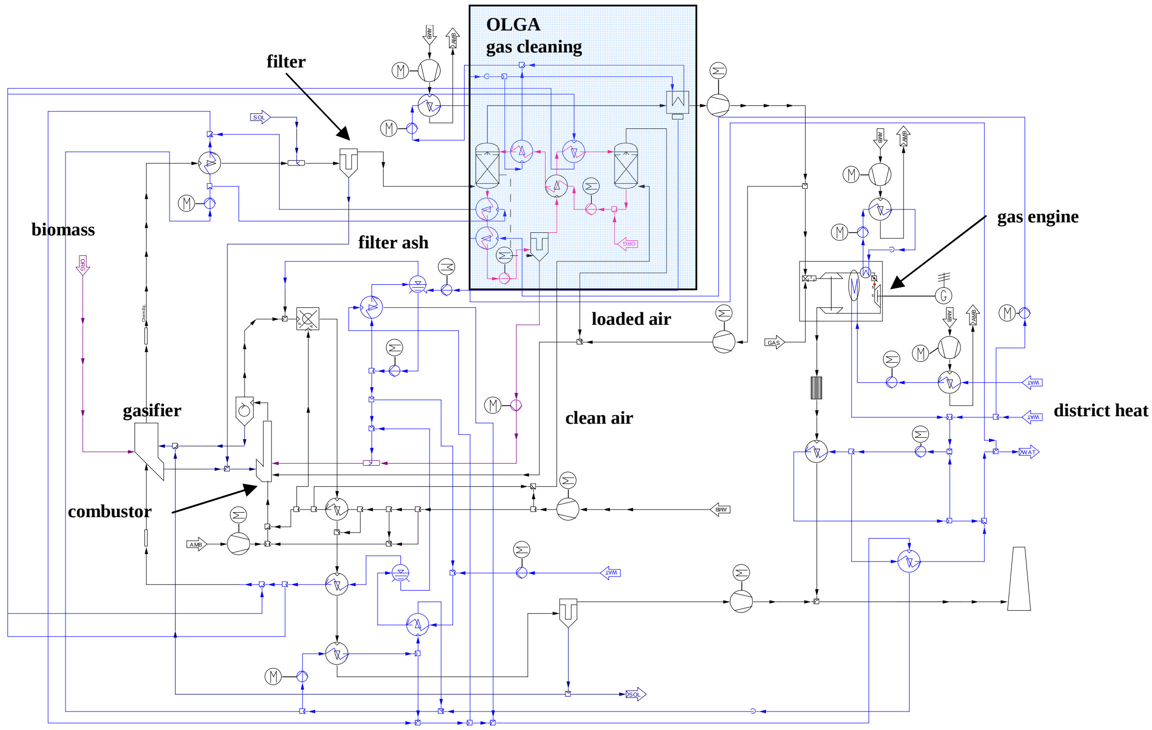


Figure 3.11. Flow chart Cases 6&7.

Table 3.8. Energy balance overall plant Case 6.

Fuel power without OLGA [kW]	19,713
Fuel power with OLGA [kW]	20,000
Chemical power gasification gas [kW]	14,800
District heating power [kW]	10,286
Electrical power generator [kW]	5,363
Electrical consumption plant [kW]	240
Electrical net power production [kW]	5,123

3.3.7 Case 7 - FICFB optimum

Case 7 differs from Case 6 by the fact that it is assumed OLGA has a possible effect on the operational degrees of freedom of the gasifier since high tar loads can be recycled by the stripped air back into the system. Therefore, it is assumed that the steam to fuel ratio can be lowered from about 0.5 to 0.35 kg/kg. This has a positive effect on the chemical efficiency of the gasifier, however, due to the catalytic effect of the steam more tar is produced. This effect was encountered by raising the tar level in the product gas to 12 g/m_n^3. The product gas still has to be cooled down for hot gas filtering which is difficult with such high tar content, since plugging is likely. Additionally, it is likely that the hot gas filter has to be pre-coated intensively to avoid plugging by tar particles. The flow chart of this plant is equivalent to Case 6. In Table 3.9 the overall mass and energy balance of the plant can be found. The chemical power of the gasification gas rises by nearly 400 kW, the electrical output can be raised by 150 kW, and the district heating output by 300 kW since less steam has to be produced and product gas can be replaced in the combustor by stripped tar. However, it has to be considered if possible problems in the product gas heat exchanger are justified by the electric and heat benefit.

Table 3.9. Energy balance overall plant Case 7.

Fuel power without OLGA [kW]	19,751
Fuel power with OLGA [kW]	20,000
Chemical power gasification gas [kW]	15,167
District heating power [kW]	10,548
Electrical power generator [kW]	5,519
Electrical consumption plant [kW]	229
Electrical net power production [kW]	5,290

3.3.8 Case 8 - FICFB dust & cooling

In Case 8 a cyclone is used for coarse particle removal before the OLGA gas cleaning system. Therefore, the temperature of the product gas after the gas cooler can be raised to about 500°C, which reduces the risk of tar plugging in the heat exchanger. The OLGA gas cleaning system has in that case to fulfil a tar removal, dust removal and cooling task. Table 3.10 shows the overall mass and energy balance of case 8.

