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Prerequisites for OCAC campaign at Sävenäs HP3 – a biomass-fired 95 MW BFB boiler

Master's thesis in Sustainable Energy Systems

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MASTER'S THESIS

**Prerequisites for OCAC campaign at Sävenäs
HP3 – a biomass-fired 95 MW BFB boiler**

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Gothenburg, Sweden 2018

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Abstract

This thesis is a qualitatively investigation, partly based on quantitative data, of the possibility and consequence of introducing oxygen-carrying bed material in the industrial biomass-fired BFB boiler Sävenäs HP3. The boiler is operated by Göteborg Energy and supplies district heat and electricity to Gothenburg. The investigated oxygen-carrying material is LD-slag, a material that is relevant thanks to its availability and low cost. Based on identified risks, an experimental campaign realization for Sävenäs HP3 is proposed. The objective is to introduce and evaluate the effect of LD-slag as bed material with as low risk as possible for the boiler operator.

The agglomeration temperature and particle size distribution is determined for LD-slag and silica sand in addition to their fluidization behaviour. It is concluded that LD-slag performs superbly as bed material, without any tendencies towards agglomeration or attrition. It is found that a batch-wise introduction of LD-slag is the most suitable experimental campaign for Sävenäs HP3. A time-varying bed composition with respect to share of different components is estimated for different batch sizes. A batch of 30 ton LD-slag is determined, ensuring at least 30/70% of LD-slag and silica sand, respectively, for three days. No modifications nor installations of new equipment to the existing process is required to implement the proposed campaign strategy.

Keywords: biomass combustion, agglomeration, industrial boiler, bubbling fluidized bed, oxygen-carrier aided combustion, LD-slag, experimental campaign.

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Nomenclature

FBC	Fluidized bed combustion
BFB	Bubbling fluidized bed
CFB	Circulating fluidized bed
OC	Oxygen carrier
OCAC	Oxygen-Carrier Aided Combustion
CLC	Chemical-Looping Combustion
ESP	Electrostatic precipitator
SNCR	Selective Non-Catalyst Reduction
LD	Linz-Donawitz

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Chapter 1

Introduction

Today, fluidized bed combustion is considered as the most attractive technology for combustion of biomass and different solid wastes. Traditionally, silica sand is used as bed material in fluidized beds. However, lately the opportunity to use an active bed material that distribute the available oxygen has received great interest. The concept of using a so-called oxygen-carrying material as bed material is called Oxygen Carrier Aided Combustion, OCAC.

The technique of OCAC originates from Chalmers University of Technology and its research concerning Chemical-Looping Combustion, CLC, which is a combustion technique to separate carbon dioxide. CLC requires a reliable and stable oxygen-carrying bed material under operation, and different materials have been proven to possess this property. One material with oxygen-carrying properties that currently have a high potential is a byproduct from iron- and steel industries called LD-slag. By successful OCAC operation, not only more beneficial operation of the boiler is achieved, the behaviour of the active bed material during operation can be further studied as well.

Currently, Chalmers is participating in a European research project called OxyCar-FBC in collaboration with Göteborg Energi, Wien University of Technology and Bertsch Energy and others. The project treats modelling, process studies and experiments with oxygen carrying materials in fluidized beds of different capacities ranging from 10 kW to 100 MW. Included in the OxyCar-FBC, an examination if an experimental campaign with oxygen carrying bed material in an industrial scale boiler should be carried out or not.

1.1 Aim and objective

In this study, the technical prerequisites of introducing oxygen carrying bed material in an industrial-scale fluidized bed are analyzed. The aim is to identify essential risks in order to design a experimental campaign where a portion of the bed is replaced with LD-slag. The fluidized bed considered is a 95 MW_{th} bubbling fluidized bed boiler located in Sävenäs, Gothenburg, operated by Göteborg Energi.

Expected outcome of OCAC operation is studied in literature while material properties of LD-slag and samples of silica sand and bottom ash are quantitatively compared. Based on the resulting differences and similarities, technical prerequisites of

Sävenäsverket for introducing LD-slag are qualitatively evaluated. A final strategy for a low-risk experimental campaign to introduce LD-slag in Sävenäsverket is constructed and presented.

For Göteborg Energi and Sävenäsverket the transition to OCAC operation can counter certain performance issues. Currently the fluidized bed of Sävenäsverket suffers from temperature differences over the cross-section of the boiler. This boosts the variation in oxygen concentration in the boiler which may imply increased emissions of non-complete combustion compounds. Additionally, too high moisture content limits the fuel intake as the supply of primary air is already operating at maximum flow rate to avoid particle entrainment. Hence, the desired benefits of OCAC in Sävenäsverket is to smoothing out the bed temperature, to decrease the emission related to non-complete combustion and to broaden the range of possible fuel qualities with respect to moisture content. Further improvements relevant for Sävenäsverket are for example improved overall efficiency by lowering the surplus of air supplied.

The resulting qualitatively evaluation and experimental campaign strategy presented in this project will provide information and aspects necessary to consider when moving from silica sand operation to operation with an oxygen carrying bed material.

1.2 Limitations

The scope of the project is not to answer if the oxygen carrying capacity is the most essential parameter of the bed material or not. The project will neither evaluate the complex chemical properties associated with the bed material. Instead, the aim is to present supporting material and identified prerequisites for an experimental campaign at Sävenäsverket.

This project is limited on the implementation of LD-slag. Even though there are more materials that exhibit oxygen carrying properties, LD-slag is considered as the most relevant substitute to be used during a campaign in the industrial boiler considering its availability, cost and impact on the environment.

Chapter 2

Background and Theory

In this chapter the underlying theory for this project is presented. To begin with, theory related to the combustion process applied at Sävenäsverket is explored regarding fluidized bed combustion, FBC, and solid biomass combustion. Further, introductory theory regarding fluidization and oxygen-carrying bed materials is presented.

2.1 Development of combustion technology

The development and commissioning of fluidized bed combustion units, FBC, have flourished since the first commercialised implementation in the early 1970's. In contrary to the grate-fired combustors, FBC units are generally more capable of keeping the emissions low and even operate with improved energy efficiency as the unit is operating with more low-graded fuel types or higher content of ash and moisture [1, 2]. The increased allowance for heterogeneous fuel mixtures is essential for biomass combustion as the incoming fuel quality is typically varying on daily basis. The increased operational performance is not the only driving force behind the transition to FBC units, but also the possible implementation through retrofit of the existing grate fired furnaces instead of installing new utilities [1].

The commercial biomass- and waste FBC boilers are today operated with silica sand as bed material. The amount of supplied air is greater than the stoichiometric amount required for full combustion to counteract the poor mixing between warm flue gases and cold combustion air.

An innovative concept to improve the FBC technology performance is to replace the silica sand with a so-called active bed material that carries oxygen. The concept is referred as Oxygen-Carrier Aided Combustion, OCAC, which originates from Chalmers University of Technology and its research concerning Chemical-Looping Combustion, CLC [3]. The active bed materials consist of one or several metal oxides that reduce and oxidize repeatedly in fuel-rich and oxygen-rich environments, respectively. OCAC improves distribution and utilization of the oxygen inside the boiler followed by an overall improved performance [4]. Similar with the transition from grate-fired furnaces to FBC units, the possible implementation of OCAC through retrofit of existing FBC units is an essential feature to gain industrial acceptance. Hence, OCAC may constitute the next step in combustion technology development, Figure 2.1.

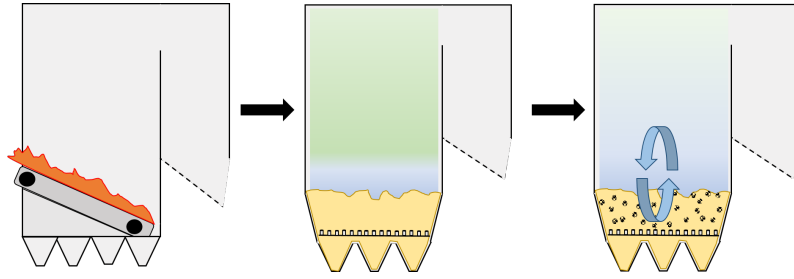


Figure 2.1: Transition in combustion technology. From left to right: grate-fired combustion, fluidized bed combustion and OCAC in a fluidized bed.

2.2 Combustion of solid fuels

Solid fuels are typically characterized by moisture content, ash content, volatiles and content of fixed carbon. Composition of solid fuels is estimated and determined by a proximate analysis. The moisture content indicates the amount of water bound in the fuel and the content of ash represents the share of incombustible inorganics that are left after full combustion. The combustible part of the fuel is either in gaseous or solid form, referred as volatiles and fixed carbon respectively. In contrast to a proximate analysis, an ultimate analysis determines the elemental composition of the solid fuel, i.e. share of carbon (C), hydrogen (H) and oxygen (O) etc.

The proximate analysis of a solid fuel is of high importance considering the propagation and evolution of the combustion process [2]. The combustion process is divided into three steps; drying, devolatilisation and combustion of the fixed carbon. The drying step represents the evaporation of the bound water while devolatilisation refers to the gasification of organics that release a combustible gas mixture. Left is the fixed carbon.

2.2.1 Combustion of solid biomass

Biomass combustion is considered as climate neutral with respect to the renewable nature of biomass, assuming that the pollution associated with transport and harvesting is neglected. This is the main reason behind the increased usage of solid biomass as alternative fuel to generate energy not only in Europe, but also worldwide [5, 6].

Compared with e.g. coal, solid biomass is considered as a difficult fuel type with reactive ashes, high volatile content, high moisture content followed by a low heating value [2, 7]. The high content of volatiles creates a challenge to ensure good mixing between the combustible gases and the combustion air.

If sufficient mixing is not achieved, undesired pollutants, i.e. partially oxidised compounds like carbon monoxide, are released.

Another challenge is the high content of reactive ashes. The incombustible compounds, i.e. ashes, consist of minerals originating from the original biomass including alkalies such as sodium, potassium, calcium and magnesium (Na, K, Ca and Mg, respectively) [7]. Presence of alkali enhances the risk of problems in the boiler; fouling and slagging [7–11]. Slagging is the result of deposits formed near the bed surface and on the furnace walls, exposed mainly to radiant heat. Fouling refer to deposits formed on the heat exchange surfaces. Slagging deposits are typically the result of formed alkaline silicates with low melting temperatures, e.g. 700-850°C (that of common alkali silicates) instead of 1450°C (that of quartz) [11]. As these compounds melt, deposits are formed and accumulated with time. These deposits may result in undesired defluidization of the bed (as a result of bed agglomeration), damaged equipment, increased costs related to maintenance and even forced shut-downs [9].

2.3 Fluidization technology

Fluidized beds increase the range of possible fuel qualities in a boiler, enhance mixing of solid fuel particles and air in addition giving a stable and effective heat transfer. In the hot bed, the fuel particles are sequentially dried, gasified and combusted.

A gas-solid system is the most common utilization of fluidization technology in FBC units. As the velocity of gas penetrating the solid bed increase, different regimes can be observed [12]. The first regime is referred as a fixed bed, where the gas velocity is low and the bed is unaffected. The second regime is called minimum fluidization where the drag force of the up-moving gas just equals the gravitational force acting on the bed particles. As the gas flow rate is increased further, regime of bubbling fluidization is reached.

The velocity that corresponds to the transition from a fixed bed into a minimum fluidization bed is called velocity of minimum fluidization, u_{mf} . Gas moving with this velocity suspends the bed particles in the gas and liquid properties of the bed is observed. As a result, lighter particles will end up on top of larger particles. This regime of minimum fluidization can be expressed as a force balance over the bed particles, Equation 2.1 [12],

$$\begin{aligned} \left\{ \begin{array}{l} \text{Drag force exerted} \\ \text{by the moving gas} \end{array} \right\} &= \left\{ \begin{array}{l} \text{Gravitational} \\ \text{forces} \end{array} \right\} \Rightarrow \\ \Rightarrow \Delta p_{bed} A_{cs} &= A_{cs} L_{mf} [1 - \epsilon_{mf}] (\rho_s - \rho_g) \frac{g}{g_c} \end{aligned} \quad (2.1)$$

where ϵ_{mf} represents the bed voidage at minimum fluidization, A_{cs} represents the cross sectional area and L_{mf} represents the bed height at minimum fluidization. Ob-

viously, the density of the solid bed particles have a great influence on the pressure difference required and hence the minimum velocity of fluidization.

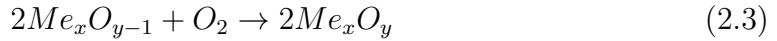
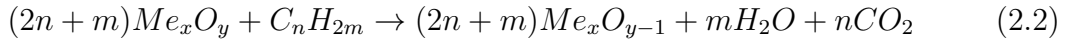
Naturally, if the gas velocity successively is increased for a fixed bed the pressure drop across the bed will to the same. However, as the velocity reaches u_{mf} the differential pressure over the bed stagnates [12]. u_{mf} can thus be determined experimentally for a specific bed if the pressure is registered as the flow rate of gas increases. To emphasise, this is valid for a bed with homogeneous particle size. If the bed consist of a wide particle size distribution instead, a segregated bed will be observed as the lighter particle will be fluidized before the larger particles. Then the bed will include a spectrum of different u_{mf} dependent on the particle sizes included.

