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Evaluation of bed-to-tube heat transfer in a fluidized bed heat exchanger in a 75 MW_{th} CFB boiler for municipal solid waste fuels

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ABSTRACT

Bed-to-tube heat transfer has been investigated for a tertiary superheater in a 75 MW_{th} Circulating Fluidized Bed (CFB) boiler in Norrköping, Sweden. The boiler is used for incineration of solid waste fuels. Two fluidized bed heat exchangers are located in loop seals, connecting the cyclones and the furnace. The heat exchangers are placed in series, with respect to the steam side, and in parallel, with respect to the particle side. The total heat transfer surface area is roughly 44 m^2, distributed over 72 tubes. The total effect transferred most often is in the range 2–6 MW. The incoming steam temperature in the first superheater is 380–400 °C, while the exiting steam temperature from the second is around 450°, at 65 bar pressure. The bed temperature in the Fluidized Bed Heat Exchanger (FBHE) is 850–875 °C. The analysis is based on operational data from two time periods (2002–2005 and 2014–2021). The two periods use different heat exchanger designs, following a retrofit in 2005. The aim of the study is to establish the bed-to-tube heat transfer coefficient in an industrial FBHE unit and investigate how it varies over different time periods, for two different bed materials and for two different designs. Also, the experimentally determined heat transfer coefficients are compared with an established heat-transfer correlation, for prediction of heat transfer from bubbling fluidized bed to tubes. Operation with two bed materials were evaluated, namely silica sand and crushed and beneficiated ilmenite. Both materials are classified as Geldart B particles. Air is used as fluidization gas in the FBHE. The analysis show, with a few exceptions, comparably low heat-transfer coefficients from bed to tube of 100–150 W/(m^2K). The results were similar for silica sand and ilmenite, but the highest measured heat transfer coefficient was for a period with ilmenite. The heat transfer was lower than expected based on literature data from FBHE units and fluidized bed boilers in general, and much lower than bed-to-tube heat transfer coefficients from lab-scale experiments and empirically derived predictive expressions. The difference could be related to one or more of several factors, such as the effect of very small tube spacing, unknown thermal conductivity of one of the layers in the tube bundle, the effect of lateral particle flow and the effect of fouling due to ash layers forming on the tube surfaces. It is suggested that it should be possible to significantly increase the bed-to-tube heat transfer by increasing the tube pitch, which is expected to improve bed mixing without increasing the risk of corrosion.

1. Introduction

A key characteristic of fluidized beds is their ability to provide high heat-transfer rates within the bed, resulting in almost uniform temperature [1]. The heat transfer in the dense zones of Bubbling Fluidized Beds (BFBs) is good, both inside the bed and from the bed to immersed objects. Fluidized Bed Heat Exchangers (FBHEs) is an example of a technology which makes use of this phenomena. In a FBHE, tubes with fluid flowing inside them are heated or cooled by being immersed in a bubbling fluidized bed. In Circulating Fluidized Bed (CFB) boilers, such devices are often placed subsequent of the cyclone. Hot bed particles from the combustion chamber are separated from the flue gases in the cyclone, from which they exit via a return leg. The hot bed particles are then cooled in a FBHE, before being returned to the furnace. The bed particles are cooled by submerged tubes, in which water/steam is evaporated/heated to generate saturated/superheated steam. FBHE units can be used to reduce the load on heat exchange surfaces in the furnace, or in the convection path where heat transfer coefficients are lower (around or below 100 $Wm^{-2}K^{-1}$, which shall be compared with FBHE units where it should be possible to reach at least 500 $Wm^{-2}K^{-1}$).

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Compared to the convection path, superheaters in the FBHE are also less exposed to chlorides, alkali and steam, which should reduce the risk of corrosion. The FBHE units can also be used to control the temperature of the superheated steam, and the temperature in the furnace.

Particle convection is responsible for most of the heat transfer in this kind of units. In CFB applications, fine sand like particles is used (generally in the size range 100–350 μm). The particles are transported through the furnace at high velocity (several meters per second), whereas the particles in the FBHE generally experience superficial gas velocities below 0.5 m/s.

Most of the research on fluidized bed heat exchangers has been focusing on use of lab-scale fluidized bed units, but there are examples of experimental studies in pilot scale/commercial scale CFB boilers [2–4]. However, in reference [2] and [3] there is no information on the bed material or about the particle size, which are very important parameters for the bed-to-tube heat transfer. Reference [4] is for a semi-commercial scale CFB boiler using a very small heat exchanger with an effect of approximately 10–25 kW, which should be compared to industrial-scale boilers which extracts several MWs. Based on recently published studies performed in laboratory setting, very high heat transfer coefficients (>800 W/(m^{2}K)) have been found at relevant temperatures in lab-scale systems with single horizontal tubes [5,6], and in studies at semi-commercial scale (>500 W/m^{2}K) [4]. It is therefore of interest to perform additional research on industrial boilers with large-scale fluidized bed heat exchangers, to evaluate if similar heat transfer coefficients are achievable for such units.

An experimental investigation in a large-scale CFB boiler adds several challenges, in comparison to tests in lab-scale. Firstly, in industrial scale the FBHE consist of complete tube bundles with several meters of immersed tube surface, often making several tube passes through the bed. In comparison, lab-scale tests typically utilize a single horizontal tube no longer than one meter. Secondly, most industrial scale CFB boilers use fuels that contains ash-forming elements, which can cause agglomeration of bed particles and fouling or corrosion of heat-transferring surfaces. The effect on heat transfer of ash elements could be severe and must be taken into consideration. Important factors which could influence heat transfer include fuel composition, combustion efficiency, temperature levels in the boiler, age of the heat exchange tubes, and the time passed since the last maintenance. Thirdly, these large-scale systems have lateral flow of particles through the fluidized bed heat exchangers, while most predictive heat transfer correlations for FBHEs do not consider the effect of particle flow. The study performed by Stenberg et al. [4] show that the circulating particles has an effect (a few percent when the particle flow was increased), and also that the orientation of the tubes in relation to the particle flow has some effect (6–10 %). Another aspect which can be seen both as a challenge, but also an interesting point of investigation, is the fact that an industrial FBHE unit is in operation over long time periods. This offers an opportunity to study how their performance with respect to heat transfer varies over time.