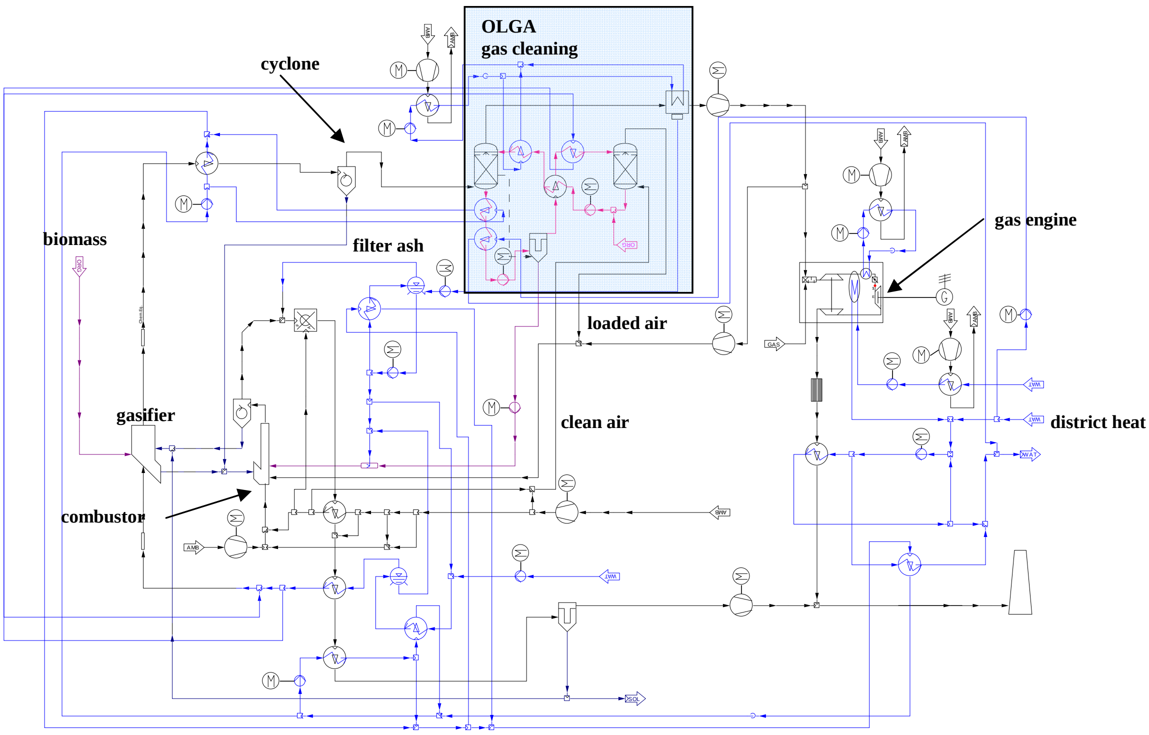


Figure 3.12. Flow chart Case 8.

Table 3.10. Energy balance overall plant Case 8.

Fuel power without OLGA [kW]	19751
Fuel power with OLGA [kW]	20000
Chemical power gasification gas [kW]	15203
District heating power [kW]	10505
Electrical power generator [kW]	5535
Electrical consumption plant [kW]	228
Electrical net power production [kW]	5306

3.4 Discussion

In this section a comparison of the different cases using the efficiencies specified in section 3.1.1 will be given. The blue bars represent the efficiency of the process, if the additional heating value of the scrubbing liquid that is combusted is not taken into account. The violet bars do include the scrubbing liquid as additional fuel and are therefore always lower.

Figure 3.13 shows the comparison of the chemical efficiencies of the different cases. No difference can be seen between Case 1*, 1 and 2, since the additional ash combustion has no effect on the chemical efficiency of the system. In Case 3 the higher amounts of fly ash and tar are removed in the OLGA gas cleaning and returned to the gasifier which increases the chemical efficiency in the product gas. An increase of the inlet temperature to the OLGA gas cleaning system (Case 4) has no positive effect on the chemical efficiency of the system. In general it can be stated that there are about 2% difference in the chemical efficiency if the fuel input (OLGA scrubbing liquid) in the OLGA gas cleaning is considered.

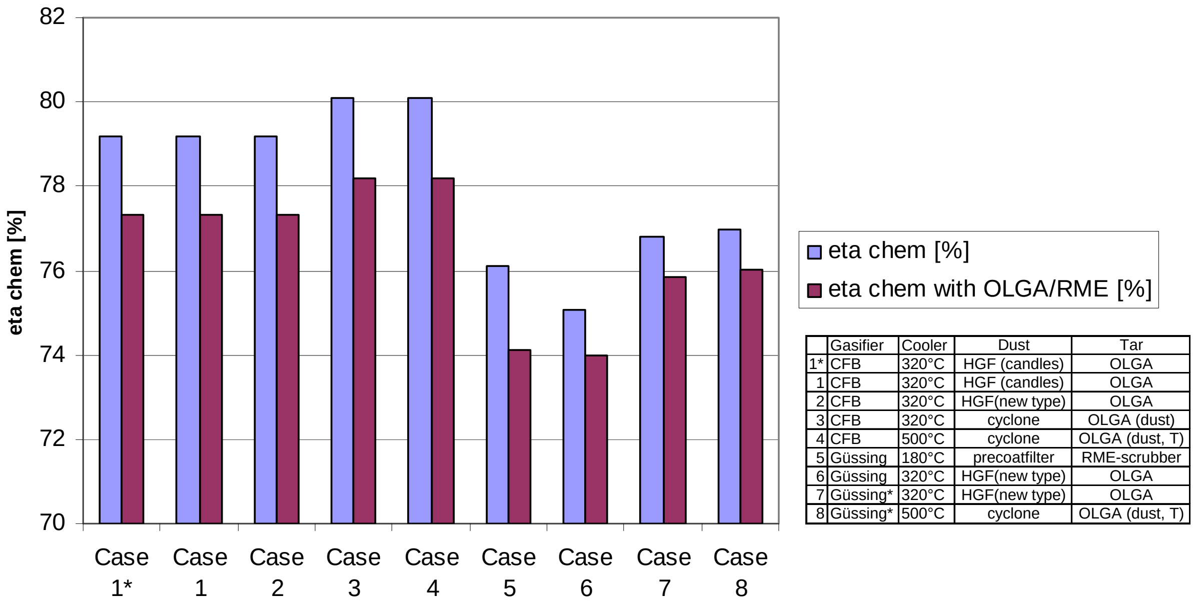
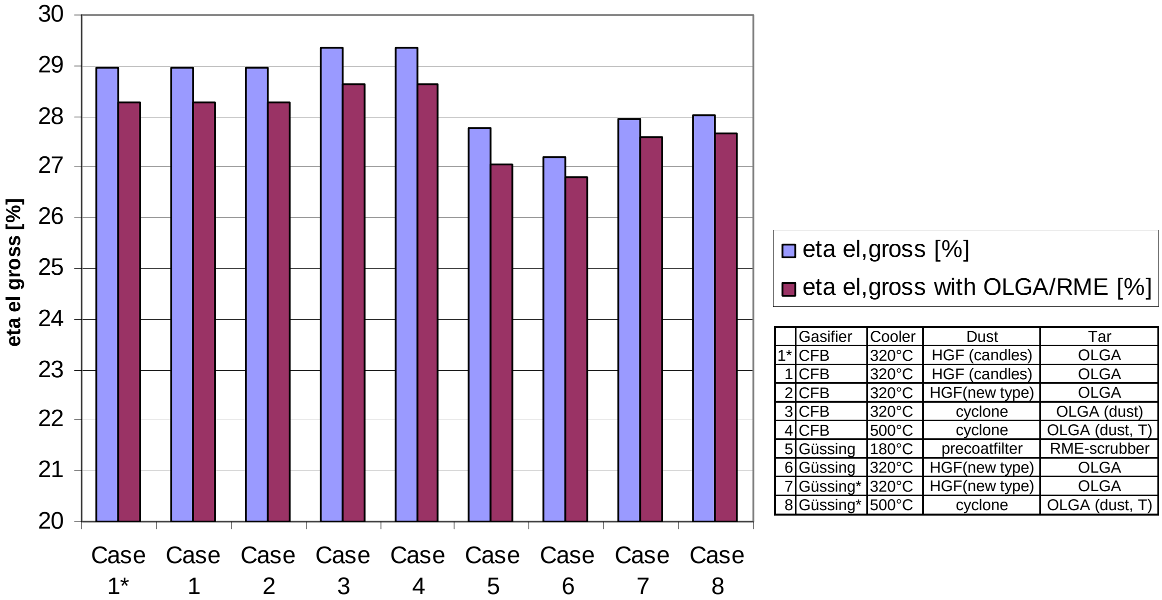