2.4 Oxygen carrier aided combustion

The principle of oxygen carrier aided combustion, OCAC, is based on replacing the inert bed material with an oxygen-carrying bed material, OC. The whole bed can be replaced, or a fraction of it. The OC should be able to be reduced and oxidized repeatedly in presence or absence of fuel in the furnace, respectively.

The oxygen distribution in the furnace is improved as the OC provides oxygen to oxygen-deficit regions. The oxygen carrier can provide oxygen either directly to the fuel particle or through releasing oxygen in gaseous form. The latter is called oxygen uncoupling and is an important feature to improve the rate of char gasification in chemical-looping combustion, CLC, where the availability of oxygen is limited [13]. However, for OCAC operation, this feature is not as important.

For a general metal oxide used as oxygen carrier in complete combustion of a general hydrocarbon fuel, the reduction and oxidization reactions can be written accordingly;



The theoretical amount of oxygen that can be used for fuel conversion per unit oxidized oxygen-carrier on stoichiometric basis is called the oxygen transport capacity of the oxygen-carrier, R_O [14]. This capacity for a specific oxygen carrier is based on the change in mass related to the release of oxygen and is determined according to Equation 2.4.

$$R_O = \frac{m_{OC,ox} - m_{OC,red}}{m_{OC,ox}} \quad (2.4)$$

Here $m_{OC,ox}$ and $m_{OC,red}$ denote the mass of the oxygen-carrier in oxidized and reduced state, respectively.

2.4.1 The "perfect" oxygen carrier

For successful OCAC operation, the substitute for silica sand requires to resist high temperatures, prohibit good physical strength to avoid attrition, good fluidization behaviour, low tendency towards agglomeration and good redox reactivity [15]. Besides performance based properties, the availability, cost and after-use management is important. At Chalmers, effort is put on research related to evaluation of different materials and their viability as OC (both synthetically produced ones but also natural occurring ores [16–19]).

2.5 LD-slag: a byproduct from steel industry

The Swedish steel production is operated and regulated by the company SSAB. SSAB is the leading producer of steel in the Nordic region. SSAB Merox is responsible for the resulting waste and byproduct handling. As 4.93 Mton crude steel was produced during 2017, about 2 Mton of byproducts was produced as well. This great amount motivates the search of new areas of application. Using LD-slag, one of the major byproducts, as a potential oxygen carrier in fluidized bed combustion is one alternative.

The integrated and most common way of producing steel includes a blast furnace and LD-converter (named after the Linz-Donawitz process). In the blast furnace hot metal is formed by melting the raw material. The raw material is iron ore, either in form of pellets or natural ore. The resulting hot metal from the blast furnace, pig iron, is further converted batch-wise into steel in the LD-converter by introducing high-pressurized oxygen in the molten mixture. The content of carbon is lowered and the pig iron is converted into steel.

The amount of waste and byproducts formed during the production process is significant. For each ton of crude steel produced, roughly 400 kg of byproducts are formed for the integrated route (blast furnace and LD-converter) of producing steel. Roughly 90 wt-% of the byproducts is in the form of slags where LD-slag constitute about 30 wt-%. LD-slag is easily extracted from the LD-converter as the slag is lighter than the molten steel and stays on the surface of the steel. The slag is discharged and cooled off in casted pockets in the ground. The resulting solidified LD-slag is then put on storage.

Currently LD-slag have different areas of application. These include internally use in blast furnace as slag former, sealing and filling surfaces in landfills, road construction materials and raw material for rock wool production used for isolation.

As the LD-slag initially consist of large chunks, the material must be crushed and processed to the desired quality. For rock wool production, SSAB Merox delivers a product referred as "LD-slag 0/2" which is further used in this project. Depending on the choice of raw material (either ores, pellets or steel scrap) the composition of LD-

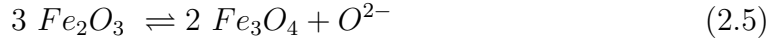
2. Background and Theory

slag can vary. But generally the composition of LD-slag 0/2 follows the specification presented in Table 2.1.

Table 2.1: Data specification of the product "LD-slag 0/2" delivered by Merox. The composition refers to weight-percentage.

LD-slag 0/2 delivered by Merox	
Composition	35.6 % CaO , 26.1 % Fe_2O_3 , 11.9 % MgO , 9.76 %, SiO_2 , 2.27 % MnO , 1.31 % TiO_2 , Al_2O_3 , P_2O_5 , Na_2O , K_2O < 1 %
Size interval	0 - 2360 μm
Size distribution	d10: 100 μm
	d50: 250 μm
	d90: 1800 μm
Bulk density	1510 kg/m^3

According to Table 2.1, iron(III) oxide is the main metal oxide present in LD-slag. Fe_2O_3 belongs to one of the first metal oxides investigated and studied for oxygen carrier purposes [20]. One possible intermediate oxidation path is from iron(III) to iron(II,III) oxide. The redox formula and the theoretical oxygen transport capacity for the given pair is presented in Equation 2.5 and 2.6, respectively.



$$\left\{ \begin{array}{l} m_{OC,ox} = n_{OC,ox} M_{OC,ox} = 159.687g/mol_{OC,ox} \\ m_{OC,red} = n_{OC,red} M_{OC,red} = 154.354g/mol_{OC,ox} \end{array} \right\} \Rightarrow \quad (2.6)$$

$$\Rightarrow R_O = \frac{m_{OC,ox} - m_{OC,red}}{m_{OC,ox}} = 0.0334$$

The viability of using LD-slag as an oxygen carrying bed material have been investigated by Chalmers, both in a 12 MW_{th} CFB boiler but also in small scale CLC pilot plants as well, 300 W_{th} and 10 kW_{th} , respectively [21, 22]. Both complete and partly substitution of the bed inventory with LD-slag was studied. LD-slag performed well during operation as biomass was used as fuel, with no attrition nor agglomeration. It showed good physical strength and the handling of LD-slag was not different compared with sand. Increased bed temperatures were observed and a change in temperature profile over the boiler as well, indicating on the oxygen buffering effect of LD-slag. As LD-slag did not absorb ashes in contrast to silica sand, a mixed bed inventory of LD-slag and silica sand was suggested to be preferable.

Chapter 3

Methodology

This thesis is a qualitative investigation, partly based on quantitative data, of the possibility and consequence of introducing oxygen-carrying material in the fluidized bed in the industrial boiler Sävénäverket. The objective was to identify and minimize the risks for a possible future experimental campaign.

3.1 Qualitative analysis

To perfectly forecast what will occur and how different phenomenon affect another when changing the bed inventory composition is near the limit of impossible. However, based on an overall description of the existing process, presented in Chapter 4, important subsystems that are likely to change when implementing an active bed material was studied. The subsystems and parameters that were used to evaluate the specific prerequisites for an experimental campaign in the industrial boiler are presented below.

- Bed material handling (delivery, storage and feeding regulation).
- Bed particles behaviour during operation.
- Ash management.
- Fluidization properties of the bed material.
- Bed temperature variation and regulation.
- Release of combustion emissions.
- Process air regulation.

3.2 Quantitative data collection

The collection of quantitative data was related to the comparison of physical properties for the different bed materials. The physical properties of LD-slag are related to the fluidization behaviour and resistance to sintering at high temperatures.

To determine LD-slag's resistance of agglomeration at high temperatures, the sintering temperature was measured by the Research Institutes of Sweden (RISE). Three samples were sent for analysis presented in Table 3.1. These include two samples of "used" LD-slag and one sample with fresh LD-slag. The used LD-slag was taken from stored bottom ash samples originating from the Chalmers 12 MW_{th}

CFB boiler. The bottom ash samples were originally taken during an experimental campaign in late autumn 2017 when LD-slag was tested as bed material [21]. During that campaign the silica sand in the bed was both completely and partially (30 wt-%) replaced by LD-slag. The sintering temperature of these two bottom ash compositions were determined by RISE.

Table 3.1: Samples sent for the agglomeration and sintering test performed in a small lab reactor operated by RISE.

Name	Material	Comment
LD	LD-slag	Novel/Fresh material.
LD30	LD-slag	Used material. 30/70 wt-% mixture of LD-slag and silica sand.
LD100	LD-slag	Used material. 100 wt-% LD-slag as bed material.

At RISE laboratory, the samples were placed in a lab reactor where they were fluidized while gradually increasing the temperature. As the temperature reached 700°C, a step-wise temperature increase by 3.5°C per minute was practised. The maximum temperature of the test was 110°C. Variation in differential pressure over the bed was used to observe and determine any tendencies of sintering.

To derive similar fluidization behaviour of the active bed material as the currently used silica sand, the difference in density and particle size distribution was investigated. This enabled to derive a preliminary desired size distribution of the LD-slag. The final LD-slag with desired size distribution together with the ordinary LD-slag was in the end fluidized and compared with the bottom ash. The different derived mixtures used in this project, and corresponding nomenclature, is presented in Table 3.2 and three of them are shown in Figure 3.1.

Table 3.2: Nomenclature used for the different mixtures discussed in this project.

Name	Comment
B95	Technical specification for the currently used silica sand.
Fresh sand	Sample of silica sand taken from the sand silo.
Bottom ash	Sample taken from the bottom ash.
LD-slag 0/2	Original and unprocessed material after delivery.
LD-slag 355	Sieved LD-slag 0/2 with fines <355 μm removed.

3.3 Experimental campaign strategy

It was determined that a mixture of LD-slag and silica sand was best suited for this campaign. As the industrial boiler is equipped with one sand silo, the resulting strategy was to introduce the LD-slag batch-wise in the bed inventory. This imply



Figure 3.1: The different mixtures extracted and investigated throughout the project. Currently used silica sand to the left, original LD-slag 0/2 to the right and LD-slag without fines, LD-slag 355, in the middle.

that the concentration of LD-slag in the bed inventory will increase with time until the loaded batch in the sand silo is consumed and original silica sand is introduced again.

3.3.1 Campaign objectives and expected results

Desired objectives and estimated results for a successful experimental campaign at Sävenäsverket are presented below.

- O1. To register increased heat development in the bed as the oxygen carriers allow combustion to occur in the bed.
- O2. To evaluate the smoothing effect of the active bed material on the cross-sectional variation in bed temperature.
- O3. Observe if the new bed composition affect the release of combustion emissions.
- O4. Analyse bottom ash samples to map any changes in the bed inventory related to agglomeration.
- O5. Register if the frequency of the pneumatic fly ash sender is changed (which imply that the share between bottom- and fly ashes is changed).
- O6. Investigate if lower surplus of air is possible compared with silica sand without having increased emissions of non-combusted compounds.

3.3.2 Bed inventory composition

To determine batch size of LD-slag for the campaign it was essential to estimate how the bed composition change with time. By approximating the bed inventory as a perfectly mixed bed with homogeneous concentration, and determining the in- and out streams as continuous flow rates, the mass balances of the bed components included was formulated, Equation 3.1.

$$\begin{cases} \frac{dx_1}{dt} = \frac{\dot{m}_{1,in} - x_1 \cdot \dot{m}_{out}}{\dot{m}_{out}} \\ \frac{dx_2}{dt} = \frac{\dot{m}_{2,in} - x_2 \cdot \dot{m}_{out}}{\dot{m}_{out}} \\ \vdots \\ \frac{dx_n}{dt} = \frac{\dot{m}_{n,in} - x_n \cdot \dot{m}_{out}}{\dot{m}_{out}} \end{cases} \quad (3.1)$$

$x_{1,2,...,n}$ represents the share of different components in the bed inventory. For the experimental campaign three components was considered (silica sand, bottom ash originating from the fuel and LD-slag). The share of each component is dependent on the total mass of the bed inventory. During normal operation in the industrial process the extraction of bottom ash is regulated to maintain a constant level of bed material. This imply that the extraction rate equals the intake rate, i.e. $\Delta m_{tot,inv} = 0$, referring to Equation 3.2.

$$\Delta m_{tot,inv} = \dot{m}_{out} - \sum_{i=1}^n \dot{m}_{i,in} \quad (3.2)$$

According to Equation 3.1, the composition of the bed inventory is regulated by three parameters; intake of ashes originating from the fuel, feed of bed material and the bottom ash extraction rate. By defining these for different operation scenarios presented in Table 3.3, the time dependent composition of the bed inventory could be derived. The resulting time varying bed composition provided important information to further construct an experimental campaign. The target bed inventory composition was set to reach at least 30/70 wt-% LD-slag/silica sand respectively (excluding the bottom ash).