The challenges involved make it more difficult to evaluate the results of studies at industrial scale, as compared to lab scale. However, it is still important to perform research also on this scale, to bridge the gap between more fundamental research experiments and science which can be applied in industrial settings. This includes also how research can be used to design new improved industrial FBHE units.

Most studies on fluidized bed heat transfer have been performed using silica sand as bed material. This is also the bed material most commonly used in industrial-scale biomass and waste incinerators. However, research performed in Sweden during the last ten years that involves > 10 commercial boilers has shown that the use of ilmenite (a titanium-iron-ore) as bed material can provide several benefits for CFB operation compared to silica sand, mainly due to its oxygen carrying properties and good ability to absorb key ash components such as potassium. The use of ilmenite as bed materials have been shown to reduce the risk of problems related to air and fuel mixing, uneven fuel feeding, CO emissions, sintering and agglomeration. It also provides a possibility to increase fuel load and reduce air-to-fuel ratio, for a given boiler. The results and conclusions of this research effort has recently been reviewed and summarized by Störner et al. [7]. Based on the possibility that other bed materials than silica sand may be applied in future boilers, it is also of interest to evaluate if this influences the bed-to-tube heat transfer coefficient. Previous work performed by Stenberg et al. [4,5] show that the use of ilmenite results in higher bed-to-tube heat transfer, possibly due to its higher density which allows it to carry more sensible heat. It is therefore of interest to evaluate if a similar trend is seen also for an industrial FBHE unit.

The primary aim of this study is to evaluate the bed-to-tube surface heat transfer coefficient to FBHEs in an industrial-scale CFB boiler operating with municipal solid waste and industrial waste. In the plant, two FBHE units operating as tertiary superheaters are located in the loop seal of the boiler, generating steam at 450 °C, 65 bar. A secondary aim is to compare the results with heat transfer correlations which are commonly used to predict heat transfer from a bubbling fluidized bed to a single tube, to examine if such expressions are applicable on industrial FBHEs. Another secondary aim is to evaluate which factors that can have a significant influence on the design of fluidized bed heat exchangers and provide ideas for improved FBHE design. Historical data since the start of the boiler in 2002 is used to perform the analysis. During this time period, both silica sand and ilmenite has been used as bed material.

2. Experimental setup

2.1. P14 – Händelöverket, Norrköping, Sweden

P14 is a CFB boiler for incineration of municipal solid waste and industrial waste. It was commissioned in 2002, as one of the first CFB boilers intended to utilize 100 % waste fuels [8]. Normally, it operates with a fuel load of approximately 75 MW and provides steam at 450 °C, 65 bar. The unit is located at a larger site with a total output of around 500 MW. The plant operates continuously all year, apart from two periods per year when it is shut down for maintenance work, once during spring (April) and once during autumn (October). The plant is presented in Fig. 1, where the key components are shown including the FBHEs which act as tertiary superheaters in this plant.

The boiler consists of a water-cooled furnace, two water-cooled cyclones and a loop seal, which also acts as a FBHE and tertiary superheater. The flue gases from the furnace passes through an empty pass and a convection path which includes superheaters and economizers. The empty path is meant to reduce the flue gas temperature slightly, to make the ash less sticky before the convection path. The flue gas cleaning includes textile filters which remove particles, lime which binds chlorides and sulfur, and active carbon to capture dioxins and heavy metals.

The fuel is top fed to the furnace. Ash (and used bed material) is removed from the furnace using primary air nozzles targeting extraction ports at the bottom of the furnace to two screws, which transport the material to an ash classifier. The classifier acts as a wind sieve separating fine particles at the top which are fed back to the furnace, while the coarse material is transported with transportation screws to ash containers. During operation with ilmenite, a magnetic separation system (Improbed Loop™) is in operation which separates ilmenite from the coarse material by means of its magnetic properties. Thus, the ilmenite can be returned to the boiler with a high efficiency. This system is used to reduce the consumption of fresh ilmenite.

2.2. Fluidized bed heat exchangers

The FBHEs at P14 are divided into two separate parts (left and right, seen from the fuel feeding point, which is on the opposite side). Both FBHEs have identical design, with its own windbox. They were redesigned and altered in 2005. This was due to strong deposit formation and