Figure 3.13. Comparison of the chemical efficiencies.

The FICFB-cases (Cases 5-8) have in general a lower chemical efficiency than the CFB cases of about 2.5%. Case 5, the Güssing plant with the RME-scrubber has the same net chemical efficiency as if the OLGA gas cleaning would be installed. A significant improvement in chemical efficiency can only be seen if the steam to fuel ratio is lowered (Cases 7&8) by approximately 2%.

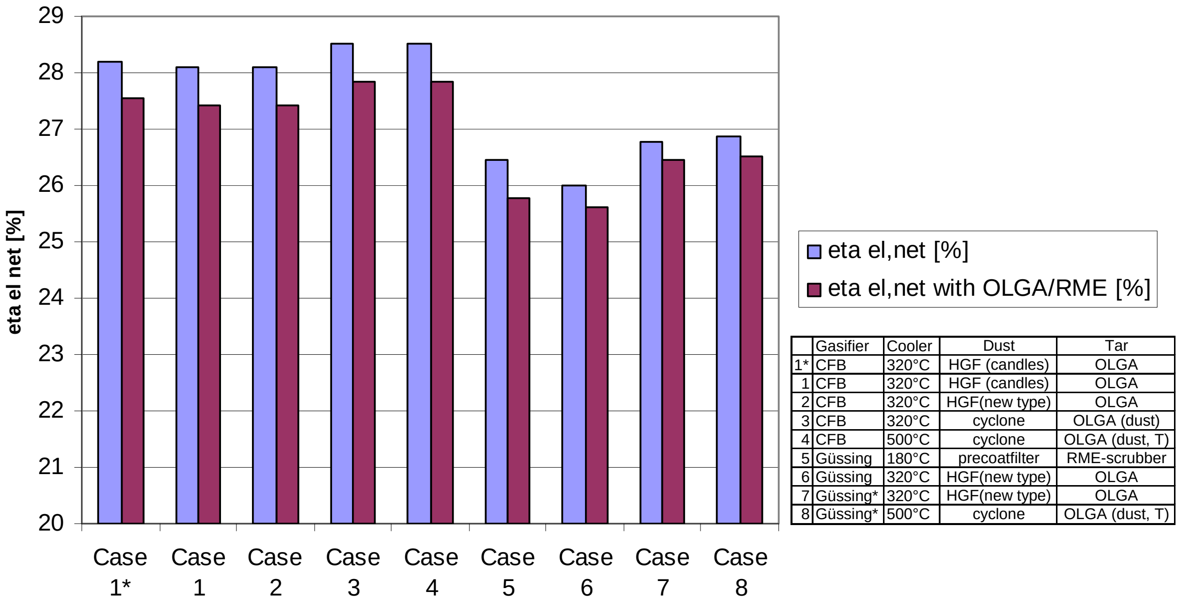
Figure 3.14 shows the gross electrical efficiencies of the different cases. As the chemical efficiency is strongly coupled to the gross electrical efficiency, the same characteristic as in Figure 3.13 can be seen. Cases 1-3 show the same gross electric efficiency of about 28% if the OLGA consumption is considered. Using OLGA also for dust removal the efficiency can be increased by about half a percent. The standard FICFB plant has an electric efficiency of about 27%, the same plant with the OLGA gas cleaning has about the same efficiency. Reducing the steam to fuel ratio (Case 7&8) the gross electric efficiency can be increased by half a percent.



	Gasifier	Cooler	Dust	Tar
1*	CFB	320°C	HGF (candles)	OLGA
1	CFB	320°C	HGF (candles)	OLGA
2	CFB	320°C	HGF(new type)	OLGA
3	CFB	320°C	cyclone	OLGA (dust)
4	CFB	500°C	cyclone	OLGA (dust, T)
5	Güssing	180°C	precoatfilter	RME-scrubber
6	Güssing	320°C	HGF(new type)	OLGA
7	Güssing*	320°C	HGF(new type)	OLGA
8	Güssing*	500°C	cyclone	OLGA (dust, T)

Figure 3.14. Comparison of the gross electrical efficiencies.