Table 3.3: Definition of the given flows for different scenarios related to the realization of the campaign.

	Normal operation	Forced extraction
Bed inventory	80 ton	80 ton
Fuel intake	11 kg/s	11 kg/s
Bed material intake	330 kg/hr	760 kg/MWh
Extraction rate	440 kg/hr	900 kg/hr

According to Table 3.3, the extraction rate of bed material is twice as high for the forced extraction scenario compared with normal extraction. As the extraction rate is higher, the feeding of bed material is increased as well to maintain constant bed level. Since the fuel intake is constant for both scenarios, the feeding rate of bed material is increased more than 100% for the extraction rate scenario.

Chapter 4

Description of the industrial boiler

In this chapter the essential functioning of the internal systems of the boiler are presented, with focus on the combustion-related systems. Considered internal systems are; fuel handling, process air, flue gas cleaning, ash and bed material handling.

4.1 Boiler overview

The industrial boiler located in Sävenäs and operated by Göteborg Energi has a major role in supplying district heat to Gothenburg. The boiler has a long and rich history of implied reconstructions and retrofits through the years since the installation in 1985. Göteborg Energi has a production mix of several facilities including external sources of heat utilized in the district heat network. Out of these facilities, the boiler in Sävenäs is among the first in the priority order and is supplying hot water during the whole winter season, early October to late May, more than 4000 hours per year [23].

The boiler is a combined heat and power production facility. It is a bubbling fluidized bed with installed thermal capacity of 95 MW. To fluidize the bed particles and keep them suspended, air is injected in the bottom of the furnace. The boiler is fired with biomass mainly consisting of mill and logging residues (referred as “GROT”). Share of each fuel type for the year 2017 is presented in Figure 4.1. This imply that fuel is non-homogeneous with varying particle sizes and fuel quality on a daily basis.

The boiler is equipped with a 14 MW steam turbine which convert saturated steam of 20 bar into electricity. Since the turbine is fed with saturated steam, the boiler is referred to a saturated steam boiler instead of a traditional superheated steam boiler. This imply that the boiler is not equipped with a superheater. However, the steam turbine is not continuously in operation, but is committed to run during periods when the demand for electricity is high and/or demand for heat is low.

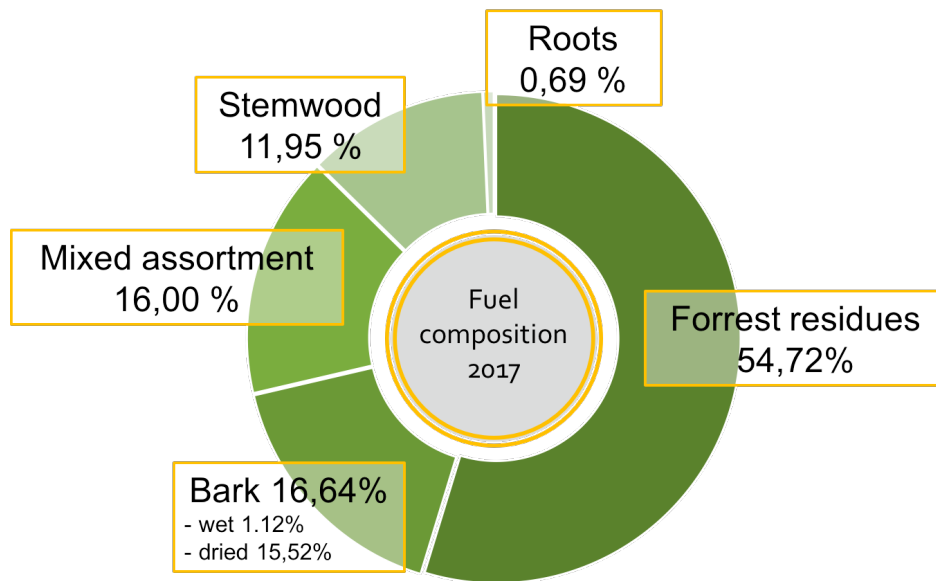


Figure 4.1: Visual representation of the fuel composition based on MWh retrieved during 2017

Consequently, there are two possible operational modes for the boiler; hot water mode or steam generation mode. During hot water mode the input of air and fuel is regulated by the pressure level in the steam dome where the steam outtake is regulated by the power output on the district water. In the steam generation mode, saturated steam for electricity generation besides hot water is supplied and the pressure is regulated by the steam turbine and condenser instead of the steam dome. The power to heat ratio, α , is usually 0.15.

To utilize the hot flue gases in the furnace leaving the fluidized bed, a primary and secondary economizer are located in the convection part of the furnace. Tube water is heated in the primary economizer, while district water is heated in the secondary economizer. As the flue gas leaves the convection path of the furnace, the gas is cleaned off particulates. The cleaned flue gas can either be recirculated back to the furnace or condensed in the final flue gas condenser before being released in the stack. The flue gas condenser makes it possible to account for the higher heating value of the incoming fuel which adds 20-25 MW of heat to the district water besides the thermal output of the boiler. The return temperature of the district water is typically 40°C and is successively heated up to a supply temperature depending on the outdoor temperature with a flow of typically 400 kg/s. The supply temperature is generally within the range 95-110°C.

Accompanied with biomass combustion, ashes are formed that have to be collected and treated continuously. Larger fractions of ashes, together with bed material, is extracted from the bottom of the furnace and stored in bottom ash containers. These are further transported by truck for after treatment. Smaller fractions of ashes, smaller particulates, is retrieved and collected from the flue gas cleaning steps including an electrostatic precipitator (ESP) and bag house filters. As the

small particulates are collected, they are first sprayed with water to bind the dust before transported by truck for later treatment. Schematic picture of the boiler is presented in Figure 4.2.

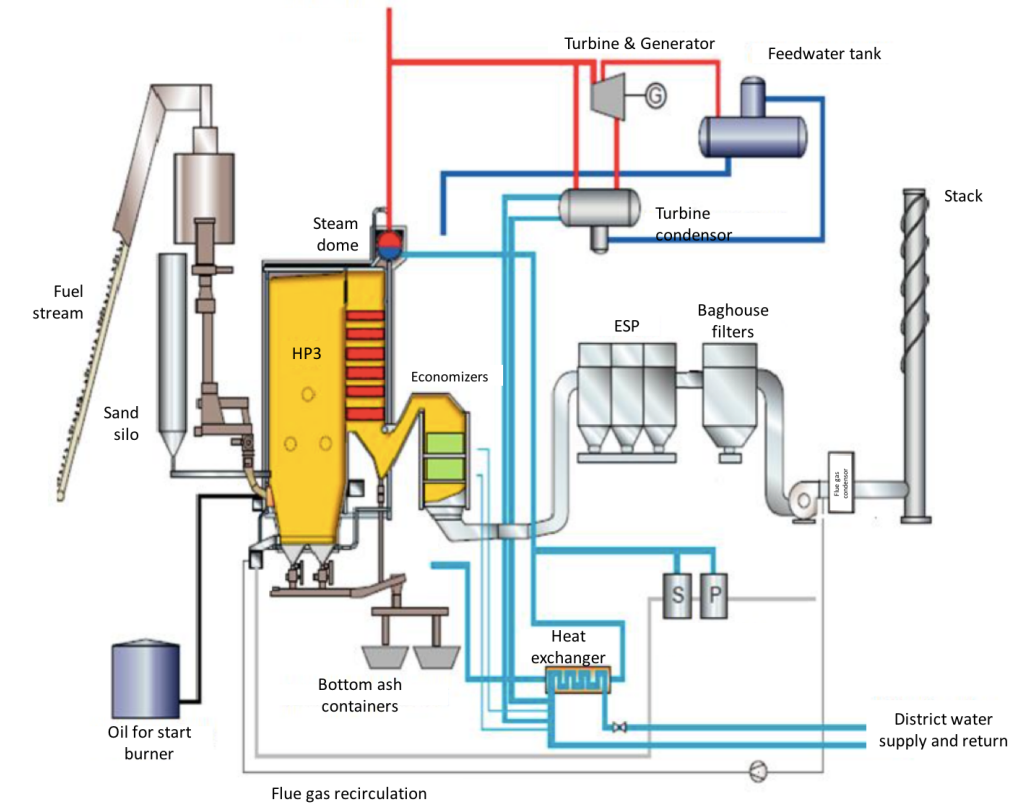


Figure 4.2: Schematic overview of the heat and power production plant of Sävenäs HP3. *Source: Göteborg Energi.*

4.2 Furnace design

The furnace have a rectangular shape, about $9.4 \times 4.3\text{m}$, with a cross sectional area of 40.83 m^2 . There are four fuel intakes placed on the front long side wall next to the fresh bed material inlet. The primary air is introduced through the HYBEX-system in the bottom of the furnace while there are four inlets for secondary air on each long side and four inlets for tertiary air on each short side according to Figure 4.3. Below the HYBEX system there are in total eight bottom ash collection pockets.

4.3 Fuel handling

To meet the desired output of heat and electricity, right amount of fuel must be introduced. During full load 11 kg of fuel is consumed each second. The fuel handling system distinguishes between outer and inner management of the incoming fuel. The outer handling system is responsible for retrieving and regulate the size

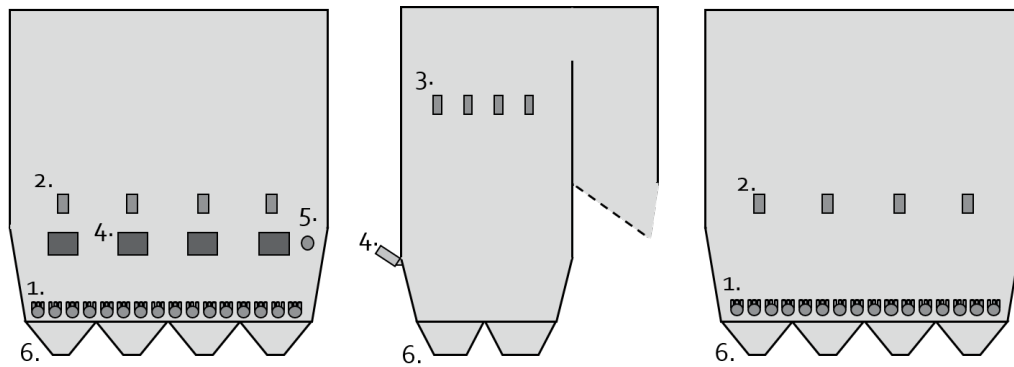


Figure 4.3: Simplified schematic of the furnace design. From left to right: front-, side- and back wall respectively seen from the outside. 1. Primary air, 2. Secondary air, 3. Tertiary air, 4. Fuel inlet, 5. Sand inlet, 6. Bottom ash pockets.

and content of the fuel before storing it in silos. The inner fuel handling system on the other hand is responsible for the distribution and intake of fuel into the furnace. The two subsystems of fuel handling is presented more in detail below.

The outer fuel handling system receives roughly 40 loaded trucks with fuel each day. The amount of fuel is weighted and its moisture content is measured. The fuel is collected in a collection pocket where a hydraulic dispenser moves and extract the incoming fuel. The fuel stream is processed to remove the content of metal and to mill too large fractions before storing it in silos. This is done by a rotating metal separator and a laminar sieve (where the larger fraction is milled by two rotating crushing wheels). The plant is equipped with two weekend stock fuel silos that each hold 5000 m^3 that can cover the fuel demand for two-three days during full load. Fuel is continuously extracted from these silos and transported by a belt conveyor to the inner fuel handling system.

The inner fuel handling system includes two intermediate daily stock silos that serve to even out disturbances in the fuel feeding system. The daily stock silos can supply the demand for one hour and the extracted fuel stream is evenly distributed over four fuel feeders. Each fuel feeder is further providing a fuel chute with fuel that is further introduced inside the furnace just above the fluidized bed.

4.4 Process air

The total intake of process air is regulated by a total air fan. The process air is divided into primary air and carrying air. Carrying air represents secondary and tertiary air in addition to the air supplied for the Selective Non-Catalyst Reduction system (SNCR). The primary air is introduced in the bottom of the furnace to fluidize the bed while the supplementary air is introduced above the bed at different levels, Figure 4.3. The intake and distribution of process air adapts the boiler load. A simplified regulation scheme of the process air is presented in Figure 4.4. During full load the total intake reaches $140\,000 \text{ Nm}^3/\text{h}$.