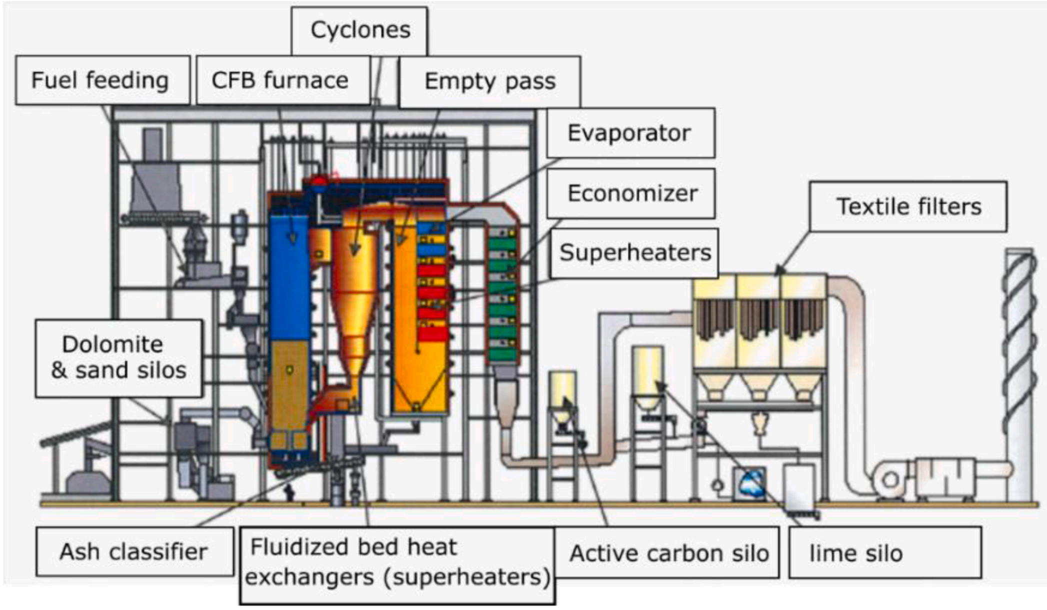

Fig. 1. Schematic view of the P14 CFB boiler in Norrköping.

rapid corrosion of the loop-seal superheater, limiting its lifetime [8]. The redesign will be further explained below.

The FBHEs are placed in parallel on the particle side and in series on the steam side. This means that all steam first passes through the right FBHE, and then through the left FBHE. This means that the steam temperature always will be at its lowest in the inlet of the right FBHE, and at its highest in the outlet from the left FBHE. In each of the FBHEs, the tubes are aligned horizontally and pass through the bed four times. The steam is evenly distributed in the 72 tubes, each of which has identical length. The tube pattern is rotated square. The bent parts of the tubes are not exposed to the fluidized bed in the current design. However, it was immersed in the fluidized bed in the original design. The difference has been considered in the heat transfer calculations.

In the original design of the tube bundle, used until 2005, the ratio between the outer and inner diameter was approximately 1.6, while in the new design that number is 2.14. The difference is explained by the current use of double-layer tubes. Essentially, a layer with poor thermal conductivity was added onto the original superheater tubes. The purpose of this arrangement is to achieve a surface temperature at the outside of the outer tube of around 700 °C. During waste incineration, the ash which causes corrosion consists of a mix of chemical compounds which makes it difficult to define a certain melting point. Rather, there is a melting range at which the ash deposits melt [9]. It is important to avoid the surface temperature range which is defined as sticky, which means that 15–70 % of the ash is melt. This is because in this range there is a significant risk of corrosion and increased risk for agglomerations. The decision to redesign the FBHE and use a double-layer tube, was to raise the temperature at the tube surface to safely maintain a temperature above this sticky temperature range. Also, it should be mentioned that the protective layer serves as a shield for erosion. Fig. 2 displays a schematic overview of the temperature profile for the present tube bundle.

The use of the protective tube layer does come at a cost in terms of reduced heat transfer. Basically, the resistance for heat transfer increases significantly, since heat now has to pass also through both the protective tube, before reaching the superheater tube. Another difference in tube bundle design that is predicted to reduce the heat transfer is the fact that the spacing between the tubes in the redesigned FBHE is smaller, which has a negative effect on the mixing of the bed material [10]. The tube pitch in the original design was 3.18D_{to} and in the new design it is 2.35 D_{to}. Consequently, the overall heat transfer coefficient (U) is decreased.

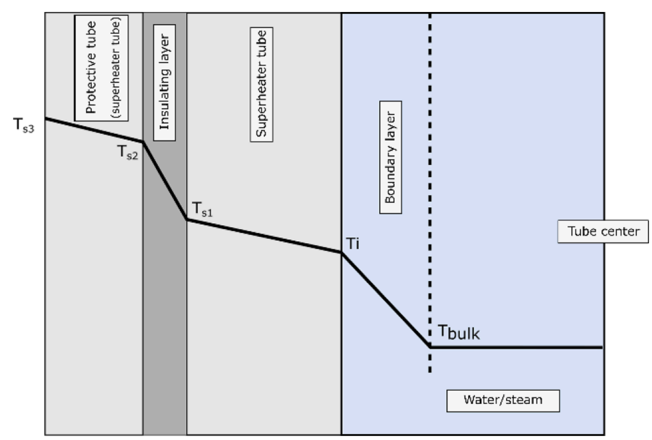

Fig. 2. Simplified temperature profile in the present fluidized bed heat exchanger.

Since 2016 the volumetric flow in the FBHE's has been measured which makes it possible to calculate the superficial gas velocity to approximately 1 m/s in both units. The actual gas velocity between the tubes in the original tube bundle design was approximately 1.46 m/s. The corresponding number for the retrofitted heat exchangers is about and 1.74 m/s. The whole fluidized bed heat exchanger is considered to be immersed in the bed. The bed temperature varies to some extent with the fuel load and the temperature difference between the heat exchangers also varies slightly, but the difference is not>20 °C.

2.3. Bed material

The silica sand which is used at P14 is Baskarp B20, provided by Sibelco Nordic AB. Ilmenite is a mineral ore rich with titanium and iron oxides that is mined mainly for production of titanium dioxide. The material used was a concentrate of Norwegian rock ilmenite provided by Titania A/S. It is a product which have been physically beneficiated to remove most slag elements. Fresh particles of both materials have an average particle size of about 200 µm. The main physical difference between them, which could potentially affect heat transfer, is the significant difference in bulk density (about 1500 kg/m^3 for silica sand and about 2300 kg/m^3 for ilmenite). The key physical and chemical properties of fresh particles are presented in Table 1.