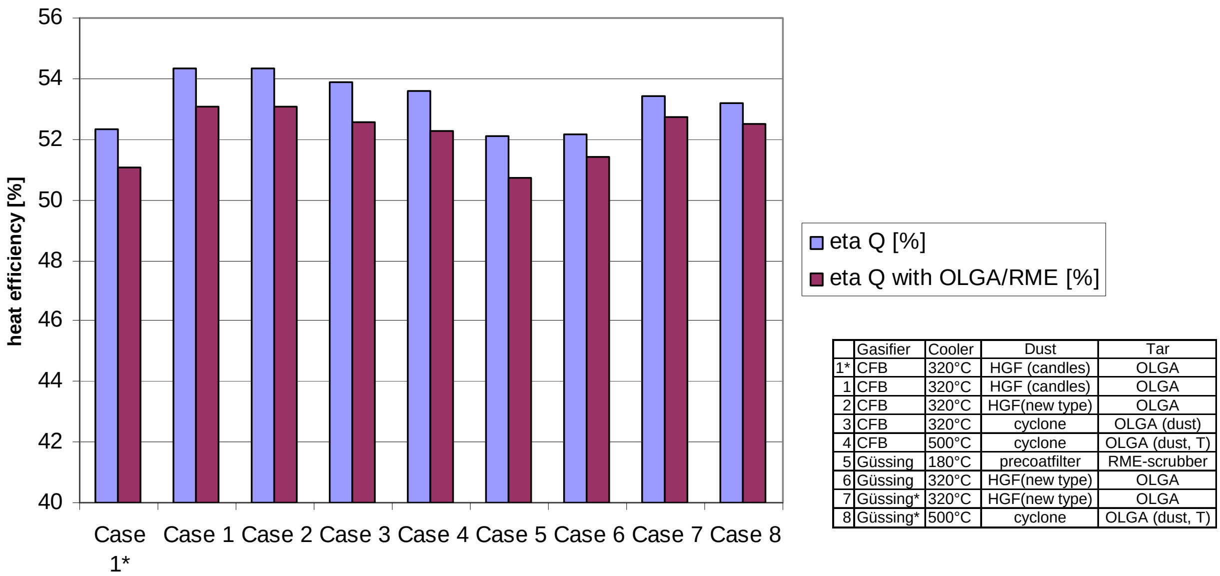
Figure 3.15 shows the net electric efficiencies of the different cases. It can be seen that the values shown are about 1% lower than the gross electric efficiencies. However, the net electric efficiencies do include only the motors and fans listed in the flowcharts and no additional consumers.



	Gasifier	Cooler	Dust	Tar
1*	CFB	320°C	HGF (candles)	OLGA
1	CFB	320°C	HGF (candles)	OLGA
2	CFB	320°C	HGF(new type)	OLGA
3	CFB	320°C	cyclone	OLGA (dust)
4	CFB	500°C	cyclone	OLGA (dust, T)
5	Güssing	180°C	precoatfilter	RME-scrubber
6	Güssing	320°C	HGF(new type)	OLGA
7	Güssing*	320°C	HGF(new type)	OLGA
8	Güssing*	500°C	cyclone	OLGA (dust, T)

Figure 3.15. Comparison of the net electrical efficiencies.

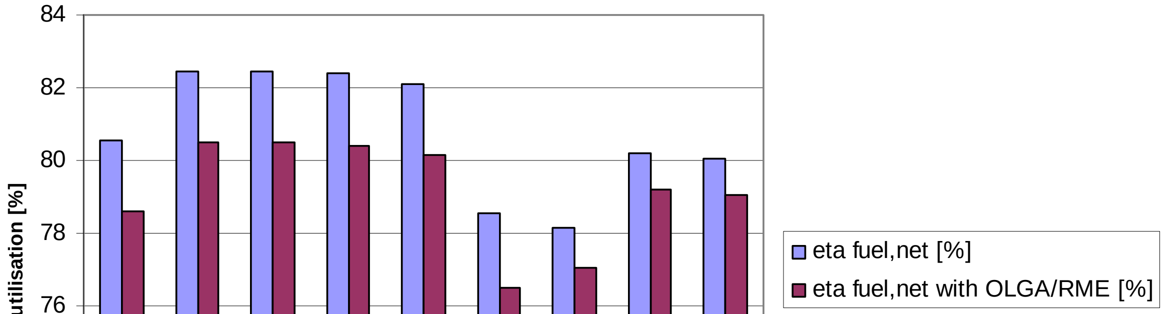
Figure 3.16 shows the comparison of the heat efficiencies. Comparing Case 1* with Case 1 and 2 the effect of the additional filter ash combustion can be seen. The heat efficiency of the plant can be raised by about 1%. In Case 3 and 4 the heat efficiency is lowered a little bit by the increase in chemical efficiency due to the tar and fly coke recirculation into the gasifier.



	Gasifier	Cooler	Dust	Tar
1*	CFB	320°C	HGF (candles)	OLGA
1	CFB	320°C	HGF (candles)	OLGA
2	CFB	320°C	HGF(new type)	OLGA
3	CFB	320°C	cyclone	OLGA (dust)
4	CFB	500°C	cyclone	OLGA (dust, T)
5	Güssing	180°C	precoatfilter	RME-scrubber
6	Güssing	320°C	HGF(new type)	OLGA
7	Güssing*	320°C	HGF(new type)	OLGA
8	Güssing*	500°C	cyclone	OLGA (dust, T)

Figure 3.16. Comparison of the heat efficiencies.

Figure 3.17 shows the total fuel utilisation. It can be seen that all CFB Cases (1-4) have approximately the same total efficiency. The higher inlet temperature to the OLGA gas cleaning system has nearly no effect on the performance of the plant. Adding the ash combustion to the system accounts for a 2% increase in the fuel utilisation, however, a cost effective realisation of an ash combustion system is necessary.



	Gasifier	Cooler	Dust	Tar
1*	CFB	320°C	HGF (candles)	OLGA
1	CFB	320°C	HGF (candles)	OLGA
2	CFB	320°C	HGF(new type)	OLGA
3	CFB	320°C	cyclone	OLGA (dust)
4	CFB	500°C	cyclone	OLGA (dust, T)
5	Güssing	180°C	precoatfilter	RME-scrubber
6	Güssing	320°C	HGF(new type)	OLGA
7	Güssing*	320°C	HGF(new type)	OLGA
8	Güssing*	500°C	cyclone	OLGA (dust, T)

Figure 3.17. Comparison of the fuel utilisation.

Possible, a fly ash recirculation to the gasifier can be the better option, but has to be discussed with the gasifier designer. Further, the dispose of the wastewater has to be considered in the economic calculations (*i.e.* in the FICFB cases the waste water is recycled to the combustor.