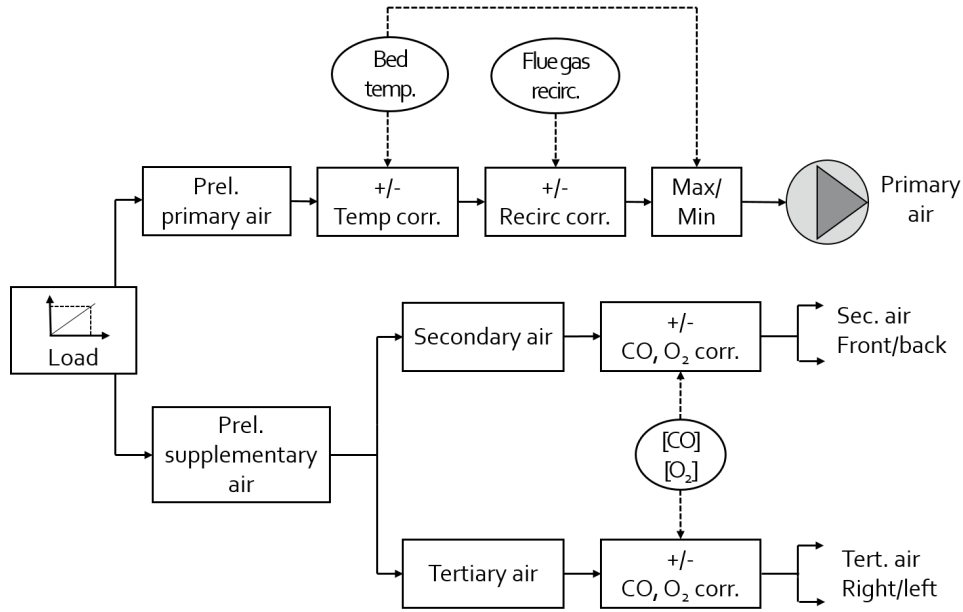


Figure 4.4: Simplified schematic of the regulation of primary, secondary and tertiary air flow rates, respectively.

4.4.1 Primary air

Approximately one third of the total process air flow is directed as primary air. The primary air is preheated above 100°C before it is injected through small nozzles in the bottom of the furnace, referred as the HYBEX system. The primary air is governing for the temperature in the bed, oxygen supply and the flue gas velocity. Too low flow rate of primary air will result in defluidization and a too high insufflation of air will cause undesired particle entrainment.

The flow rate is initially determined by the desired load signal and further up- or down regulated by the registered bed temperature and flow of recirculated flue gases. The flow rate is then limited to be within the predefined maximum and minimum velocity limits, visualized in Figure 4.4.

Not only the flow rate of primary air is of importance but also the air quality considering the moisture and oxygen content. Three alternatives are possible; introducing dry air, introducing air with higher moisture content and introducing recirculated flue gases to decrease the content of oxygen. Air with increased moisture content originates from the rotating heat exchanger in the flue gas condenser. A higher content of moisture has a cooling effect on the bed. In contrary, the dry primary air originates from preheated atmospheric air and is contributing with increased input of oxygen per volume of air (compared with wet air). Recirculation of flue gases and mix it with primary air is practised when the bed temperature is too high. This limits the supply of oxygen introduced in the bed which is followed by a drop in temperature.

The minimum and maximum velocity constraints for primary air is related to the minimum gas velocity to keep the bed fluidized and maximum gas velocity to avoid particle entrainment and deposits on the furnace wall. The upper limit is a trade-off between increased supply of oxygen and increased risks of deposits on the furnace walls induced by splashing of the air bubbles as they break the bed surface. As the deposits eventually fall down to the bed again, they adversely affect the fluidization quality of the bed. The maximum gas velocity is currently set to 1.36 m/s. This velocity corresponds to different flow rates for different temperatures as the flow rate is measured in normal cubic meters. The corresponding primary air flow rates derived from the application of thermodynamic processes for ideal gases is presented in Figure 4.5 together with the lower limit as well. The lower limit is defined for different boiler loads. The limit presented in Figure 4.5 is taken for a load signal of 100% (i.e. the highest lowest limit).

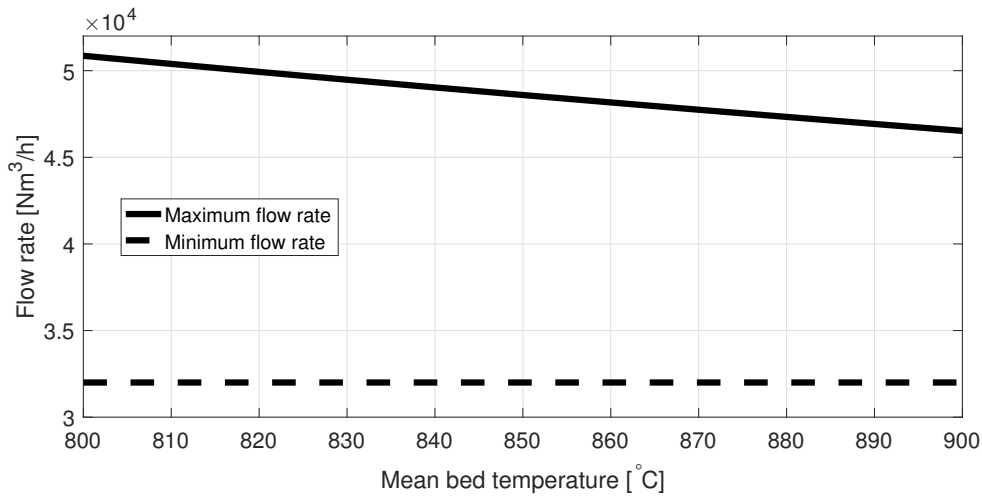


Figure 4.5: Mean bed temperature's impact the upper and lower limit of primary air flow rate. The maximum limit adapts to bed temperature to avoid exceeding the predefined upper limit of gas velocity in the bed while the lower limit is regulated by the load signal instead.

4.4.2 Secondary and tertiary air

The objective of introducing supplementary air, i.e. secondary and tertiary air, is to ensure full combustion of the flue gases in the furnace leaving the bed. Before entering the boiler, the supplementary air is preheated electrically but also by exchanging heat with the flue gases in the flue gas condenser to a temperature of 80°C.

The initial amount of supplementary air to deliver is determined by the actual load. Then, the flow rate is corrected by the registered level of oxygen (O_2) and carbon monoxide (CO) in the flue gases leaving the convection section. If the registered level of O_2 in the flue gases is too low, or the registered level of CO is too high, additional supplementary air is introduced and vice versa. The corrected flow rate

of secondary air is then equally distributed over the eight air feeders, four feeders on front and back side wall respectively. The same holds for tertiary air but on the right and left side wall instead, Figure 4.4.

4.5 Flue gas passage

The flue gas passage includes the convection path of the furnace, different cleaning steps, heat recovery in the flue gas condenser and finally the exhaust by the stack.

As the flue gases leaves the radiation part of the furnace it carries a temperature about 950°C. The total flow rate of flue gases exceeds well above 190 000 Nm³/h during full load. To recover this great amount of heat, the hot gases initially enters the convection part of the furnace, see Figure 4.6. Here the temperature drops as the flue gases passes through large tube packages where boiler water is heated. Further the flue gases are cooled by exchanging heat with the primary and secondary economizer. As the flue gas leaves the convection part, the temperature have gradually decreased to approximately 140°C.

The flue gases contain relatively high content of water vapour, almost 25%. The water originates from the high moisture content of the incoming fuel and water vapour formed during combustion. In the flue gas condenser, located after the flue gas cleaning, the heat of vaporization of water is utilized by condensing the water vapour in the flue gas. This is done by decreasing the temperature of the flue gases down to roughly 35-40°C by exchanging heat with cold return district water and primary air, forcing the water vapour to condense. Consequently, in the flue gas condenser the temperature is efficiently decreased from 140°C to 40°C besides condensing the water vapour. At this point the flue gas temperature and hence quality of heat is too low for further heat exchange and is therefore released by the stack.

Besides heat recovery, the gas is cleaned. The different cleaning procedures installed includes the Selective Non-Catalytic Reduction (SNCR), Electrostatic Precipitator (ESP) and baghouse filters. Together these control and regulates the content of nitrogen oxides, dust and small particulates in the exhaust gases. The nitrogen oxides is already taken care of inside the furnace, by the SNCR system. Here ammonia is injected to react with the nitrogen oxides to form nitrogen and water. The ESP and bag house filters separates the dust and small particulates (i.e. entrained fractions of bed particles and light-weighted fuel compounds) from the flue gases. Additionally, the flue gas condenser support the dust separation as the condensed water efficiently bind the eventually remaining dust after the filters.

Emissions and temperature is measured along the whole flue gas passage. As the gas exit the furnace, levels of oxygen and combustion emissions (CO, SO_x, NO_x etc.) are registered before entering the ESP. Further, after the ESP, emissions of sulphur and nitrogen oxides is registered as well in addition to oxygen and carbon monoxide. However, most of the emission measurements are done in the stack where a Fourier

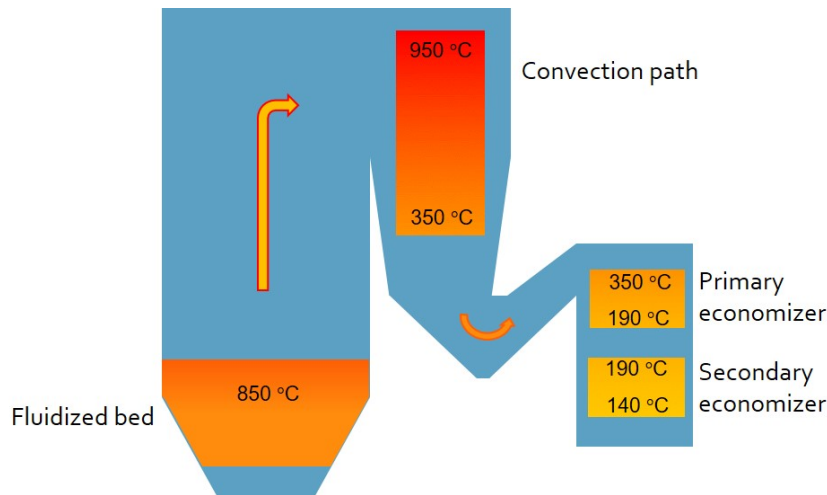


Figure 4.6: Overview of the gradually decrease in temperature of the flue gases in the convective part of the furnace.

Transform Infrared Spectroscopy instrument (FTIR) is installed. The FTIR measures the content of oxygen, nitrogen- and sulphur oxides, in-complete combustion compounds, dust and ammonia. The temperatures along the passage are registered in the top of the furnace, before and after the economizers, after and between the ESP and baghouse filters, before the flue gas condenser and inside the stack.

4.6 Ash handling

During operation it is crucial to manage the produced ashes by collection and storage. The ash originates from both fuel and bed particles. The ash formed during combustion is divided into two waste streams; bottom ash and fly ash. These ashes are sought to be utilized in the best environmental way possible. No traditional landfill have been practised during recent years for the produced ashes. Instead, since biomass has a high content of nutrients, ash spreading in the forest to enable nutrient recirculation have been done. However, because of the high content of bed material in the bottom ash it is only the fly ash that have been used for nutrient recycling. Bottom ash on the other hand have been used for soil improvement and soil production for covering top-layer for landfills.

4.6.1 Bottom ash

In the bottom of the furnace eight water-cooled water packets are located. Here the heavier fractions of fuel ashes together with bed particles are collected. These pockets are equipped with pneumatic regulated discharge gates that control the extraction of bottom ash. The extraction of bottom ash is either directed directly to containers for removal or to a rotating sieve. The sieve separates bed material from ashes to enable recirculation of bed material back to the furnace.

Consist mainly of bed material, but also include larger fractions of the inorganic compounds that originate from the fuel. During 2017, out of 2687 ton of bottom ash extracted only 681 ton originated from the fuel. This represent 25 wt-% while the rest consist of bed material.

4.6.2 Fly ash

Small fractions of ashes and bed particles are entrained by the moving air along the furnace. These are finally collected at the ESP (collecting larger particulates) and at the bag-house filters (collecting smaller particulates). The resulting waste collected in the bottom of the ESP and bag-house filters is transported to a fly ash silo by compressed air in the ash sender. Water is sprayed on the fly ashes before loaded on trucks to bind the dust. 1957 ton of wet fly ash was collected during last year, 2017, with a moisture content of roughly 25 wt-%. Consequently, 1468 ton of dry fly ash was collected which originates from the fuel alone. Because of its high content of nutrients it is transported to a treatment company which enable the nutrient-rich fly ashes to be recirculated back to forest ground and hence close the loop for the nutrients.

4.6.3 Mass balance estimation

To establish an estimation of ash content in the delivered fuel, a mass balance for the boiler is set up. Data is used for the years between 2015-2017. The inflow of solid particles is represented by the intake of biomass considering its ash content and inflow of sand material. The outflow of solid particles is represented by the bottom ash extraction together with the material collected by the ESP and bag house filters. The loss of particulates with the exhaust gas is neglected. The actual in- and out flow of solid particles are presented in Table 4.1.

Table 4.1: Mass balance for the boiler between 2015-2017. The amount of bottom ash includes both ash and bed material collected in the pockets. The amount of fly ash is presented in dry basis based on a moisture content of 25 %.