Further, operation of a smaller biomass fired CFB plant with silica sand and ilmenite suggests that both silica sand and ilmenite changes significantly during operation, with individual particles growing larger and becoming more porous [4]. A sample of actual bed material was collected from material which had passed backwards through the nozzles at the bottom of the fluidized bed heat exchangers. The bed sample was examined and sieved. The results show that the mean particle size is just below the mean particle size of the fresh bed material. 4 % of the bed sample was smaller than 90 μm and 2 % was>500 μm. These quantities were not included in the estimation of the average particle size. At the time of writing, it was not possible to retrieve relevant bed samples with ilmenite in operation at P14.

The precise effect of momentarily differences in particle properties cannot easily be taken into consideration in the current study, which relies on historical data. Further assumptions about the properties of the bed material are listed and commented on below.

3. Methodology

3.1. Data collection of experimental results

The estimations of the experimental bed-to-tube heat transfer coefficient in the FBHE were made based on the following assumptions:

- Data was collected only for periods of operation when the fuel feed was at least 65 MW_{th}. This was to ensure that all data referred to periods with substantial solids flow through the FBHE units.
- In the calculations of operational hours, the average values on an hourly basis were used (one data point was retrieved per second).
- The steam flow is assumed to be evenly distributed over all the tube loops.
- The heat transfer in the tubes is calculated by using a correlation describing a gas flowing through a circular tube. The heat transfer coefficient on the inside is calculated as an average from the following two equations:

$$h_i = \frac{k_{steam}}{d_i} * j_h Re Pr^{0.33} \tag{1}$$

$$h_i = \frac{k_{steam}}{d_i} * 0.023 Re^{0.8} Pr^{0.4} \tag{2}$$

where k_{steam} = thermal conductivity of the steam, j_h = heat transfer factor on the tube side, Re = Reynold's number = ρud/μ and Pr = Prandtl's number = c_pμ/k. The overall heat transfer coefficient is calculated using Eq. (3) and the bed-to-tube heat transfer coefficient is calculated using Eq. (4). The heat load delivered to the heat exchanger is calculated by multiplication of the heat capacity for the steam, the steam flow and the temperature difference over the fluidized bed heat exchanger.

$$U = \frac{Q}{A\Delta T_{average}} \tag{3}$$

$$h_o = \frac{1}{\frac{1}{U} - \frac{d_o \ln(d_o/d_i)}{2k_w} - \frac{d_o}{d_i h_i}} \tag{4}$$

- The fluid properties for the inside of the tubes are set to steam at 425 °C, 65 bar.
- The bed temperature is assumed to be the value which is measured in the right and left fluidized bed heat exchanger respectively.
- The heat transfer area is similar for the original and the present design (44.4 m^2 in the original design and 44.3 m^2 in the present design).
- The heat transfer coefficient through the superheater tube is calculated based on the thermal conductivity of high temperature steel which is set to 20 W/(mK), which is reasonable for the temperatures in the superheater tubes. Similar values are determined for the protective tube in stainless steel. The thermal conductivity in the insulation material is based on table values for Portland cement presented by Ichim et al. [13], which is 0.2–3.63 W/(mK). The value chosen was 0.34 W/(mK), which results in a temperature at the outside tube surface equal to 700 °C for operation with sand during 2019 (and which seems correct).
- Additional resistance for heat transfer on the outside of the tubes in the form of ash deposits and effects of corrosion were not considered (but will be commented on below).
- The surface temperatures on the tubes are calculated based on the assumed resistances for heat transfer through the different parts of the tube package (superheater tube, cement, protective tubing).
- In comparison with different periods, the fluidization is assumed to be similar. To support this assumption, it should be mentioned that measurements of the gas flows in the FBHE units (which have been done since 2016) show that the gas flow are rather constant and results in a superficial gas velocity of around 1 m/s in both FBHE units.
- For most of the experimental data presented, only one of the FBHE units (the left one), is included in the results. Overall small differences were seen between left and right FBHE.

Table 1
Key properties and elemental composition [12] of the bed materials.

	Sand (bed sample)	Fresh sand	Fresh ilmenite
Mean average particle size (μm)	186	200	198
Bulk density (kg/m^3)	1458	1458*[1]	2343*[2]
Minimum fluidization velocity u_{mf} (cm/s) [9]	1.25	1.45	2.30
Terminal velocity u_t (cm/s) [10]	95.1	107.7	163.4
Si (wt%)	NA	42.3	0.93
Fe (wt%)	NA	0.35	36.46
Ti (wt%)	NA	0	26.89
Mg (wt%)	NA	0	2.16
Al (wt%)	NA	2.59	0.34
Ca (wt%	NA	0	0.23
Mn (wt%)	NA	0.89	0.04
K (wt%)	NA	1.66	0.02
Cu (wt%)	NA	0	0.01
Total (wt%)		47.80*[3]	67.27*[3]

*[1] It is assumed that the density of the fresh sand is close to that of used sand.
*[2] Based on previous measurements on ilmenite.
*[3] Balance is oxygen.

3.2. Estimation of bed-to-tube heat transfer coefficient with laboratory-derived correlations

In the evaluation of the expected bed-to-tube heat transfer coefficient from fluidized bed to tube wall with correlations, the following assumptions have been made:

- The bed temperature is assumed to be equal to the measured bed temperature in the FBHE.
- The sphericity of both bed material particles is assumed to be 0.65 and the particle emissivity is assumed to be 0.9.
- The heat capacity for the particles is assumed to be 900 J/(kgK) for both materials, and the thermal conductivity is assumed to be 0.27 W/(mK).
- The particle size is assumed to be the mean particle size of the fresh bed material.
- When data is missing to calculate the superficial gas velocity, a value of 1 m/s is assumed.