Comparing the standard FICFB case with the RME gas cleaning to the OLGA system only a little influence can be seen. If the steam to fuel ratio can be lowered as assumed in Cases 7 and 8 the OLGA gas cleaning system can help to increase the efficiency of the FICFB gasifier. However, then they design of the product gas cooler is more challenging.

4. Economic evaluation

This chapter describes the approach and reports the results of the economic assessment. For eight different cases of a 20 MW_{th} CHP plant, a 10th of its kind, the economic feasibility was determined and compared, using cost figures for gasifiers, coolers, OLGA, etc. from the project partners. The main aim of the economic assessment is the comparison of the different cases and the nominal investment costs or internal rate of return are of less importance.

4.1 Cases assessed

The different cases assessed are described in Chapter 2. The data from the energy and mass balances (Chapter 3) was used to calculate the power and heat production and the sizes of the equipment. Different from the energy and mass balances, for the economic calculations no on-site ash combustion was assumed in the CFB cases, as the accompanying investment costs could not be obtained. The power and heat yields for these cases were, therefore, slightly adapted to the figures given in Table 4.1.

Table 4.1. Used efficiencies of power and heat production for the different cases.

Case	Net efficiency electric [%]	Net efficiency heat [%]
1	27.53	51.08
2	27.53	51.08
3	27.94	50.60
4	27.93	50.60
5	25.78	50.72
6	25.62	51.43
7	26.45	52.74
8	26.53	52.53

4.2 Costs parameters

4.2.1 Investment costs

The investment costs calculated are valid for a 10th plant of its kind and expressed in 2006 euros. The cost figures for all the different parts of the plant were obtained from the project partners: Foster Wheeler Oy (CFB gasifier, coolers, filters) TU Vienna (Güssing gasifier and gas clean-up, gas engines) and Dahlman (OLGA, cyclones, condenser).

The total capital investment is a sum of many factors, as is shown in Figure 4.1 [6]. In general, the total capital investment is estimated based on the 'in-side battery limits' (ISBL) costs and then using several factors, depending on the type of plant, to estimate 'off-side battery limits' (OSBL) costs, indirect costs, working capital and start-up costs. However, costs estimates starting with bare equipment costs are also possible.

The ISBL costs include all the equipment, piping, instrumentation etc. including the installation of all these components. Civil and structural works, as well as process buildings, are also included. Contrary to the general use of ISBL costs, in this assessment, also the R&D, design and engineering and licences for all these components are included, because these were already included in the cost figures that were obtained from the project partners. These are costs that are usually included in the indirect costs. Thus, the here used ISBL costs, which are also used to calculate the other costs, are higher than ISBL costs as they are used generally in other assessments.

The OSBL include costs factors such as service facilities, product storage and land. OSBL costs are generally in the order of 45-50% of the ISBL costs. As in this case, the ISBL costs are actually include more than what is generally used, in order to calculate the OSBL costs a lower factor of 35% of the ISBL costs is used.

Indirect costs include the construction of the plant, contractor's fees, etc. They also usually include R&D, design and engineering of the equipment and licenses. However, as mentioned earlier, in this assessment these costs factors are already included in the ISBL costs. Indirect costs can vary a lot, from 18% to even 43% of the combined ISBL and OSBL costs. In this case 18% was used, since a considerable part of the indirect costs is already included in the ISBL costs.

Working capital can vary from 12-28% of the fixed capital investment. In this assessment the lower value of 12% was chosen, because the process is mainly capital intensive and not so much labour intensive and, therefore, at least for wages not much working capital is necessary. The nominal amount resulting from the 12% of the fixed capital investment actually compares very well with the amount of cash equal to all the operational costs and sales revenues for a period of three months, which is an alternative method to calculate working capital.

The start-up costs are costs such as initial labour costs and raw material usage in start-up. The start-up costs are generally in the order of 8-10% of the fixed capital investment and in this assessment a figure of 9% is used.

<table>
<tr><th colspan="5">Total Capital Investment</th></tr>
<tr><th colspan="3">Fixed Capital Investment</th><th rowspan="3">Working Capital</th><th rowspan="3">Start-up Costs</th></tr>
<tr><th colspan="2">Direct Costs</th><th rowspan="2">Indirect Costs</th></tr>
<tr><th>ISBL costs + R&D
(onsite)</th><th>OSBL Costs
(offsite)</th></tr>
<tr><td>Purchase & Installation of
• Process equipment
• Piping & appurtenances
• Instrumentation & Controls
• Electric equipment & materials
• Civil & structural
• Process Buildings
• Up-front R&D
• Up-front license
• Equipment design and engineering</td><td>• Yard Improvements
• Auxiliary buildings
• Service facilities
• Storage/distribution
• Land</td><td>• Construction
• Contractor’s fee
• Contingencies</td><td>• Inventories
• Salaries/wages due
• Receivables less payables
• Cash</td><td>• Modifications
• Start-up labour
• Loss in production</td></tr>
<tr><td>52%</td><td>18%</td><td>13%</td><td>10%</td><td>7%</td></tr>
</table>

Figure 4.1. Average breakdown of total capital investment, with the percentages used for this study. Adapted from reference 6.

As the ISBL costs are different for the different cases, all the other costs will also be different when using a standard cost breakdown. However, there is no reason why these costs should be different for different cases, e.g. there is no reason that land costs should be different for erecting the CFB or the Güssing gasifier or why e.g. late payments for power sales should be different between cases. Therefore, the figures for OSBL costs, indirect costs, working capital and start-up costs were averaged over all the cases and these averages were used for all the cases.

Reliable investment costs for wastewater treatment, *i.e.* ammonia removal from the water coming out of the condenser could not be obtained. Costs for wastewater treatment have been included in the operational costs (see section 4.2.2). The investment for a DeNOx is included, but only so far to achieve the Austrian norms, not the Dutch ones. It was not possible to obtain the additional costs for the Dutch norms, but since these costs would apply to all cases, the impact on this study by neglecting this budget will be minimal.

4.2.2 Operational costs

For labour costs it was assumed that one person is present continuously and that five shifts are necessary. In addition, supervision costs, laboratory costs and plant overheads are taken into account. Maintenance is estimated as 5% of the fixed capital investment per year. The insurance of the plant is estimated at 1% of the fixed capital investment per year.

