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The heat transfer between an immersed surface of moving lignite and small particles in a fluidized bed equipped with an inclined slotted distributor

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ABSTRACT

In the process of fluidized bed combustion, the particle-convective component (h_{pc}) can evidently affect the overall heat transfer coefficient (h_o). Therefore, a vigorous mixing of fuel that enhances heat transfer is required. The use of an inclined slotted distributor (IS distributor) is a promising technique to improve the gas-particles contact and particle dispersion. In the present study, experiments were carried out in a fluidized bed equipped with an IS distributor, and the surface-renewal theory was used to investigate the heat transfer between an immersed surface of moving lignite and small particles. The effects of excess gas velocity (U_e), immersed surface diameter (d_l), and small particle size (d_p) on h_o were discussed, and a comparison with the conventional fluidized bed was made. The test results confirm that the use of a bed equipped with an IS distributor is effective to enhance the heat transfer performance when compared to a conventional bed. Based on the test results obtained at U_{mf} and different U_e, the parameters of the surface-renewal theory were determined, and the correlations were established for predicting h_o. A good agreement between the predictions and test results could be achieved when the radiative component (h_r) is negligible.

1. Introduction

Low-rank coals (LRCs), which are mainly lignite, account for a large portion of the world coal reserve [1]. In China, LRCs, especially lignite, play a significant role in supplying primary energy because of their abundance, easy access, and low mining cost [2,3]. Fluidized bed combustion as one of the lignite combustion technologies is widely applied in power industry. In combustion, a definite amount of lignite is supplied to the bed at low temperatures, together with additions of small limestone particles, offering significant advantages such as fuel flexibility, in-bed sulfur capture, and relatively low NO_x emissions [4–6]. The heat liberation in the combustion of the fuel components of the carbon, hydrogen, sulfur is consumed in heating the lignite, as well as the limestone. Therefore, an efficient combustion requires the vigorous mixing between moving lignite and small limestone particles, which enhances heat transfer rate.

The major factor that affects the intensity of mixing is the arrangement of the gas flow through the distributor. Practically, various distributor designs have been created for a fluidized bed. The designs can be generally categorized into two configurations: static and rotating distributors [7]. According to the direction of gas entering the bed, the distributors can be further classified, namely, normal direction (perforated-type distributor), lateral direction (multi-vortex tube, bubble caps), and inclined direction (inclined slotted distributor) [8–10]. A fluidized bed with a rotating distributor is an efficient gas-particles reactor and thus has been widely applied to improve the heat transfer. As stated in the literature [7,11], the rotational motion of the distributor greatly improves the particle mixing and the distribution of gas-particles over the cross-section, thus enhancing the heat transfer rate inside the bed. Recently, inclined slotted distributor (IS distributor) has attracted much attention in the application of a fluidized bed. The gas entry through the inclined slots of the distributor is divided into two velocity components: The vertical velocity is responsible for fluidization, and the horizontal velocity contributes to the swirling movement in a bed [10,12,13]. Without using an electric motor or mechanical rotation of a distributor, excellent gas-particles contact and particle dispersion could be achieved using an IS distributor.

The location and magnitude of immersed surfaces in a fluidized bed greatly affect the heat transfer process and the contribution of heat transfer mechanism. Previously, many experimental studies have been made on the heat transfer between a fluidized bed and immersed surfaces. These studies focused on the following aspects: heat transfer on tubes submerged in bubbling dense beds, heat transfer for membrane-walls in a large-scale CFB combustor, heat transfer on water walls of a

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Nomenclature

A_I	surface area of immersed surface, m^2
$C_{p.p}$	specific heat capacity of particulate phase, J/kg k
$C_{p.g}$	specific heat capacity of gas phase, J/kg k
$C_{p.I}$	specific heat capacity of immersed surface, J/kg k
d_I	immersed surface diameter, mm
d_p	small particle size, μm
h_o	overall heat transfer coefficient, w/m^2 k
h_r	radiative component, w/m^2 k
h_{gc}	gas-convection component, w/m^2 k
h_{pc}	particle-convective component, w/m^2 k
h_1	heat transfer coefficient of gas film, w/m^2 k
h_2	heat transfer coefficient of penetration layer, w/m^2 k
k_g	thermal conductivity of gas, w/m k
k_p	thermal conductivity of particulate phase, w/m k
$k_p{}^0$	thermal conductivity of particulate phase with stagnant gas, w/m k
k_s	thermal conductivity of small particle, w/m k
$m(d_p)$	function with d_p as independent variable, –
$n(d_p)$	function with d_p as independent variable, –
R_1	thermal resistance of gas film, m^2 k/w
R_2	thermal resistance of penetration layer, m^2 k/w
$R_p{}^c$	thermal resistance of particle-convective, m^2 k/w
t	time, s
t_0	the time when placing the immersed surface into the fluidized bed, s
U_e	excess gas velocity, m/s
U_s	superficial gas velocity, m/s
U_{mf}	minimum fluidizing velocity, m/s
V_I	volume of immersed surface, m^3
ε_{mf}	voidage in particulate phase at minimum fluidizing velocity, –
Re_{mf}	Reynolds number of immersed surface at minimum fluidizing velocity, –
Pr	Prandtl number, –
δ	gas film thermal resistance constant, –
τ	residence time, s
ρ_p	density of particulate phase, kg/m^3
ρ_g	density of gas phase, kg/m^3
ρ_s	density of small particle, kg/m^3
ρ_I	density of immersed surface, kg/m^3
φ	parameter in calculation of k_p, –
θ_w	wall temperature, °C
θ_I	temperature of immersed surface, °C