- Based on a previous study [5], an expression suggested by Grewal & Saxena [14] was determined as the most suitable expression to estimate the bed-to tube heat transfer at elevated temperatures. The contribution from radiative heat transfer has been added to the convective contribution based on the temperature of the outside tube surface. The method presented here is overall similar to the method presented in earlier studies [4,5].
- In order to account for the difference between a single horizontal tube and a complete tube bundle, a correction factor was calculated both for the original design and the current design based on the specification for the heat exchanger. The correction factor was calculated to 0.87 for the original design and to 0.78 for the current design, using the correction factors presented by Basu et al. [11]. The bed-to-tube heat transfer coefficient is calculated as (h_{conv} + h_{rad}) *Correction factor.

4. Results

4.1. Estimation of the bed-to-tube heat transfer coefficient from experimental data

The results indicate that the steam flow is rather constant for the test periods (22–25 kg/s). This is true also for the bed temperature (850–870 °C). The results for 2002 (the first year of operation) are presented (including all hours during the year) for both FBHE units in Fig. 3.

It can be observed that the delivered heat load is approximately 4 MW in each FBHE unit at the very start of the plant, but that it is gradually reduced with time. The reduction in the delivered heat load is most likely due to ash deposits, which are known to build up on heat transfer tubes during operation. Another factor that could potentially contribute is that the bed particles present specifically in the FBHE units gets larger with time, even though there is continuous withdrawal of bottom ash in the furnace. This would have a negative effect on the bed-to-tube heat transfer. Still, fouling due to buildup of ash deposits seems likely to be the main culprit.

The results for 2020 are presented in Fig. 4. The FBHE's has now been outfitted with protective layers.

It can be observed that the delivered heat load is significantly smaller with the new heat exchanger design. Since the heat transfer surface is similar for the two designs, the overall heat transfer coefficient is significantly higher with the original design.

The complete results (including calculated bed-to-tube heat transfer coefficients) for the period 2002–2005 are presented in Table 2. This was before redesign of the FBHEs with a protective layer. As explained earlier, the right FBHE is located upstream of the left FBHE, with respect to steam flow, while the units are mounted in parallel with respect to solids flow. As a consequence of this arrangement, ΔT_{right} and P_{right} are usually higher than ΔT_{left} and P_{left} (Table 3)

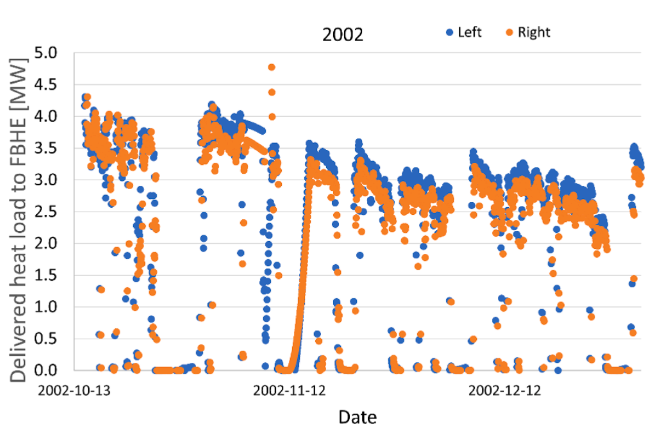

Fig. 3. Estimated delivered heat load in left and right fluidized bed heat exchanger (FBHE) during 2002.

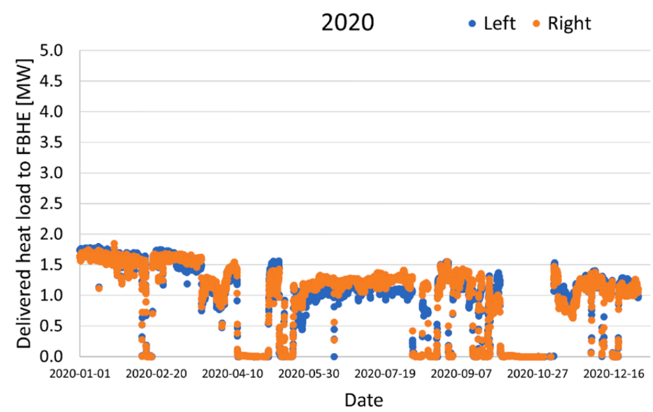

Fig. 4. Estimated delivered heat load in left and right fluidized bed heat exchanger (FBHE) during 2020.

During the period 2002–2005 it can be observed that the delivered heat load (and the temperature difference) in the left and right superheater was at its highest point in 2002, just after the start of the boiler. The temperature in the bed was rather constant for the different occasions. The calculated heat transfer coefficients from bed to tube are lower than expected, based on previous experiments at lab-scale and semi-commercial scale. The estimated bed-to-tube heat transfer coefficients determined at this plant was generally around 100 W/(m^2K).

It can be observed that the heat transferred from bed to tube generally decreased with time, but it can also be seen that there are yearly variations. During the years 2003 and 2004 results for several periods during the year are presented to display the effect of planned stops for maintenance work which usually take place during April and October each year. For example, the period with the lowest bed-to-tube heat transfer coefficient for example (25/9–8/10 in 2003) is just before maintenance work. The subsequent period after maintenance (24/11–31/12 in 2003) has more than twice as high heat transfer coefficient to the tube surface. This underscores the importance of fouling and possibly also formation of agglomerations are for heat transfer.

Another factor which is difficult to separate from fouling is the age of the tubes. Here, it shall be pointed out that the tubes in the old design had to be replaced annually. However, is likely that the decrease in heat transfer due to fouling is more important.