	2015	2016	2017
Solid fuel [<i>ton wet basis</i>]	160 770	145 874	158 886
Sand consumption [<i>ton</i>]	1878	1864	2006
Bottom ash [<i>ton</i>]	2647	2504	2687
Fly ash [<i>ton</i>]	1512	1454	1468
Estimated ash content in fuel [<i>wet basis</i>]	0.0142	0.0144	0.0135
Estimated ash content in fuel [<i>dry basis</i>]	0.0256	0.0258	0.0245

As can be seen in Table 4.1, the amount of fuel and sand introduced are somewhat equal over the years and greater share of the fuel ashes end up as fly ash.

The amount of bottom ash collected consist of both ash and bed material. By subtracting the amount of bottom ash with the total intake of sand over the year, the amount of fuel ashes that end up as bottom ash is achieved. In contrast, all fly ash particles can be assumed originating from the fuel. This because larger fractions of solids in the flue gases are separated and captured right before the economizers and directed to the bottom ash, Figure 4.2. However, bed material could be damaged and break during operation which results in fine particles that can be carried away with the flue gas. But such tendencies have not been observed. Hence, the amount of fly ash originates from fuel ashes alone while bottom ash consist of a mixture between ashes and bed material (mostly bed material).

4.7 The fluidized bed

The bed must withstand high temperatures and resists physical abrasion. During operation the bed temperature is within 800-900°C, typically 850°C and the penetrating flow of primary air causes stress on the particles as they collide with each other and the furnace walls. But, during combustion of biomass the bed material react with the ashes and enriches substances that may lower the sintering temperature of the particles. Hence, to avoid bed agglomeration continuous regeneration of the bed is required. The consumption of bed material in the boiler is dependent on the quality of the solid fuel, load, bed temperature and quality of the sand.

4.7.1 Bed material

Bed material used at Sävenäsverket is silica sand. The sand supplier is Baskarp/Sibelco, and the quality of the sand delivered is referred to "B95". Since the boiler type is BFB, larger particles are used in comparison with e.g. a CFB. Particle diameter is in range of 1 mm. The technical specification and composition of the sand used is presented in Table 4.2 and 4.3, respectively.

The sand is currently delivered by truck and fed at a pressure around 40 bar into the sand silo. The sand silo is placed close to the front wall of the furnace where the bed material is introduced. The sand silo has a cylindrical shape with a conical bottom. The diameter of the silo is 1.97 m and the total height of the silo is 14.12 m, where the cone constitute 1.54 m. This result in a total volume close to 40 m³ that hold 50 ton of silica sand. A screw conveyor is located at the bottom of the silo, feeding fresh sand to the furnace.

4.7.2 Bed inventory

The furnace contain 80 tons of bed material and the furnace is fed with 3.5 kg sand per MWh. Hence, during full load roughly 330 kg of fresh sand is fed each hour. The extraction of bed material is regulated by the outtake of bottom ash. The average

Table 4.2: Data specification of the silica sand used as bed material delivered by Baskarp referred as 'B95'. The values of d10, d50 and d90 are used to characterize the distribution of particle diameters. E.g. 90% of the sand particles have a particle diameter less than 1.6 mm.

Data specification traditional silica sand 'B95'	
Size range	177 - 2.8*10 ³ μm
d10	390 μm
d50	900 μm
d90	1.61*10 ³ μm
Sintering temperature	1270 °C
Bulk density	1500 kg/m ³

Table 4.3: Composition of the silica sand used as bed material delivered by Baskarp referred as "B95". Presented as weight-percentage.

Composition of silica sand 'B95'	
SiO ₂	90.5 %
Al ₂ O ₃	4.9 %
Fe ₂ O ₃	0.5 %
K ₂ O	2.0 %
Na ₂ O	1.2 %
Other trace elements	0,9 %

residence time in the furnace for a sand particle is 242 hours. However, the whole bed inventory is not actively fluidized. There is a dead volume in the bed below the primary air injection in the bottom ash extraction pockets. Here the bed is packed, i.e. inactive, instead of fluidized, see Figure 4.7. As the fixed bed height, i.e. the height of bed material above the air nozzles when not fluidized, h_{fixed} is given as well as the cross-sectional area of the furnace, A_{cs} , and density of sand, ρ_{sand} , it is estimated that 50% of the total bed inventory is active or inactivate, respectively, according to Equation 4.1.

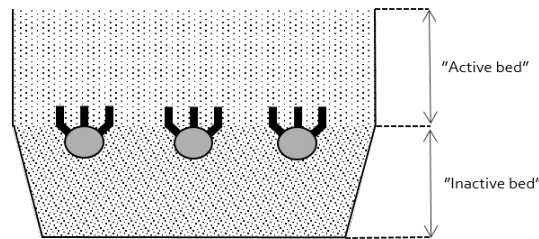


Figure 4.7: Simplified picture of the primary air nozzles in the bottom of the furnace. An in-active part of the bed is located below the nozzles.

$$\frac{m_{\text{active bed}}}{m_{\text{tot. bed inventory}}} \approx \frac{A_{cs} * h_{fixed} * \rho_{sand}}{m_{\text{tot. bed inventory}}} = \frac{39\,838.5\,kg}{80\,000\,kg} = 49.8\% \quad (4.1)$$

To decrease the consumption of sand, a rotating sieve have been installed. The objective of the sieve is to separate sand particles out of the bottom ash to be reintroduced in the furnace. The working principle of the sieve is illustrated in Figure 4.8. More fine and coarse particles extracted from the sieve are removed and collected as bottom ash while the fraction in between is recirculated back to the furnace.

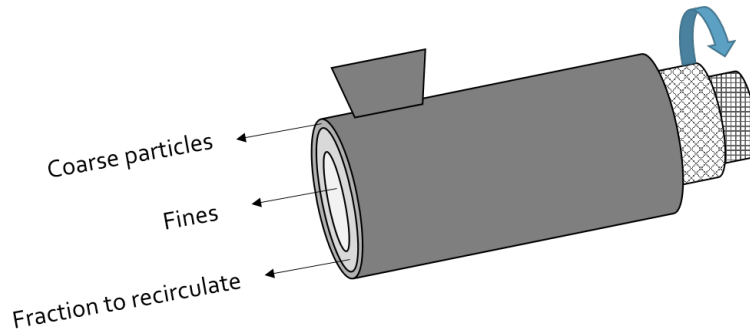


Figure 4.8: Schematic of the existing rotating sieve installed to decrease the consumption of fresh bed material.

To regulate the in- and outtake of bed material to maintain a certain level of bed material inside the furnace, the differential pressure over the bed is measured. The differential pressure meter is measuring the difference in pressure right above the air nozzles and over the bed. The reference differential pressure over the bed is 65 mbar. Hence, by comparing the registered differential pressure over the bed with the reference value it is possible to observe and regulate the level of bed material inside the furnace. A too high differential pressure indicate an increase in amount of bed material which imply that the intake of sand is greater than the outtake and vice versa for a lower differential pressure. The extraction of bed material typically follows the same magnitude as the inflow, 3.5 kg/MWh, but forced outtake can be practised as well where roughly 10-11 kg/MWh extraction is possible. Hence, the differential pressure is the most important parameter to use for controlling the amount of bed material.

4.7.3 Bed temperature

To observe and regulate the bed temperature is of great importance. Too high temperature in the bed for a longer period of time will increase the risk of sintering of bed particles, and a too low bed temperature will lower the rate of combustion. Hence, to keep track of the temperature, eight temperature sensors are located in the bottom of the bed. They are located 40-50 cm above the primary air nozzles and spread over the cross-section according to Figure 4.9. During operation, two mean values of these eight temperature sensors are used as input for different regulation schemes. The two mean bed temperatures used are "T4 max" and "T6 mean". T4 max is a mean value of the four highest temperatures registered out of the eight sensors and T6 mean is a mean value of the temperatures except the lowest and

highest one. Observing the registered temperature of each sensor it is clear the bed is exposed to irregular temperature variations. The bed temperature is significantly lower at the front wall where the fuel is fed. This indicates that the fuel is not distributed over the whole bed, but stays near the front wall instead and accordingly cools the bed because of the fuel's moisture content. By extracting the mean value of each sensor after one week of operation a temperature map of the bed can be constructed. The result is presented in Figure 4.9.

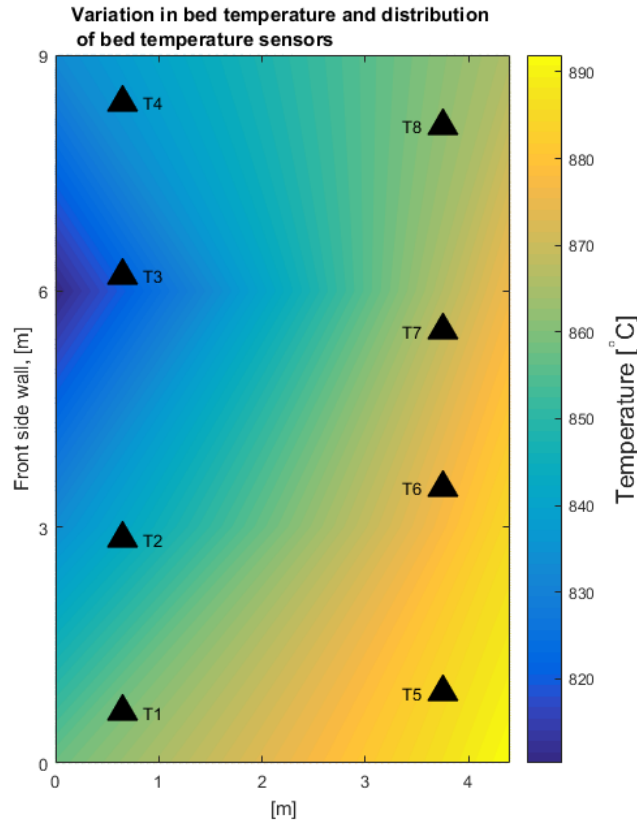


Figure 4.9: Overview of the temperature sensors located in the bed and the registered temperature variation. Sensors T1-T4 are located near the front wall of the furnace where the fuel is fed. The fuel's moisture content is contributing to the lower temperature compared with the opposite wall.

The temperature of the bed is dependent on the amount of oxygen and quality of fuel that is introduced. The principle of bed temperature regulation by controlling the flow rate and quality of the primary air have already been presented in Section 4.4. But, the quality of fuel is not possible to control in the same way. The fuel quality is dependent on the truck delivery and is fluctuating on daily basis. The content of moisture and dry matter of the fuel can only be controlled to a certain degree because of the limited possibilities of mixing the incoming fuel. Hence, fluctuating quality of the fuel is inevitable and must be met by regulating the process air. The already mentioned measures to be taken in order to control the bed temperature

can therefore be summarized accordingly;

- By regulating the flow rate of recirculated flue gases.
- By regulating the flow rate of primary air (including the regulation of load signal and hence fuel intake).
- By introducing either wet or dry primary air.

4.7.4 Quality of fluidization

The differential pressure over the bed is important for other reasons but amount of bed material as well. The differential pressure give valuable information regarding variations in sand quality that could be caused for instance by agglomeration. The problem of sintering followed by agglomeration creates larger and heavier particles which can be observed by an increase of differential pressure over the bed. This imply that an increase of differential pressure indicate that increased extraction of bottom ash and bed material should be practised to maintain not only constant level of bed material, but also to maintain good quality of the bed particles.

Other indications of sintering and agglomeration can be temperature variations over the bed as the temperature tend to decrease at locations where the particles melts and may prohibit oxygen to penetrate into the bed through the bottom nozzles. As the particles agglomerate, the rate of fluidization is decreased as well because of heavier bed particles. This should be followed by less efficient combustion and hence give a modest increase in emissions of non-combusted compounds. These trends should be observed to avoid major break downs that require the boiler to be shut down in order to replace the bed material.

Chapter 5

Results

This chapter presents the results from the quantitative data collection, qualitative analysis and the final strategy of an experimental campaign. For further discussion and clarification of the results, the reader is referred to Chapter 6, Discussion.

5.1 Quantitative data collection

The quantitative data collected includes material sintering temperature, determination of particle size distribution of fresh silica sand, bottom ash and LD-slag respectively and the fluidization behaviour of bottom ash and LD-slag. These are presented below.

5.1.1 Sintering temperature

As the temperature is increased while fluidizing each of the samples, the differential pressure over the bed is decreasing slowly. However, as sintering starts to develop the differential pressure decreases more rapidly. Complete agglomeration is observed as the differential pressure over the bed sample reaches bottom. The temperature where complete agglomeration occurs is referred to as T_3 . Before reaching complete agglomeration, two additional temperatures are registered to map when the rate of change in differential pressure is observed. This first and second increased rate of change, i.e. tendency towards sintering and agglomeration, are referred to as T_1 and T_2 respectively. T_1 , T_2 and T_3 are registered for each of the samples tested in the lab reactor. The derivation of the three temperatures for sample LD30 is presented in Figure 5.1 while the summarized results for all three samples together with one sample of silica sand are presented in Table 5.1.