Wood is used to cost 4 €/GJ delivered at the factory gate. Bed material, nitrogen and scrubbing oil costs are based on the amounts and costs delivered by the project partners. Waste waster treatment costs for the removal of ammonia were difficult to obtain. Known figures vary from 1 €/m^3 of water to as much as 50 €/m^3 of water. For the calculations 10 €/m^3 of water was assumed.

Ash disposal costs were estimated based on prices in The Netherlands, which are relatively high compared to other European countries. For ash without carbon, from the Güssing gasifier, disposal costs are estimated at 150 €/t. According to new EU regulations, ashes with high carbon content, from the CFB gasifier, cannot be deposited anymore. Therefore, in the system modelling these ashes were combusted for heat generation. However, since costs figures for ash combustion could not be obtained, in the economic assessment the original system was used. For the ashes it was assumed that they were used for other purposes off-site at a cost of 220 €/t.

4.2.3 Plant and economic parameters

The profitability of a plant depends very much on economic conditions and possible subsidies. It is difficult to predict these conditions; the economic factors assumed are mostly based on current or near-future conditions and are given in Table 4.2.

Table 4.2. Assumed plant and economic parameters.

Parameter	Value	Motivation
Depreciation period	15 year	General figure
Interest rate	6 %	Value for a non-risk investment
Corporate tax	25 %	Dutch figure for 2007 (also close to Western European average)
Power sales price	32 €/MWh	Current market price
Green power support	97 €/MWh	Dutch subsidy for Green power < 50 MW_e
Heat sales price	6 €/GJ	More or less average western European heat price, actual prices can be 3-15 €/GJ.
Operational time	7500 hours	Value used in calculations for the Dutch green power subsidies
Period of heat sales per year	7500 hours	Full utilisation of heat is necessary for economic feasibility

4.3 Economic method

As a measure of economic feasibility, the 'internal rate of return' (IRR) was used. The IRR is the discount rate that results in a net present value of zero for a series of future cash flows. Its formula is given in Equation 7. The IRR does not provide any information on the period after the depreciation time, so when comparing different systems, they should have the same depreciation time. Usually a IRR of 12-15% is required for a project to proceed, but in the case of energy projects, in which product sales are very much guaranteed, as low as 10% can be accepted.

$$\sum_{t=0}^{\text{Depreciation}} \frac{\text{Cash flow}}{\left(1+\frac{IRR}{100\%}\right)^{t}} = 0 \qquad \text{Equation 7}$$

4.4 Results

Based on the investment costs, operational costs and the revenues, for all cases the IRR was calculated and the results are given in Table 4.3. In general, it can be concluded that under the assumed economic conditions the IRR's are in the range of 4.5-6.5. Since 10% is seen as the minimal IRR required, all the cases can be considered not viable. However, the results are sufficient for comparison of the different cases.

Table 4.3. Results of economic assessment.

Case	Total capital investment (M€)	IRR (%)
1	28.7	4.6
2	28.6	4.6
3	28.6	4.9
4	28.5	4.9
5	24.2	6.3
6	24.9	6.1
7	26.2	6.0
8	26.2	6.0

4.4.1 Comparison of the cases

When comparing the cases, the results show that there is only very small difference between Case 1 and Case 2, which only differ in the type of filters used. The investment costs for Case 3 are also almost equal to Case 1 and 2, but this system has a somewhat higher power yield and thus higher revenues. Case 4 is almost similar to Case 3; the slightly lower costs for the cooler do not make a large difference. The Güssing benchmark (Case 5) has lower investment cost, lower operational costs (mainly due to lower maintenance and ash disposal costs), but also lower revenues, because the power yield is much lower than for the CFB cases. Overall, it has a better performance than the CFB cases. Case 6, Güssing including OLGA, requires a higher investment, but has a higher total yield. However, the higher yield is caused by higher heat yield and the power yield is actually lower than for Case 5. Since power is valued much higher than heat the actual revenues are in this case, in fact, lower than for Case 5. Case 7 clearly requires a higher investment than Case 5 and 6, mainly because of a more expensive cooler, which is necessary for the higher tar loading. Since the heat and power yield are higher, the revenues are also much higher compared to Case 5 and 6, although this does not completely compensate for the higher investment. Case 8 has a slightly higher investment than Case 7 and the higher revenue does not make up for this.

4.4.2 CFB versus FICFB

The cases with a CFB gasifier are more expensive than the ones with the FICFB gasifier. This is caused by higher costs for the gasifier, but also by higher costs for most downstream equipment, since the gas flow is larger in the CFB cases (10,291 vs. 6,085 m_n^3). Of the higher investment costs for the CFB cases compared to the Güssing cases (and in specific Case 4 and 8) little less than half of the difference is caused by the fact that the downstream equipment for the CFB cases is more expensive, because the CFB case has a higher gas flow (10,291 m_n^3/h vs. 6,085 m_n^3/h). The remainder is because of the higher costs for the CFB gasifier than for the FICFB gasifier. This might come as a surprise as there no obvious reason why a CFB gasifier would be more expensive than a Güssing gasifier. A possible explanation might be that for Foster Wheeler Energia Oy that made the cost estimate for the CFB gasifier, the scale of 20 MW_{th} is too small compared to what they generally supply and, therefore, relatively expensive. The FICFB gasifier, however, has so far only been built at 8 MW_{th} (in Güssing).

4.4.3 FICFB with OLGA

Replacing the current tar removal method for the Güssing gasifier, the biodiesel scrubber, with OLGA (Case 5 vs. 6) does not result in a better economic performance. Next to the somewhat higher investment costs for system with OLGA, there are no economic benefits as only the heat yield increases, whereas the power yield even decreases. When the FICFB gasifier is optimised (Case 7), there are significantly higher revenues from the increased power and heat yields. However, under the current assumptions, this is not enough to compensate for the higher investment costs. The latter are mainly caused by higher cooler costs, because in the optimised case the tar load of the gas is much higher than in the FICFB reference case. Still, the difference between the benchmark and optimised case is not very large and e.g. a 25% reduction in the investment costs for OLGA would be sufficient to make the optimised case equally economically viable to the benchmark case. This means that there is an incentive for further optimisation of OLGA.