During 2005–2014, the plant was operated normally, using the new FHBH design and silica sand as bed material. This data has not been included in this study. In 2014, operation of the boiler with ilmenite as bed material was tested, and later adapted. In table 3, the results for the time period 2014–2021 are presented. Apart from the parameters described before, the superficial gas velocities are also included in this table, and the bed material type. As expected, it can be seen that the delivered heat load is lower compared to the earlier time period. Also, the variability in heat transfer coefficient from the bed to the tube surface however is higher in the current heat exchanger. For one period the estimated heat transfer coefficient is approaching 500 W/(m^2K), but this can likely be traced to the fact that the assumptions for the insulating layer where the heat transfer coefficient through the insulating tube layer is around 160 W/(m^2K). If the heat transfer through the tube walls was high (meaning that the heat transfer on the outside is limiting the overall heat transfer), an increase in heat transfer on the outside would result in a proportional increase in the U-value. In this case however, h_o must increase significantly to explain changes in the U-value which most likely is not telling the full picture, since it is expected that the bed-to-tube heat transfer is worse than the present design where the spacing for successful bed mixing is lower which is indicated by Basu for example [11].

Table 2
Estimated average temperature differences, delivered heat loads, estimated bed-to-tube heat transfer coefficients for different time periods with the original FBHE (2002–2005).

Year	Date	M_{steam} [kg/s]	$T_{FBHE,left}$ [°C]	$T_{FBHE,right}$ [°C]	ΔT_{left} [°C]	ΔT_{right} [°C]	P_{left} [MW]	P_{right} [MW]	$h_{o,\ left}$ [W/m^{2}K]
2002	30/10–31/12	23.2	854.7	853.2	51.4	53.2	2.90	3.00	187
2003	1/1–31/12	22.5	869.4	861.6	34.5	37.7	1.93	2.10	112
2003	1/1–2/5	22.7	868.2	866.0	35.5	36.6	2.04	2.07	117
2003	1/6–6/9	22.6	871.5	856.1	35.4	40.1	2.03	2.24	115
2003	1/6–15/7	23.1	869.5	859.3	37.9	39.0	2.22	2.23	127
2003	15/7–6/9	22.3	873.1	853.4	33.4	41.1	1.88	2.25	105
2003	25/9–8/10	19.9	873.4	869.6	18.6	18.5	0.93	0.93	50
2003	24/11–31/12	22.8	866.3	852.0	36.8	47.3	2.09	2.69	122
2004	1/1–31/12	22.6	870.4	870.0	32.4	41.6	1.84	2.36	104
2004	1/1–8/5	22.6	868.6	867.5	30.8	39.2	1.76	2.23	99
2004	26/5–31/7	22.8	868.7	866.6	34.5	43.6	1.97	2.49	114
2004	26/5–1/7	22.8	863.0	863.6	41.0	48.5	2.35	2.78	140
2004	1/7–31/7	22.7	874.7	869.9	27.6	38.4	1.57	2.19	88
2004	17/8–31/12	22.4	873.2	870.6	32.6	42.6	1.83	2.39	103
2005	1/1–31/12	22.3	873.0	870.0	26.7	32.1	1.50	1.81	83

Table 3
Estimated average temperature differences, delivered heat loads and estimated bed-to-tube heat transfer coefficients in the left FBHE for different periods with the current FBHE (2014–2021).

Year	Date	Bed material	M_{steam} [kg/s]	$T_{bed,left}$ [°C]	$T_{bed,right}$ [°C]	ΔT_{left} [°C]	ΔT_{right} [°C]	P_{left} [MW]	P_{right} [MW]	$u_{o,left}$ [m/s]	$u_{o,right}$ [m/s]	$h_{o,\ left}$ [W/m^{2}K]
2014	1/1–19/11	Sand	23.93	911	881	18.5	20.2	1.12	1.22			101
2014	19/11–31/12	Ilmenite	25.80	912	882	17.9	20.6	1.17	1.34			109
2015	1/1–6/2	Ilmenite	26.00	906	881	21.6	20.5	1.42	1.34			172
2015	31/3–14/4	Sand	26.54	913	829	24.8	22.4	1.66	1.48			260
2015	15/6–31/12	Sand	25.67	906	879	21.0	19.7	1.37	1.27			157
2016	1/1–1/2	Sand	24.67	912	893	17.8	22.6	1.11	1.40	1.07	1.07	99
2016	23/2–1/3	Ilmenite	25.33	910	897	19.3	22.9	1.24	1.46	1.03	1.03	123
2016	12/8–11/9	Ilmenite	25.02	905	888	21.6	22.1	1.37	1.39	1.01	0.96	159
2017	1/1–25/1	Sand	26.06	905	877	20.3	21.5	1.34	1.41	0.99	0.99	151
2017	22/3–17/4	Ilmenite	25.08	906	882	19.7	19.8	1.25	1.25	0.99	0.99	128
2018	31/8–8/9	Sand	23.76	899	888	19.2	22.5	1.15	1.35	1.03	1.03	113
2018	20/1–30/4	Ilmenite	25.88	903	891	18.9	24.7	1.24	1.60	1.02	0.96	128
2018	13/11–19/12	Ilmenite	26.55	904	889	27.9	26.5	1.87	1.76	1.02	1.02	446
2019	1/1–15/12	Sand	25.46	904	888	23.8	25.0	1.54	1.59	1.03	1.03	216
2019	16/12–31/12	Ilmenite	26.84	910	888	25.2	24.4	1.72	1.63	1.03	1.03	302
2020	1/11–15/1	Ilmenite	26.75	909	891	25.5	24.9	1.73	1.66	1.03	1.03	303
2020	15/1–23/1	Ilmenite	26.20	909	888	25.0	24.7	1.66	1.61	1.03	1.03	264
2020	20/2–2/3	Ilmenite	26.26	904	889	25.3	24.6	1.68	1.62	1.03	1.02	286
2020	17/3–31/12	Sand	24.50	901	892	18.1	19.3	1.13	1.19	1.02	0.98	106
2021	1/1–23/2	Sand	24.33	896	892	16.0	15.8	0.99	0.97	0.94	0.93	85

Fig. 5 can be used to compare the temperature differences for the two different bed materials which have been used in the CFB boiler. The results indicate that there is no clear difference between sand and ilmenite. This is not in agreement with previous studies, where ilmenite was found to result in higher bed-to-tube heat transfer for similar particle sizes [4,5].