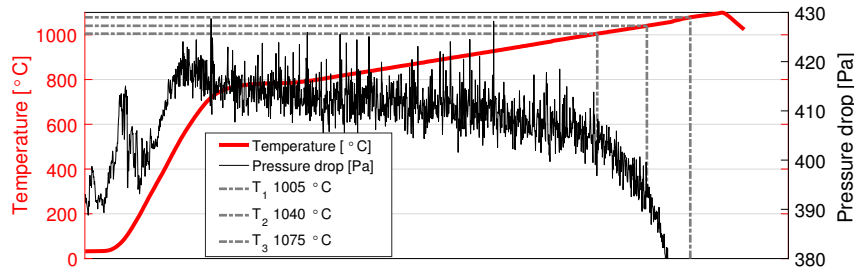


Figure 5.1: Differential pressure over the bed sample LD30 with increasing temperature.

Table 5.1: Experimental result of investigating the sintering and agglomeration behaviour of the bed material under hot conditions.

	LD-slag			Silica sand 100%
	Fresh	30%	100%	
T_1 [°C]	-	1005	960	1030
T_2 [°C]	-	1040	-	1050
T_3 [°C]	-	1075	-	1060

According to Table 5.1, LD-slag shows no tendency towards sintering followed by agglomeration at high temperatures. The test proves that LD-slag has a slightly higher resistance towards sintering compared with fresh silica sand.

5.1.2 Bed material particle size

The particle size distribution was initially compared between the fresh sand sample and its technical specification, Figure 5.2. Further, the fresh sand sample was compared with the bottom ash sample and the result is presented in Figure 5.3.

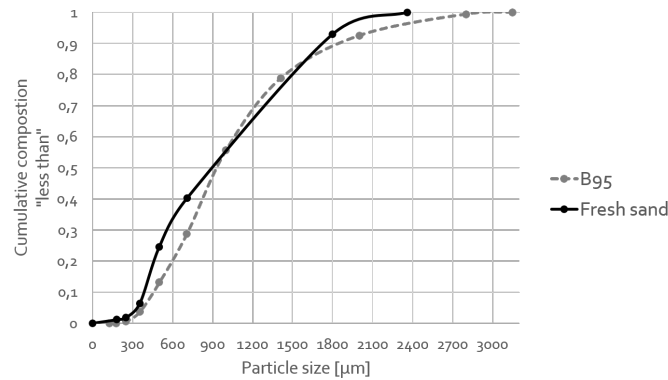


Figure 5.2: Particle size distribution comparison between the fresh sand sample and the corresponding technical specification.

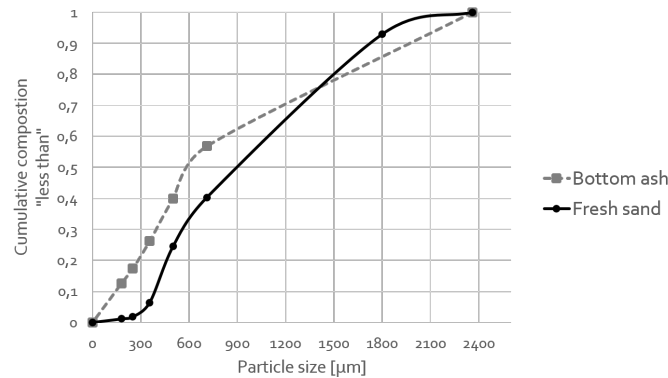


Figure 5.3: Particle size distribution comparison between the sand and bottom ash

As the relationship between the fresh sand sample and bottom ash was established, a comparison between fresh sand and the original product LD-slag 0/2 delivered by Merox was performed. The results are presented in Figure 5.4.

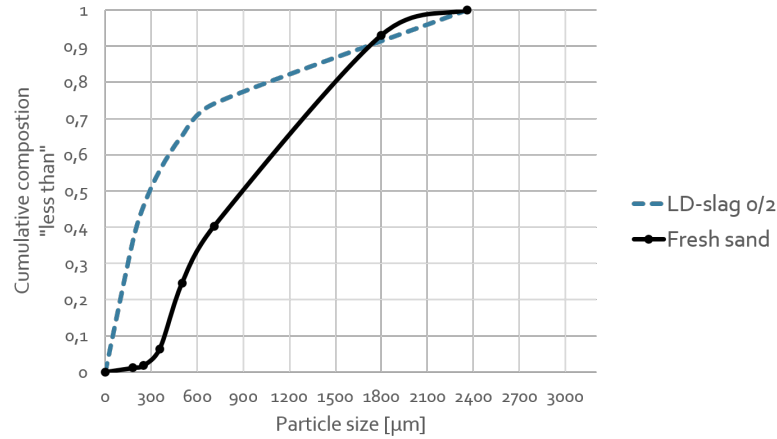


Figure 5.4: Particle size distribution comparison between the fresh sand and original LD-slag 0/2.

Further, fines was removed from the original LD-slag 0/2 to have a size distribution more similar to the fresh sand. The resulting mixture is referred as LD-slag 355 and its size distribution together with fresh sand and LD-slag 0/2 is presented in Figure 5.5.

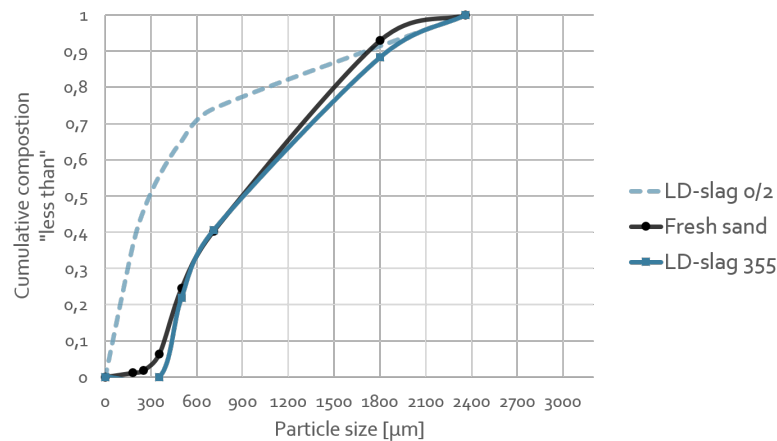


Figure 5.5: Particle size distribution comparison between the fresh sand and LD-slag 355.

5.1.3 Fluidization behaviour

The resulting fluidization behaviour of bottom ash, LD-slag 0/2 and LD-slag 355 is presented in Figure 5.6. The pressure is measured above the bed against the atmosphere.

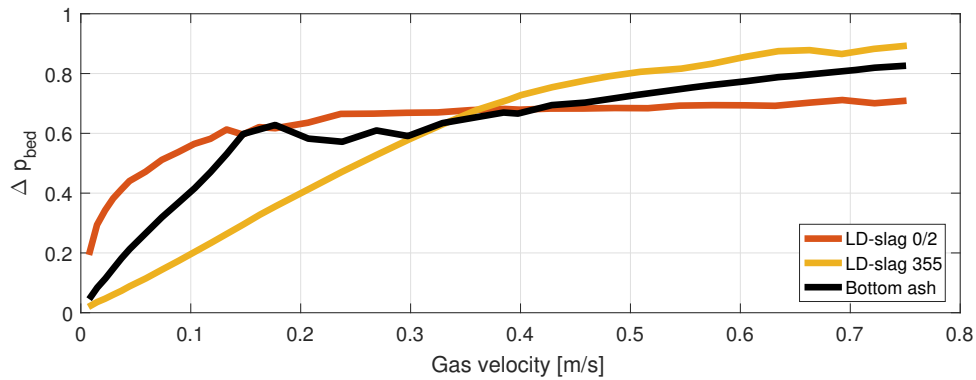


Figure 5.6: Differential pressure over the bed for increasing air flow rates for bottom ash, LD slag 0/2 and LD slag 355.

Figure 5.6 shows that LD-slag 355 requires higher gas velocities to be fluidized. This imply that that the minimum flow rate of primary air must be changed if LD-slag 355 is used as bed material. On the other hand, LD-slag 0/2 has similar velocity of minimum fluidization as the bottom ash and no changes should be required.

5.2 Qualitative analysis of introducing LD-slag

The qualitatively evaluation of introducing LD-slag 0/2 in the bed inventory of Sävenäsverket, based on quantitatively data, is summarized and presented in Table 5.2.

Table 5.2: Summarising of the qualitatively evaluation of introducing LD-slag 0/2 in the industrial process based on some quantitative data.

Technical aspects		Silica sand	LD-slag 0/2
Bed material handling	Feeding to silo	Pressurized delivery trucks, 40 bar.	No changes required.
	Storage in silo	50 tons stored, enough for 5-6 days of operation.	As the density is slightly lower, no risk of violating the weight limit.
	Feeding to furnace	Fed by a screw conveyor.	No changes required.

continues on next page...

Continuation of Table 5.2

Technical aspects		Silica sand	LD-slag 0/2
Bed particles behaviour during operation	Density [kg/m^3]	$\rho_{\text{bulk}}=1.58$	$\rho_{\text{bulk}}=1.51$
	Reaction with ash particles	Have to regenerate bed particles.	Not studied. Regeneration necessary as well.
	Alkali absorption	Successfully absorb alkali to an extent.	Poor alkali absorption properties.
	Sintering properties	Sintering tendencies around 900-950 °C.	No agglomeration tendency below 1000°C.
Ash management	Bottom ash	Non-hazardous. Currently used for soil improvement products.	Non-hazardous. Can still be managed by the same company.
	Fly ash	Nutrient recirculation practised.	Risk of LD-slag in fly ashes (higher amount of fines).
Fluidization properties	Characteristics	Segregated bed.	Segregated bed.
	Velocity of minimum fluidization	0.18	0.15-0.20
	Differential pressure over the bed	6.5 kPa.	Small deviation for same amount of material because of the density difference.
Bed temperature	Temperature variations	Temperature variations over the bed.	More stable bed temperatures expected.
	Cooling of the bed	Cooled by limiting the supply of oxygen, either by flow rate or recirculating flue gases.	More heat development in the bed is expected. But cooling is regarded as sufficient.
Emissions	Emissions of CO	Relatively high emissions of CO occur	Both increasing and decreasing effects of CO are expected. Net effect is unsure.
Process air	Air-to-fuel ratio	$\lambda \simeq 1.35$	Possible operation with lower λ expected.

5.3 The experimental campaign

Based on the qualitative evaluation presented in Table 5.2, a low-risk experimental campaign program was constructed. The strategy is presented after the risks and associated suggested actions to minimize certain risks are presented.

5.3.1 Risk reduction

Some of the uncertainties associated with introducing LD-slag as bed material were possible to minimize and even remove. The ones considered for the campaign and proposed measures are presented in Table 5.3.

Table 5.3: Proposed actions to minimise certain risks investigated

	Risk	Proposed action
R1	Risk of having different regime of fluidization and differential pressure over the bed with LD-slag compared with silica sand.	Introduce a <i>small batch</i> of LD-slag before the campaign to register any drastic changes. Section 5.3.2
R2	Risk of bed agglomeration due to lost alkali absorption (increased amount of alkali for a lower amount of silica sand as 30 wt-% is replaced by LD-slag).	Decrease average time in the bed of the silica sand particles by applying forced regeneration before the campaign.
R3	Risk of not reaching the desired fraction of LD-slag in the bed inventory.	Estimate how bed composition varies with time to chose a correct <i>batch size</i> of LD-slag. Section 5.3.3
R4	Risk of very time consuming experimental campaign.	Speed up the process by <i>forced extraction</i> when suitable. Section 5.3.4
R5	Fluctuations in fuel stream quality.	Keep as homogeneous fuel mixture as possible during the campaign.
R6	LD-slag causes a stop in the bed material feeding screw conveyor.	Forced outage of the remaining LD-slag in the sand silo from the bottom.

5.3.2 Small batch introduction

It is important that the small batch is big enough to affect the system sufficiently but also small enough to reduce the risk of undesired disturbances. 10 ton of LD-slag is considered as representative amount. The estimated time varying composition during the small-batch test is presented in Figure 5.7.

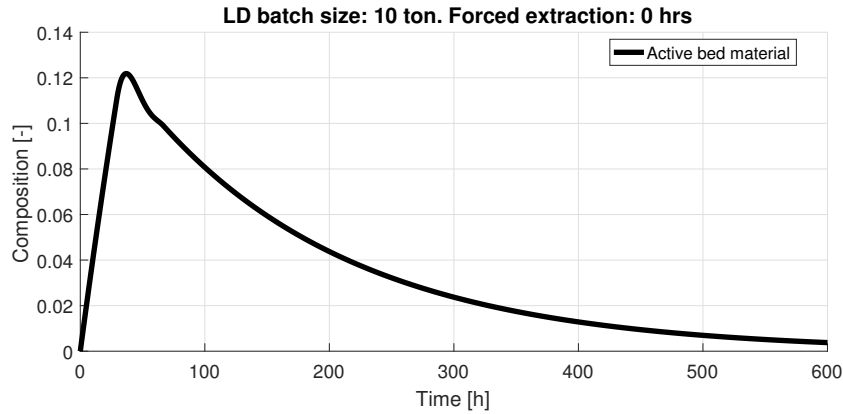


Figure 5.7: Small batch. Variation of LD concentration in the bed inventory with time.