4.4.4 Dust removal and cooling by OLGA

In addition to tar-removal, it is possible to use of OLGA for dust removal and additional cooling in the 320-400/500°C range. For dust removal this would mean replacing the filter(s) by cyclones and including an ESP in OLGA. For additional cooling it means that the cooler is somewhat smaller and that OLGA has a higher inlet temperature. For both situations, the differences in investment costs are minimal, as can also be seen in differences in investment costs between Cases 2, 3 and 4 and between Cases 7 and 8. As for the economic performance, dust removal is favoured in the CFB case, as the power yield is higher when OLGA is used for dust removal. From the economic point of view, there is no preference for dust removal by OLGA or not in the FICDB case and for cooling by OLGA or not in both CFB and FICFB cases.

4.4.5 The 20 MW_{th} system

The choice for a 20 MW_{th} system poses some difficulties. First, the scale of 20 MW_{th} is quite large to use gas engines and the gas engines are a considerable cost factor. However, it is still too small to use a gas turbine and possibly a steam cycle, which would give a higher efficiency and possibly a better economic performance. Secondly, as already mentioned, a CFB gasifier supplied by Foster Wheeler Energia Oy for this scale would be relatively expensive. A cost estimate for this scale from a company that has built several CFB gasifiers in the range of 5 MW_{th} gave a much lower investment cost.

Thirdly, for the coolers used in the systems with high-tar load (Case 1-4, 7, and 8) are much more expensive than the water cooled fire-tube used for low tar load (in Case 5 and 6). The more complex and expensive coolers, which are supplied by Foster Wheeler Energia Oy, are selected in this assessment for cases with high tar-load. The inexpensive double pipe (fire-tube) coolers can also be used, however, to avoid tar condensation they need to be air-cooled. This design is not preferred as in case of leakage there is the risk of product gas getting in contact with hot air.

In order to evaluate the effects of a cheaper gasifier and/or product gas cooler, all the cases were recalculated first by assuming that an inexpensive cooler can be used for the systems with high tar load, second by assuming the lower cost estimate (which is actually lower than the costs of the FICFB gasifier) for a CFB gasifier and third by doing both together. The results are shown, along with the original calculation, in Table 4.4.

Table 4.4. Effect of cooler and gasifier costs on the economic assessment.

Case	IRR (%)	IRR (%) Lower cooler costs	IRR (%) Lower CFB gasifier costs	IRR (%) Lower cooler & CFB gasifier costs
1	4.6	6.5	7.1	9.5
2	4.6	6.5	7.1	9.5
3	4.9	6.8	7.4	9.8
4	4.9	6.8	7.4	9.8
5	6.3	7.1*	7.2*	8.1*
6	6.1	6.9*	6.9*	7.8*
7	6.0	7.8	6.8*	8.8*
8	6.0	7.8	6.8*	8.7*

* The input for these systems has not changed, but since ISBL costs of other cases have decreased, the OSBL, indirect, working capital and start-up costs, which are averaged over all cases, have decreased. This means that values within a column can be compared very well, but values between columns to a lesser extent.

4.4.6 Lower cooler costs

If a low cost cooler can be used in all cases, the FICFB cases still have a better performance than the CFB cases. However, the difference between the FICFB reference case and FICFB optimised is now remarkable. Although the investment for the benchmark case is still lower than for the optimised case, the increase revenues from the higher power and heat yield in the optimised case now more than compensate this. Whether or not a low cost cooler can be used depends. At least, it is more likely that a double pipe cooler (water cooled) can be used in a 20 MW_{th} Güssing system than in a 20 MW_{th} CFB system, which would require an air cooled cooler. For the 20MWth optimised FICFB case it is also more likely that an air cooled cooler is used. Also, if the tar-load is actually still too high for a low cost cooler or an air cooled cooler is not permitted, given the significant difference in economic performance, primary measures to reduce tar formation in the gasifier are worth considering. This is actually already done at Güssing to allow the use of the water-cooled product gas cooler.

4.4.7 Lower CFB costs

When the alternative cost estimate, from a small-scale CFB supplier, is used for the CFB costs, the CFB cases are slightly favoured over the Güssing cases. This alternative CFB cost estimate is lower than the costs for the Güssing gasifier and more than compensates for the higher downstream equipment costs due to the higher gas flow. If both the lower cooler costs and the lower cost estimate for the CFB gasifier are applied, the CFB cases 3 and 4 actually become very close to economic viability.

4.5 Discussion

The economic assessment revealed that all the 20 MW_{th} systems are economically unattractive. When comparing the different systems, the ones with a CFB gasifier are less attractive than the ones with a Güssing gasifier, because of higher costs for the downstream equipment (larger gas flow from CFB) and higher costs for the CFB gasifier. However, there is no straightforward reason why a CFB would be more expensive than a Güssing gasifier. When a lower budget estimate for the CFB gasifier, obtained from a small-scale CFB supplier, is used, the CFB cases are favoured over the FICFB cases.

From the economic assessment it can be concluded that replacing the biodiesel scrubber with OLGA in the FICFB reference case has a negative economic effect. Also, when the gasifier is optimised in the system with OLGA, the higher investment costs are not fully compensated by the higher revenues from power and heat. However, the difference between the benchmark and optimised case is not very large, which means that there is still an incentive for further optimisation of OLGA. Furthermore, the higher investment costs for the optimised case are mainly caused by higher cooler costs. It might actually be possible to use a low cost cooler like a double pipe cooler operated with air as cooling agent. Also another more active catalyst inside the gasifier could be taken to reduce the tar content (dew point). This might then lead to a significant advantage of the Güssing optimised system compared to the FICFB reference system. For the CFB cases, a low costs cooler could improve the economic viability as well. If a low cost cooler could be used and also the lower CFB investment costs would apply, the CFB systems actually come very close to economic viability.

Using OLGA for dust removal and cooling in the 320-400/500°C range, in addition to tar-removal, mostly has very little impact as the difference in investment costs and yields are minimal. Only the dust removal by OLGA in the CFB system has a clearly positive economic effect, because power output from this system is higher.

5. Conclusions

Gasification of biomass to convert the solid biomass in a combustible product gas is the key-step in integrated biomass gasification systems for the production of electricity and heat, in so-called bio-CHP plants. After cooling and cleaning the product gas can be applied in a prime mover (*i.e.* gas engine). To make the product gas suitable for a gas engine the gas must be cooled and all impurities dust, tars, NH_3 and HCl must be removed.