commercial CFB boiler, and heat transfer for an external heat exchanger (evaporator, superheater, or reheater) in CFB boilers [14–18]. Although some experiments were made on heat transfer for a fixed surface in conventional fluidized beds, the results cannot be directly applied to this work for the mobility of immersed surface and the bed with a different type of distributor. Accordingly, the heat transfer in a fluidized bed equipped with an IS distributor is unclear. To achieve reasonable predictions on the ignition and combustion temperatures of lignite for a better configuration design, it is essential to study the heat transfer between the immersed surface of moving lignite and small limestone particles in a fluidized bed equipped with an IS distributor. Since the lignite size is always larger than the limestone particles in an industrial fluidized bed, measurement of heat transfer for spherical lignite moving in a bed of small limestone particles was conducted to simulate the actual conditions. The purpose of this work was to compare the effect of different factors on the heat transfer coefficient of moving lignite for using two beds (conventional fluidized bed and fluidized bed equipped with an IS distributor). Finally, the obtained results were used to determine the parameters of the surface-renewal theory for better predicting the heat transfer coefficient of moving lignite in a fluidized bed equipped with an IS distributor.

2. Heat transfer in a fluidized bed

The overall heat transfer coefficient, h_o, between an immersed surface of moving lignite and a fluidized bed is mainly composed of three components: gas-convection, h_{gc}, particle-convection, h_{pc}, and radiation, h_r:

$$h_o = h_{gc} + h_{pc} + h_r \tag{1}$$

The operating temperature for a lignite fluidized bed boiler varies from 600 to 900 °C [19]. The contribution of the radiative component, h_r, becomes significant only when the temperature in the bed is above 600 °C [20]. Most of the investigations focused on the convection part, and the bed temperature lower than 600 °C were always picked to ensure that h_r is negligible. This is the reason that the test temperature is chosen to be 600 °C in the following experimental section.

In an industrial fluidized bed, the contribution of gas-convection h_{gc} to the overall heat transfer coefficients, h_o, is much less than that of particle-convection h_{pc} due to the high heat capacity of the particles and low minimum fluidizing velocity (U_{mf}) [21,22]. Based on the following correlation proposed by Agarwal [23], the h_{gc} between gas and the immersed surface can be determined.

$$\frac{h_{gc} d_I}{k_g} = 2\varepsilon_{mf} + 0.693 Re_{mf}^{1/2} Pr^{1/3} \tag{2}$$

The particle-convective component, h_{pc}, clearly affects the overall heat transfer coefficient, h_o. Different theories have been proposed for determining the h_{pc}, the most commonly used theory for immersed surface in a fluidized bed is the surface-renewal theory. The major feature of the surface-renewal theory is the assumption that particles in random motion; some of the particles remain on the immersed surface for some time and then replaced with new particles. Various researchers have used this theory to investigate the heat transfer behavior in a fluidized bed. Karimipour et al. [24] developed the concept of temperature penetration depth to evaluate the heat transfer behavior near the wall of gas-solid fluidized beds according to the surface-renewal theory. Blaszczuk et al. [22] used the surface-renewal theory to assess the effect of bed particle size on heat transfer process from the core region to the water walls with rifled tubes in a 966 MWth CFB boiler. Zarghamia et al. [25,26] predicted the particle residence time on a heat exchanger surface based on particle behavior near the surface using the surface-renewal theory. Dutta et al. [18] improved the surface-renewal theory using a derived correlation for the wall coverage and other modified information from commercial CFB boilers. This theory always has the form as follows [27]:

$$\frac{1}{h_{pc}} = \frac{1}{h_1} + \frac{1}{h_2} \quad \text{or} \quad R_{pc} = R_1 + R_2 \tag{3}$$

The thermal resistance from the particulate phase (R_{pc}) refers to two resistances (R_1, R_2) to heat transfer in series, in which R_1 and R_2 are the thermal resistance of gas film and penetration layer, respectively.

The expression of the thermal resistance R_1 is shown in Eq. (4) [28,29]:

$$R_1 = \frac{1}{h_1} = \frac{d_p}{\delta k_g} \tag{4}$$

As a result of the low heat capacity of gas compared to the particles and low gas-convection contribution, R_1 is mainly from the resistance to conduction across the gas film with a thickness of d_P/δ between the