As expected, the concentration of LD-slag increases until the batch of LD-slag in the sand silo is consumed. The maximum fraction of LD-slag in the bed inventory is 12% which is reached after roughly two days.

5.3.3 Batch size

How the share of LD-slag varies with time for different batch sizes is presented in Figure 5.8. The composition in wt-% presented includes the bottom ash. This imply that the target 30/70% mixture of LD-slag and silica sand (excluding the bottom ash) is represented by 20.2 wt-% of LD-slag in the following figures.

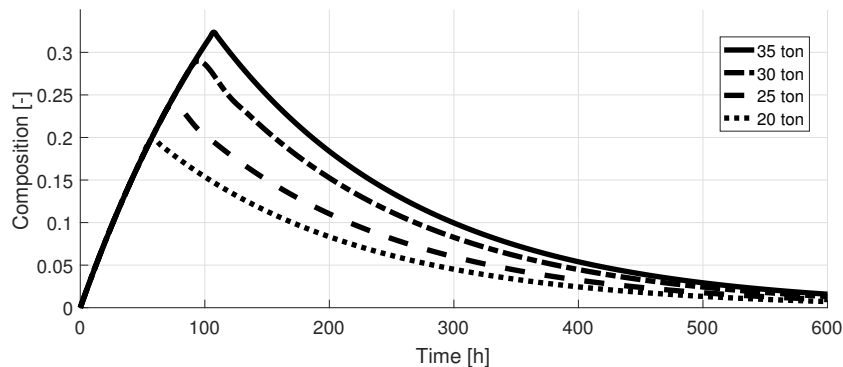


Figure 5.8: Variation in LD-slag concentration with time for different batch sizes.

As the feeding rate is equal for all batch sizes, the increasing fraction of LD-slag the first hours is the same according to Figure 5.8. Highest fraction of LD-slag is reached for the biggest batch size, 35 ton. How the total composition of the bed varies with time is presented in Figure 5.9 where a batch size of 30 ton of LD-slag is investigated.

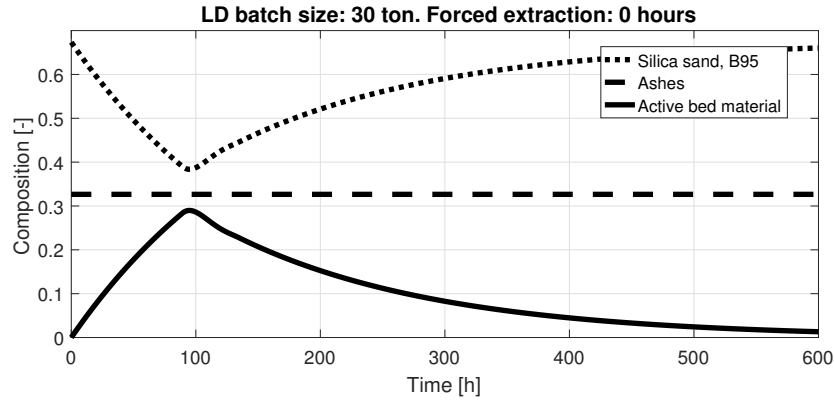


Figure 5.9: Variation in bed inventory composition for 30 ton of LD-slag.

As expected, the fraction of ashes is constant while the fraction of silica sand and LD-slag are decreasing and increasing, respectively.

5.3.4 Forced extraction

For normal extraction and intake rates the campaign is protracted considering the time duration. To speed up the campaign forced extraction when applicable is proposed. For the small batch introduction forced extraction of bed material can be practised right after the maximum concentration have occurred. For the small batch introduction, 10 ton, forced extraction after 48 hours is investigated. The variation in composition of LD-slag in the bed inventory with time as forced extraction is practised, is presented and compared with normal extraction rate in Figure 5.10.

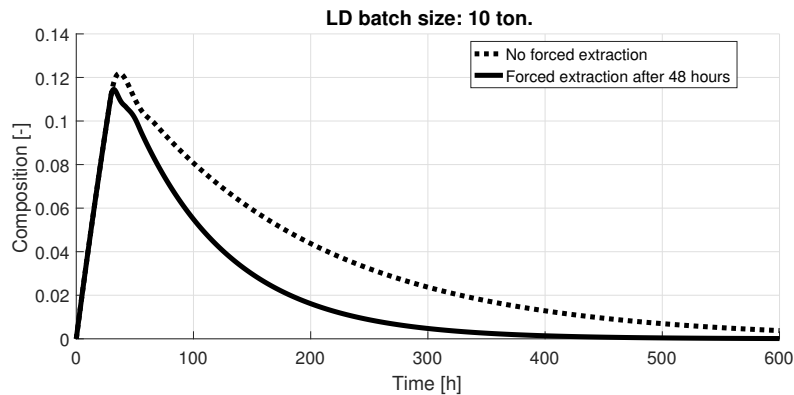


Figure 5.10: Variation of fraction of LD-slag in the bed. Comparison between normal extraction rate and the scenario of forced extraction after 48 hours.

As the big batch of LD-slag is introduced successively, forced extraction can be used to speed up the time to reach desired bed composition and to regenerate the bed in the end. For this campaign, with the batch size of 30 ton LD-slag, it was investigated to apply forced extraction the first 15 hours and then again after 120 hours. The resulting time varying bed composition is presented in Figure 5.11.

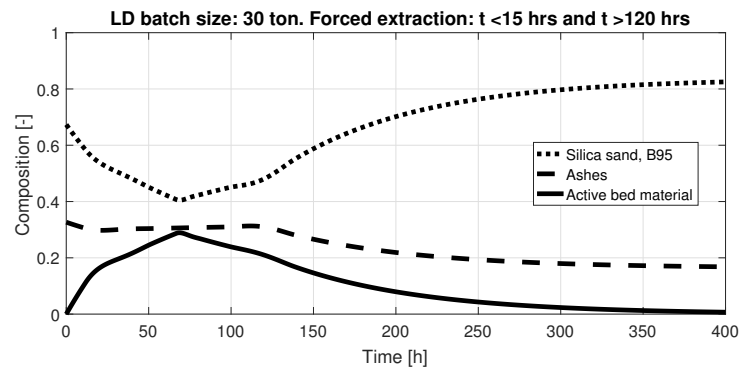


Figure 5.11: Variation in bed inventory composition with time as 30 ton of active bed material is introduced successively while forced extraction is practised the first 15 hours and after 120 hours respectively.

This result in a time period of roughly 3-4 days (85 hours) with sufficiently high share of LD-slag in the bed inventory (referring to the desired bed composition formulated for the campaign).

5.3.5 Campaign realization

The campaign realization consist of four steps, Figure 5.12, based on the measures presented in the risk analysis. The entire campaign, all steps included, is estimated to take two weeks in total. Out of these two weeks, actual operation with sufficiently high share of LD-slag in the bed constitute five to six days.

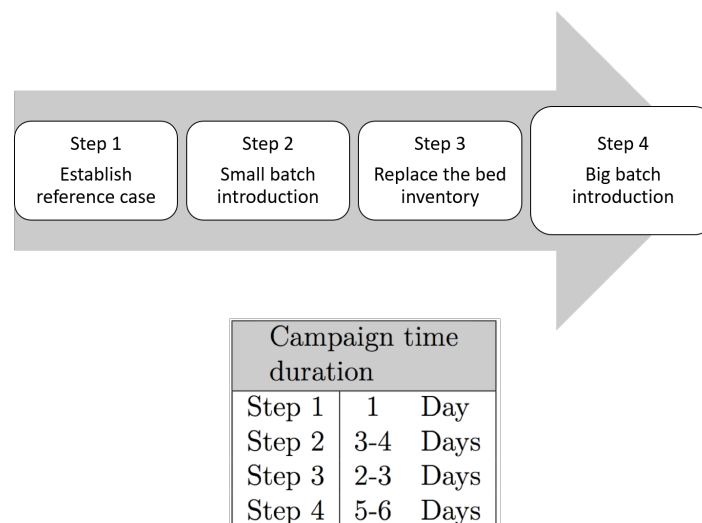


Figure 5.12: Overview of the experimental campaign layout

5.3.5.1 Step 1: Establish reference case

Establish and collect reference data related to silica sand operation running with the reference fuel mixture. This includes bed and flue gas temperatures, feeding rates of sand, fuel and air, extraction rates of bottom and fly ash and level of emissions registered.

5.3.5.2 Step 2: Small batch introduction

Introduce a small batch of LD-slag, 10 tons, in the sand silo which will be successfully introduced in the furnace and the bed inventory as the differential pressure over the bed will keep normal extraction and feeding rates. The objective is to register any unexpected and undesired effects. 10 ton of LD-slag is regarded to be large enough to impact the boiler operation sufficiently but also small enough to be considered as a short duration disturbance. Forced extraction is to be practised after 48 hours. According to the estimated time-varying bed composition, Figure 5.10, the maximum concentration of LD-slag is reached after 32 hours.

5.3.5.3 Step 3: Replace the bed inventory

In accordance with the risk analysis, a complete regeneration of the sand particles in the bed will be practised before introducing the big batch of LD-slag. Hence, step 3 of the experimental campaign is to continue with forced extraction after step 2. This to extract and remove all the remaining content of LD-slag in the bed inventory from the small batch introduction but also to decrease the average age of the silica sand particles in the bed.

5.3.5.4 Step 4: Big batch introduction

30 tons of LD-slag will be loaded in the sand silo. By practising forced extraction the first 15 hours and after 120 hours with normal extraction rate in between, the desired bed composition is kept for roughly 85 hours, Figure 5.11. Further, the maximum concentration is reached after 70 hours before it starts to decrease again as the batch is consumed and traditional silica sand is fed instead.

5.3.5.5 Campaign evaluation and data collection

Data to collect during the campaign is both fly- and bottom ash samples but also the system response on lowering the air-to-fuel ratio. The size distribution and density of the ash samples should be determined. Selected ash samples should be sent for composition analysis to derive the share of LD-slag. For estimation purposes during the campaign, a magnetic separator could potentially be used to estimate the content of iron in the bottom ashes and hence the fraction of LD-slag in the bed. But the magnetic separability of LD-slag from silica sand must be established beforehand.

During the first day of the campaign the reference scenario for silica sand operation should be established. This includes the overall performance of the boiler in addition to the system response on lowering the air-to-fuel ratio. At least one sample of fly- and bottom ash should be extracted as well to be compared with.

During the following 5-7 days (step 2 and 3), small share of LD-slag will be present in the bed inventory. One bottom ash sample should be extracted in the end of step 3 to establish the starting bed composition before initializing step 4.

At least three bottom ash samples are desirable to collect each day during big batch introduction, step 4. This can be accomplished by collecting samples with a eight hour interval, e.g. at 05.00, 13.00 and 21.00, respectively. Sieve analysis and density estimations can be practised in meantime to observe any trends/variations by comparing with reference ash sample. To collect data regarding the system response on lowering the air-to-fuel ratio, it is suggested to be repeated at least three times between 35 and 120 hours of operation after initializing step 4, when the bed composition should be similar to the desired one.

Chapter 6

Discussion

In this chapter the results presented in previous chapter, Chapter 5, are discussed. The investigated material properties of LD-slag are discussed together with a general discussion regarding the presented qualitatively evaluation. Then the proposed experimental campaign is discussed.

6.1 Material properties

The material properties investigated for LD-slag during this project considered sintering behaviour, desired particle size of the product and the resulting fluidization behaviour.

6.1.1 Sintering behaviour

It is observed that LD-slag successfully resist high temperatures without showing any tendencies towards agglomeration, Table 5.1. In comparison with fresh silica sand, LD-slag show similar if not even slightly better resistance of agglomeration at high temperatures. The sintering temperatures for LD-slag together with already experienced operation with LD-slag in the Chalmers boiler, clearly indicate that the high temperature operation will not impede the use of LD-slag as bed material.

6.1.2 Particle size

For the fresh sample of sand, the particle size distribution agrees with the technical specification, Figure 5.2. However, when compared with bottom ash it is observed that the bottom ash contain larger share of fines. This despite roughly 75 wt-% of the bottom ash consist of sand. The higher content of fines in the bottom ash either originates from the fuel ash or from a non-representative sample. Bottom ash samples are extracted by opening a pocket in the bottom ash conveyor. If large particles is stuck in the pocket, only small particles (i.e. fines) are possible to pass which will result in a non-representative sample with higher content of fines. Hence, this must be considered before establishing the relation between the particle size distribution of fresh sand and bottom ash.