The main technical challenge in the implementation of integrated biomass gasification plants is the removal of tar from the product gas. “Tar” is equivalent to a major economic penalty in biomass gasification due to fouling and loss of availability. At ECN the OLGA tar removal technology is developed, which is based on applying an organic scrubbing liquid. The operational conditions of the first OLGA design required cooling of the gas to below 320-340°C (due to stability of the applied washing liquid) and essentially complete de-dusting of the product gas with an upstream high-efficient hot gas filter.

The cost-effectiveness of integrated biomass gasification plants can be improved when the capital and operational costs are reduced and the reliability and the number of operational hours are increased. Identified system improvements are:
1. Making the hot gas filter superfluous by making the OLGA suitable for combined tar and dust removal.
2. Increase the outlet temperature of the cooler to >400°C to minimise the risk of tar condensation and resulting fouling of the gas cooler.

Both system improvements can be realised by innovative extension of the OLGA functionalities. In this study the impact of these OLGA extensions on the system efficiency and economic potential were assessed for eight selected systems (Table 2.1).

System assessment
CHP systems based on a circulating fluidised bed (CFB) gasifier with an OLGA for combined dust and tar removal operated downstream a cyclone have on average a 1%-point higher chemical efficiency that the corresponding systems with a hot gas filter. This is caused by higher amounts of fly ash (*i.e.* mainly carbon) and tar that are removed in the OLGA and returned to the gasifier. In these systems the chemical efficiency reaches 78% based on the biomass input (*i.e.* corrected for the recycling of spent scrubbing oil to the gasifier). The extension of the OLGA functionalities improves the efficiency. The other system modification of increasing the inlet temperature to the OLGA gas cleaning system has no positive effect on the chemical efficiency of the system.

Systems based on the FICFB gasification technology have in general a 2.5%-point lower chemical efficiency than the corresponding CFB systems. The main reason for this difference is the high steam consumption (*i.e.* steam to fuel ratio) in the FICFB gasifier. A significant improvement of 2%-points in chemical efficiency of the FICFB system can be achieved if the steam to fuel ratio is lowered. Reducing the steam consumption is possible as the gas inlet temperature of the OLGA is much higher than of the standard RME scrubber (*i.e.* 400 versus 140°C).

The gross electrical efficiencies of the different cases are coupled to the chemical efficiencies and the same trends are observed. Maximum efficiencies are approximately 28.5%. The standard FICFB plant has an electric efficiency of about 27%; the same plant with the OLGA gas cleaning and lower steam consumption has an efficiency of more than 27.5%. The net electric efficiencies are approximately 1%-point lower than the gross electric efficiencies. The heat efficiencies for similar CFB and FICFB systems are similar, *i.e.* approximately 52%.

Summarising it can be concluded that the use of an OLGA for combined dust and tar removal increases the system efficiency, while changing the cooler temperature has no efficiency effect (although it is desired for operational point of view). The total fuel utilisation in the CFB systems is more than 80%, while the standard FICFB case has a 4%-point lower utilisation. Installation of an OLGA unit to replace the RME scrubber has little effect. However, the OLGA allows also, in theory, the reduction of the steam consumption in the gasifier and this results in an increase of the fuel utilisation to 79%.

Economic evaluation

The economic assessment revealed that all the 20 MW_{th} systems are economically not very attractive (*i.e.* IRR below 10%) under the conditions assessed. In the base-case assessment, the FICFB systems are more attractive than the CFB-based systems, in spite of the higher efficiencies of the CFB systems. This results partly from the higher costs for the downstream equipment due to the higher product gas flow, but mainly from the higher costs of both the CFB gasifier and the product gas cooler. The impact of the two latter effects is quite significant.

When a more competitive budget estimate for a CFB gasifier is used, the attractiveness of the CFB cases increases with 2.5%-points. In that systems based on CFB and FICFB are equally attractive, which means that the higher costs for the downstream equipment in the CFB systems are compensated by the higher efficiencies. In the bases-case a relative expensive product gas cooler is selected that is cooled with steam. An alternative is to use a much simpler and cheaper fire-tube cooler that is cooled with air (as in operation and tested for 700 hours downstream a CFB gasifier at ECN). Installing the alternative and cheaper cooler results in an increase of the IRR with 1.9%-point.

Summarising, it can be concluded that the IRR of the FICFB reference case is 8.1%, that the economic viability can be increased to 8.8% when an OLGA unit and an air-cooled fire-tube cooler is installed. However, the highest IRR of 9.8% is realised in the system based on a CFB gasifier with an OLGA (and fire-tube product gas cooler).

Concluding

The cost-effectiveness of integrated biomass gasification CHP plants can be improved when an OLGA unit for combined tar and dust removal is part of the system.

Appendix A Symbols and abbreviations

Symbols	Description	Unit
h	specific enthalpy	J/kg
hhv	higher heating value	MJ/m_s^3
lhv	lower heating value	MJ/m_s^3
$\dot{m}$	mass flow	kg/s
P_{el}	electrical power	kW
P_{th}	thermal power (fuel power)	kW
$\dot{Q}$	district or process heat	kW
R	general gas constant (R = 8.31451)	J/(mol·K)
s	specific entropy	J/(kg·K)
T	temperature	K

Abbreviations	Description	Unit
CFB	circulating fluidised bed	
CHP	combined heat and power	
DLL	dynamic link library	
FICFB	fast internal circulating fluidized bed	
HGF	hot gas filter	
MDK	model developer kit	
PSE	process simulation environment	
RME	rapeseed methyl ester	
SFB	stationary fluidised bed	

Subscripts	Description	Unit
0	ambient conditions	
chem	chemical	
cons	consumption	
el	electric	
Fuel	fuel (biomass)	
gross	gross value	
GC	gas cleaning	
GE	gas engine	
input	input	
m	mechanic	
net	net value	
own	own consumption	
PG	product gas	
prod	production	
Q, q	heat	
s	isentropic	
SCR	selective catalytic reduction	
th	thermal	
wf	water free	
wt	weight	

Greek Symbols	Description	Unit
α	power to heat ratio	-
Δp	pressure drop of equipment	bar
η	efficiency	-

6. Literature

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