Also in the new design, there is a tendency that the heat transfer to

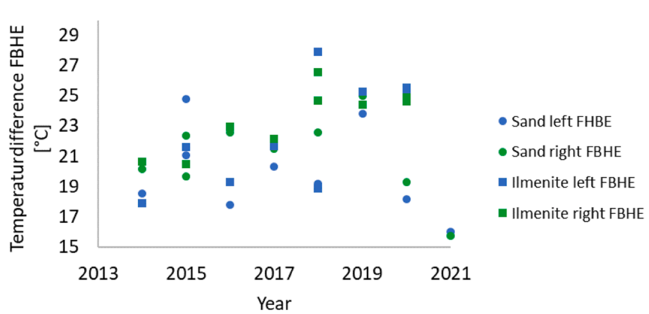

Fig. 5. Estimated temperature difference between left and right fluidized bed heat exchanger at different superficial gas velocities during the period 2016–2021.

the FBHE decreases with time after maintenance stops. However, the effect is much less significant than for the original FBHE design. This is not surprising, since the added resistance has a larger impact on an overall heat transfer coefficient, than fouling could have in the original design. This observation supports the thesis that fouling is responsible for reduced heat transfer in the units during long periods of operation.

Although the right and left fluidized bed heat exchanger usually perform similarly, there have been some exceptions historically. During one maintenance stop, it was observed that a large lump of bed material had been formed at the top of the left FBHE unit, see Fig. 6.

The lump consisted of a mix of bed material, ash and steel wires. It is probable that the lump contributed to uneven distribution of the fluidization gas, so that the gas to a higher extent passed through the lower parts of the tube bundle. This would result in poor mixing in the upper part of tube bundle. It could also have an impact on the circulation of the bed material going from right to left in Fig. 7.

These observations were made for the left fluidized bed heat exchanger. By studying the difference between left and right FBHE (where no lump was identified) during this period since the last maintenance stop during 2021, an interesting difference was seen in terms of heat transfer where $\Delta T_{right\ FBHE}$ - $\Delta Tl_{eft\ FBHE}$ changed from around 1 °C to almost 6 °C from the maintenance stop in spring to the stop during

Fig. 6. Two images of a fluidized heat exchanger (from the top to the left and from the side to the right) where a lump had formed at the top of the tube bundle.

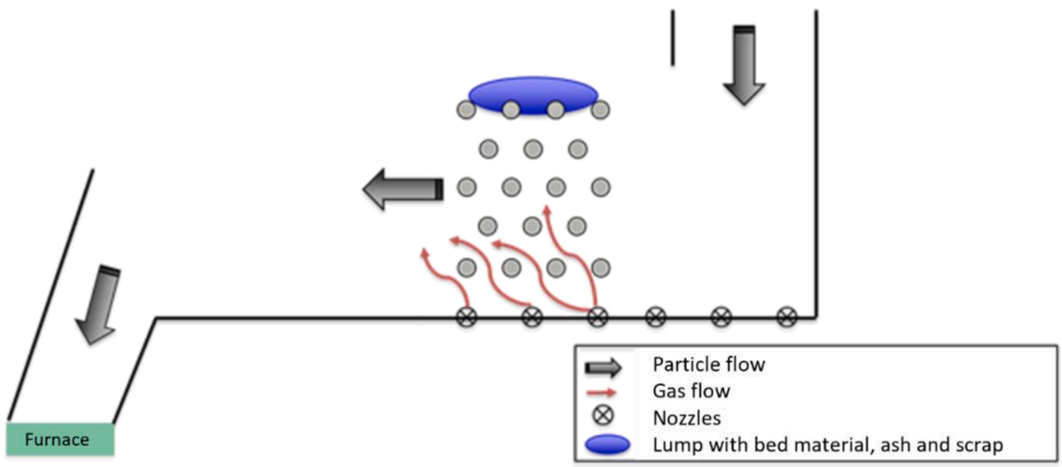

Fig. 7. Illustration of the fluidized bed heat exchanger from the side including a visualization of where the lump was located at the top of the heat exchanger.

autumn. The difference was 4.7 °C on average which was approximately 20 % lower for the left FBHE unit. This shows that the effect of the lump was significant on the temperature difference over the fluidized bed heat exchanger. Since the fluidized bed heat exchangers had been operated in the same way during this period, it is likely that it was the lump that caused this difference. The risk that such a lump is formed in the fluidized bed heat exchanger should be higher with smaller tube pitch. These observations can be seen as an argument for that improved and more even heat transfer to the fluidized bed heat exchangers can be expected with a larger tube pitch.

The experimentally determined bed-to-tube heat transfer coefficients were also compared with values predicted by heat transfer correlations for four different time periods. The results are presented in Table 4.

It can be concluded that the experimentally determined bed-to-tube heat transfer is significantly lower than the values predicted with heat transfer correlations. Essentially, the overall heat transfer coefficient is reduced to a very significant extent in the current design, as compared to what could be expected based on heat transfer correlations. The results are similar for both bed materials (only cases with sand are included in the presented results in Table 4). The reason for why the heat transfer coefficient predicted by correlations is lower in 2019 is because the radiative heat transfer coefficient is significantly lower for that case, because of the different tube surface temperature. The results are not in line with the expectations. The difference is discussed and explained in the following points:

- The measured bed-to-tube heat transfer coefficients are lower than expected. The values for the bed-to-tube heat transfer coefficients for this fluidized bed heat exchanger are generally in the range 50–200 $W/(m^2K)$, whereas previous work rather indicates heat transfer coefficients corresponding to at least 500 $W/(m^2K)$. The measured heat transfer coefficients are also significantly lower than those predicted with heat transfer correlations.
- The superficial gas velocity in this FBHE is higher than those commonly applied in this type of device. Here, the superficial gas velocity is most often around 1 m/s. To the best of our understanding, superficial gas velocities of 0.3–0.5 m/s are used in many designs, albeit this obviously depends on manufacturer and site conditions.