The product LD-slag 0/2 delivered by Merox consists of higher share of fines compared with the fresh sand, Figure 5.4. Hence, if LD-slag 0/2 is introduced the increased amount of fines will most likely result in undesired particle entrainment. The entrained particles are then collected either as bottom ash if collected in bottom of the convection section or as fly ash collected by the flue gas cleaning. This implies that LD-slag 0/2 is not preferable for long term operation. However, it is regarded as a suitable product for an short term experimental campaign as the product already exist and no additional process cost will be needed.

6.1.3 Fluidization behaviour

The resulting differential pressure over the bed as the gas velocity is increased for LD-slag and bottom ash agrees with the shape of a segregated bed, according to theory. A segregated bed was also observed during the experiment as the fines started to fluidize and positioned over the larger particles which did not yet fluidize. Similar fluidization behaviour for LD-slag 0/2 as for bottom ash was observed, Figure 5.6. This is motivated by the similar amount of fines. In contrast, higher velocities of minimum fluidization was observed for LD-slag 355 because of the lack of fines. If it is assumed that fines originates from the fuel, similar fluidization behaviour as the current bed inventory is expected for LD-slag 355. If not, LD-slag 0/2 is appropriate to achieve similar fluidization behaviour.

6.2 Qualitative evaluation

As presented in Table 5.2, Chapter 5, several subsystems and parameters in Sävenäsverket are related and important to consider when introducing LD-slag. How the different systems are affected vary, and some of them are most likely not affected at all. The discussion behind the construction of the qualitatively evaluation presented in Table 5.2 is motivated and clarified below.

Considered the bed material handling system, no modifications are necessary. The particle distribution between LD-slag and sand similar, expect a higher share of fines, in addition to the particle characteristics which imply that the handling system will be able to operate as usual. However, this holds for an experimental campaign. If long term operation is to be considered, a solution to supply a mix of silica sand and LD-slag must be designed since only one sand silo is installed.

Slightly less density was measured for LD-slag 0/2 compared with silica sand. Considering the regeneration of bed material, LD-slag does not absorb alkali in contrast to silica. Hence, regeneration of the bed will be of even higher importance when including LD-slag. If the rotating sieve would be in operation, decreased consumption of LD-slag is possible.

LD-slag is classified as non-hazardous. The resulting bottom ash during the campaign is possible to be managed by the same company. No guarantee given that the utilization practised for silica sand is possible for LD-slag. Especially regarding

the nutrient recirculation currently practised for silica sand. As LD-slag 0/2 is used during the campaign, the amount of fly ash is expected to increase because of the fines. This can be observed by studying an increase in signal frequency of the fly ash sender.

The change in fluidization behaviour have been discussed already. However, as a consequence of a lower density, the reference differential pressure of 65 mbar set for silica sand corresponds to a higher fixed bed height for LD-slag 0/2. As only 30% will consist of LD-slag during the campaign, no significant change is expected.

The fluidized bed is currently suffering of a variation in bed temperature across the cross-section, Figure 4.9, Chapter 4. At the fuel inlet on the front wall the bed temperature is more than 50°C lower than the opposite wall. By the hypothesis that LD-slag will be reduced in the fuel-rich region, supplying oxygen for combustion, this temperature difference will be balanced out to a certain degree. Additionally, as the combustion is moved further down in the bed as a result of OCAC, higher heat development in the bed is expected. To keep the desired temperature, cooling is practised by recirculating flue gases instead of air. The flue gas recirculation fan is currently running at roughly 40% of its capacity. This motivates that the bed cooling system is sufficient to compensate for an increased heat development in the bed when needed to maintain the bed temperature at 850°C.

Again, according to the expected result of OCAC operation, the excess air supplied should decrease. Further, according to Figure 5.6, LD-slag 355 requires higher primary air flow rates to be fluidized. If used, the minimum limit of primary air flow rate must be increased to ensure that defluidization is avoided.

6.3 The experimental campaign

A stable and successful operation during the campaign is required to conduct a sufficient evaluation. As the operational conditions change, it is challenging to pin down the reason behind it. Certainly as it is near impossible keeping the boiler environment constant over a long period of time. The biggest contribution behind this challenge is the fluctuating fuel quality. Hence, it is highly desired to maintain a certain fuel mix constant for as long as possible.

For this reason, trends appearing in the boiler are more importantly to consider rather than confirming a specific relation between two parameters. The objectives for the experimental campaign was formulated accordingly, Chapter 5.3.5, and the construction of the campaign as well, Chapter 3.3. Systematic reduction of the air-to-fuel ratio, collection ash samples and observation of temperature trends are included in the campaign design and bring important information for the evaluating each of the different objectives.

The risks of having different fluidization behaviour impact the overall regulation of the boiler. If higher gas velocities are required to keep the bed fluidized, the lower

limit of primary air flow rate must be increased. If lower gas velocities are required instead, the upper limit must be decreased to avoid particle entrainment. LD-slag 0/2 had the same behaviour as the bottom ash in contrast to LD-slag 355 which required higher gas velocities.

If the silo is loaded with a too large batch of LD-slag than required, the share of LD-slag and time duration with LD-slag operation will extend the desired design values. This brings increased risks as well, e.g. bed agglomeration.

An additional action, beside those presented already, to decrease the time duration of the campaign is to forcing the bed pressure to drop. This decrease the time duration as the share of LD-slag is initially allowed to increase rapidly. The bed pressure is recommended not to go below 5.5 kPa. The proposal is therefore to decrease the pressure from 6.5 to 5.5 kPa through extraction of bottom ash only (corresponds to 12.3 tonnes). Then, LD-slag is fed to increase the bed pressure back to 6.5 kPa again, reaching a high share of LD-slag fast.

Chapter 7

Conclusions

By gathering information related to the process of Sävenäsverket it can be concluded that:

- An experimental campaign of introducing oxygen-carrying bed material in Sävenäsverket is possible to realize. No restrictions in the process of Sävenäsverket are identified. No additional equipment is required to be installed, nor any modifications of the existing equipment.
- The campaign is constructed as a batch-wise introduction of LD-slag in the bed inventory. Batch size of 30 tons is suggested. This imply that the share of active bed material will successively increase until the batch is empty and the share starts to decrease again as silica sand is introduced. The desired bed composition is determined to 30/70% LD-slag and silica sand, respectively. This concentration is withhold for at least three days in the proposed campaign.
- By evaluating the experimental campaign, it should be possible to evaluate if OCAC operation in Sävenäsverket is a promising solution to excising fluctuations and variations present in the boiler.

By studying material properties of LD-slag and by comparison with silica sand it is found that:

- LD-slag resist high temperatures. For 100% LD-slag sample maximum temperature of the lab reactor, 1100°C was reached without complete bed agglomeration. For silica sand B20 complete agglomeration was reached at 1060°C. LD-slag have been operated without showing any tendencies of attrition or bed agglomeration.
- LD-slag doesn't absorb much alkali in contrast to silica sand. The loss of alkali absorption is countered by operating with 30/70% bed mixture of LD-slag and silica sand.
- A suitable LD-slag product exists already that can be used for the experimental campaign, no pretreatment is necessary. The size distribution is somewhat similar to the bottom ash, while including more fines when compared to the fresh sand.

- The LD-slag product considered, LD-slag 0/2, have similar fluidization behaviour as the bottom ash.
- The characteristics of both sand- and LD-slag based beds are segregated when fluidized because of the wide particle size distribution of the materials.
- LD-slag is classified as non-hazardous and the current company handling Sävenäsverket's bottom ash approves an involvement of LD-slag in the ashes.

Bibliography

- [1] M. Junginger J. Koornneef and A. Faaij. Development of fluidized bed combustion — An overview of trends, performance and cost. *Progress in Energy and Combustion Science*, 33(1):19 – 55, 2007.
- [2] J. R. Grace T. Fouilland and N. Ellis. Recent advances in fluidized bed technology in biomass processes. *Biofuels*, 1(3):409–433, 2010.
- [3] H. Thunman, F. Lind, C. Breitholtz, N. Berguerand, and M. Seemann. Using an oxygen-carrier as bed material for combustion of biomass in a 12-MW_{th} circulating fluidized-bed boiler. *Fuel*, 113:300 – 309, 2013.
- [4] F.Lind, A. Corcoran, and H. Thunman. Validation of the oxygen buffering ability of bed materials used for OCAC in a large scale CFB boiler. *Powder Technology*, 316:462 – 468, 2017. Fluidization for Emerging Green Technologies.
- [5] European Environment Agency. Renewable energy in Europe - Recent growth and knock-on effects, 2017.
- [6] M. Balat and G. Ayar. Biomass energy in the world - Use of biomass and potential trends. *Energy Sources*, 27(10):931–940, 2005.
- [7] A.A. Khan, W. de Jong, P.J. Jansens, and H. Spliethoff. Biomass combustion in fluidized bed boilers: Potential problems and remedies. *Fuel Processing Technology*, 90(1):21 – 50, 2009.
- [8] A. Demirbas. Combustion characteristics of different biomass fuels. *Progress in Energy and Combustion Science*, 30(2):219 – 230, 2004.
- [9] J. D. Laumb, B. C. Folkedahl, and C. J. Zygarlicke. Chapter 4 - Characteristics and behaviour of inorganic constituents. In B. G. Miller and D. A. Tillman, editors, *Combustion Engineering Issues for Solid Fuel Systems*, pages 133 – 170. Academic Press, Burlington, 2008.
- [10] E. Brus, M. Öhman, and A. Nordin. Mechanisms of bed agglomeration during fluidized-bed combustion of biomass fuels. *Energy & Fuels*, 19(3):825–832, 2005.
- [11] J. Werther, M. Saenger, E.-U. Hartge, T. Ogada, and Z. Siagi. Combustion of agricultural residues. *Progress in Energy and Combustion Science*, 26(1):1 – 27, 2000.
- [12] D. Kunii and O. Levenspiel. Chapter 3 - fluidization and mapping of regimes. In D. Kunii and Octave O. Levenspiel, editors, *Fluidization Engineering (Second Edition)*, pages 61 – 94. Butterworth-Heinemann, Boston, second edition edition, 1991.

- [13] T. Mattisson, A. Lyngfelt, and H. Leion. Chemical-looping with oxygen uncoupling for combustion of solid fuels. *International Journal of Greenhouse Gas Control*, 3(1):11 – 19, 2009.
- [14] E. Jerndal, T. Mattisson, and A. Lyngfelt. Thermal analysis of chemical-looping combustion. *Chemical Engineering Research and Design*, 84(9):795 – 806, 2006. Carbon Capture and Storage.
- [15] J. Adanez, A. Abad, F. Garcia-Labiano, P. Gayan, and L. F. de Diego. Progress in chemical-looping combustion and reforming technologies. *Progress in Energy and Combustion Science*, 38(2):215 – 282, 2012.
- [16] A. Lyngfelt. Chemical-looping combustion of solid fuels – Status of development. *Applied Energy*, 113:1869 – 1873, 2014.
- [17] T. Mattisson, A. Lyngfelt, and P. Cho. The use of iron oxide as an oxygen carrier in chemical-looping combustion of methane with inherent separation of CO₂. *Fuel*, 80(13):1953 – 1962, 2001.
- [18] A. Abad, T. Mattisson, A. Lyngfelt, and M. Rydén. Chemical-looping combustion in a 300 W continuously operating reactor system using a manganese-based oxygen carrier. *Fuel*, 85(9):1174 – 1185, 2006.
- [19] D. Jing, T. Mattisson, M. Ryden, P. Hallberg, A. Hedayati, J. Van Noyen, F. Snijkers, and A. Lyngfelt. Innovative oxygen carrier materials for chemical-looping combustion. *Energy Procedia*, 37:645 – 653, 2013. GHGT-11 Proceedings of the 11th International Conference on Greenhouse Gas Control Technologies, 18-22 November 2012, Kyoto, Japan.
- [20] A. Lyngfelt, B. Leckner, and T. Mattisson. A fluidized-bed combustion process with inherent CO₂ separation; application of chemical-looping combustion. *Chemical Engineering Science*, 56(10):3101 – 3113, 2001.
- [21] F. Lind M. Rydén, M. Hanning. Oxygen carrier aided combustion of wood chips in a 10 MW circulating fluidized bed boiler using steel converter slag as bed material. *Manuscript in preparation*, to be submitted 2018.
- [22] M. Rydén A. Lyngfelt P. Moldenhauer, C. Linderholm. Experimental investigation of chemical-looping combustion and chemical-looping gasification of biomass-based fuels using steel converter slag as oxygen carrier. *1st International Conference on Negative CO₂ emissions*, Göteborg, Sweden, May 2018.
- [23] Göteborg Energi. Miljörapport 2016 Sävenäsverket, 2017.