Table 4
Comparison of experimentally determined bed-to-tube heat transfer coefficients and estimations based on correlations for four different time periods with the original and the current FBHE design.

Year	Date	FBHE	$U_{o,\ experiment}$ $[W/m^2K]$	$U_{o,\ corr}$ $[W/m^2K]$	$h_{o,\ experiment}$ $[W/m^2K]$	$h_{o,\ corr}$ $[W/m^2K]$
2002	30/10–31/12	Original	159	465	187	846
2004	1/1–31/12	Original	95	463	104	854
2019	1/1–15/12	Current	73	75	216	690
2021	1/1–23/2	Current	48	77	85	844

- The tubes are thick and the resistance through the tube wall is large. It should be mentioned that it is difficult to find dimensions of operational tube bundles for industrial FBHE units, but the expectation from the authors (in terms of ratio between outer diameter and inner diameter where steam is flowing) is that common dimensions are in line with the original design of this unit.
- The tube spacing is small. This accentuates even more the second point above, i.e. high gas velocity between the tubes. This could affect the fluidization of the bed material as well as the lateral transport of bed material through the FBHE. Based on the observations related to lumps of materials forming at the top of the tube bundle, it is suggested that an increase in tube spacing should have a positive impact on bed mixing and heat transfer from bed to tube. It should also be mentioned that the heat transfer correlations from literature (including Grewal & Saxena used here) are based on the superficial gas velocity. Most such these correlations are derived from experiments with single tubes and have not used real heat exchangers to validate their correlations. It is possible that the effect of high real velocity in designs with small tube spacing is not accounted for in a suitable manner in the expression of Grewal & Saxena, even though the correction factor is considered to adjust the effect of multiple tubes.
- Significant variations in heat transfer from bed to tube are seen over time. This is not so surprising. Most experimental studies take place in environments without presence of bed ash and the tests are usually performed during shorter periods of time. This plant utilizes fuels with high content of aggressive ashes and the heat transfer was monitored for several years. Thus, it is expected that the results vary significantly over time. It is observed that heat transfer increases almost every year just after the planned stop for boiler maintenance work. This makes sense since maintenance involves e.g. removal of bed agglomerates and maintenance of surfaces for reduced fouling.
- When comparing the two FBHE designs, the assumptions related to the insulation tube layer for the new heat exchanger has a significant effect on the calculated bed-to-tube heat transfer coefficient. In the current tube design the resistance over the insulating tube layer is dominating the overall resistance for heat transfer in the FBHE unit which makes it more difficult to evaluate the bed-to-tube heat transfer and compare the two designs.

Based on the observations in this work, it is suggested a new FBHE design includes larger tube spacing in the horizontal direction to improve the heat transfer coefficient from bed to tube per heat transfer area. Since the available volume in the FBHE unit is limited, the number of tubes must be reduced which in turn affects the total heat transfer area. It is likely that this could be compensated for by improved heat transfer but in case this is not possible, the injection of feedwater could be reduced between the right and left fluidized bed heat exchanger to maintain a suitable temperature on the outgoing steam of 450 °C.

This design should allow for operation with a lower superficial gas velocity and therefore also a lower risk of tube erosion. It should also reduce for risk of lumps forming between the tubes, which could have several undesired effects. A lower risk of tube erosion with lower gas velocities and better mixing of the bed material in the heat exchanger should contribute to a longer lifetime of the heat exchanger as well.

5. Conclusions

This work evaluates the bed-to tube heat transfer at fluidized bed heat exchangers in a 75 MW_{th} CFB boiler fired with municipal solid waste and industrial waste, over time periods that encompasses several years. Two different designs for the FBHE were investigated. The old design is rather conventional, while the new design includes additional protective tubing and an insulation layer to reduce the risk of corrosion. It was possible to draw the following conclusions in this study:

- The experimentally determined bed-to-tube heat transfer coefficient (h_o) were in general quite low (100–150 W/m^2K), both for the original FBHE design and the new design. The heat transfer was significantly lower than previous experimental investigations at high bed temperatures in lab scale, and also lower than estimations calculated via correlations from literature.
- The new design is believed to have too narrow tube spacing. This seems to affect the bed-to-tube heat transfer in a negative manner, likely due to poor mixing. It is believed that bed mixing is less good and even completely interrupted in parts of the tube bundle. Operation with larger tube spacing should allow for better mixing, improved heat transfer and open the possibility to have lower superficial gas velocities.
- However, it should be acknowledged that it probably is difficult to find the right balance between heat transfer and corrosion risks in FBHE units, especially where fuel input in this plant includes significant amount of alkali.
- To perform accurate estimations of bed-to-tube heat transfer and especially for comparisons of different heat exchangers designs, it is important to have the limitations for the overall heat transfer on the outside of the tubes. This is difficult in this plant where the current design involves an insulation layer with high resistance for heat transfer.
- Fouling on the heat exchanger tubes seems to increase with time since the last maintenance work in general but the effect is lower in the present FBHE unit where the overall heat transfer coefficient is lower.

CRediT authorship contribution statement

Viktor Stenberg: Methodology, Formal analysis, Investigation, Data curation, Writing – original draft, Visualization. **Magnus Rydén:** Methodology, Writing – original draft, Writing – review & editing, Supervision, Project administration, Funding acquisition. **Fredrik Lind:** Methodology, Investigation, Resources, Data curation, Writing – review & editing, Supervision.

Declaration of Competing Interest

The authors declare that they have no known competing financial interests or personal relationships that could have appeared to influence the work reported in this paper.

Data availability

The authors do not have permission to share data.

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