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Thermo-fluid characterization of flue gas flows through a packed bed

Nat Vorayos*, Nakorn Tippayawong and Bancha Thanapiyawanit

Department of Mechanical Engineering, Chiang Mai University, Chiang Mai 50200, Thailand

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Abstract

Enhancement of thermal performance in an incinerator and flue gas treatment with a randomly packed bed of uniform spheres was investigated in a custom-made experimental facility and reported in the present paper. Pressure drop, temperature evolution and heat transfer characteristics were evaluated for a range of superficial velocities, Reynolds numbers and bed geometries. Results revealed that increases in both superficial velocity and bed thickness caused a rise in pressure drop across the packed bed in similar fashion to the Ergun equation but with different coefficients at low Reynolds number between 60-300. The two constants were empirically determined to be 68.5 and 4.95. It was also found that the packed bed affects axial temperature distribution from the incinerator chamber to stack. For the same heating rate, higher temperature was achieved in the chamber with the packed bed, and peak temperature was reached at a rate of 7-10% faster than that without the packed bed during the startup period.

The downstream side of the thicker bed appeared to have lower temperature than the thinner bed for the same axial position, demonstrating less flue loss during the transient state. The incinerator chamber proved to have higher temperature rising rate and reached higher maximum temperature with the presence of the packed bed.

Keywords: Flue gas; Incinerator; Low reynolds number; Packed bed; Spheres

1. Introduction

Pollutants emitted from waste incinerators include volatile organic compounds, unburned hydrocarbons, carbon residue, soot, and ash. Sufficiently high temperature is essential to obtain destructive conditions for various pollutants arising from an incinerator. Normally, a secondary chamber is utilized in order to ensure complete destruction of these pollutants in flue gas at high temperature for a certain period of residence time. Typically, the required condition is to have temperature at greater than 800℃ for at least 1.0 second. Higher chamber temperature results in lower value of required residence time and may reduce the overall chamber size. This may be achieved via adoption of a second fuel burner, a thermal oxidizer or an expensive catalytic converter. Alternatively, distributed resistant flow or flow through porous media can sustain uniform flow and prolong the residence time of the fluid. It is able to provide more complete reaction by providing (i) a larger surface area for heat transfer and fluid contact, (ii) better turbulence or mixing, (iii) longer residence time, and (iv) repetitive contacts. A packed bed can offer these advantages for flue gas treatment. Although flow becomes restricted, it can be kept at higher temperature for longer duration. A packed bed consists of packed solid material in a storage container through which a fluid is circulated. There are four packed bed structures, depending on the arrangement of balls in the bed: body-centered cubic, simple cubic, face-centered cubic, and random type. At the same Reynolds number, the face-centered cubic structure yields the lowest pressure drop and the second lowest is random type [1] Porosity or void ratio in a packed bed depends on the bed structure and ratio of bed to particle diameter, D_e/d_P [2] The local porosity causes fluctuation in velocity

*Corresponding author. Tel.: +66 5394 4146, Fax.: +66 5394 4145
E-mail address: nat@dome.eng.cmu.ac.th
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profile, especially at the position adjacent to the wall. This causes fluid at this location to have shorter residence time than at the core. However, this problem can be neglected in the case of D_e/d_P greater than 10 [3, 4] The heat transfer mechanism in a packed bed is dominated by convection. This is because flows in a packed bed generate blockage and wake coalescence so high that local velocities become more turbulent than free stream velocities near both ends [5].

With these characteristics, packed beds are extensively used in chemical and process industries as thermal storage, reactors, filters or heat exchangers. The interest in packed beds for these applications is clearly reflected in a large number of previous studies. Relevant recent literature includes that of Wen and Ding [6] whose work involved investigation of transient and steady state heat transfer of a gas through a packed bed at constant wall temperature. Effective thermal conductivities and convective heat transfer coefficients were derived and heat transfer correlations were compared with previously published works. Schroder et al. [7] presented transient heat transfer data between hot gas and particles of packed beds at relatively low Reynolds number. They reported that experimental data showed a large variance especially at low Reynolds number. They proposed logarithmic probability distribution functions for the mathematical treatment of packed bed heating. Laguerre et al. [8] investigated heat transfer between wall and a packed bed at low velocity airflow. They found that the air velocity and position along the wall affected heat transfer greatly. Singh et al. [9] reported the results of an experiment to investigate heat transfer and pressure drop characteristics of a packed bed system with large-sized packing material of different shapes. Heat transfer and flow correlations were developed for Nusselt number and friction factor. Gunjal et al. [10] considered fluid flow through an array of spheres with different periodically repeating arrangements of particles. Single-phase flow through these geometries was simulated by computational fluid dynamics. Montillet [11] presented and discussed pressure drop evolution in a packed bed of spheres with emphasis on transition to turbulent flow regime. Liu and Hsieh [12] reported that a combustion reaction in a packed bed can lower flammability limit due to mechanism of mixing and heat transfer from reaction zone to upstream and the ability to retain heat in the packed bed. At high excess air condition, less pollution is obtained in comparison with free flame combustion. Jugjai et al. [13] employed a packed bed as a heat recovery device. Two packed beds were constructed at both ends of the chamber as a heat exchanger to absorb the flue gas heat before leaving. This mechanism was found to improve combustion efficiency and ensure uniform temperature distribution in the chamber. Features of the packed bed combustor are capable of achieving high destruction efficiencies for different gaseous species. Nevertheless, it must take into account the effect of pressure drop across the bed in order to assess the suitability of the packed bed combustor.

During the literature survey, it was observed that reports on low velocity and low Reynolds number in a packed bed were rather scarce. Studies on utilization of high temperature packed bed for flue gas treatment were even fewer. In this paper, the main aim is to investigate the thermal performance of hot flue gas flow through a combustion chamber with a packed bed. The pressure drop and temperature evolution in a combustor with and without a packed bed during transient operation were measured. The motivation arises from an ongoing study on treatment of flue gas from a waste incinerator where the hot packed bed may offer an attractive alternative.

2. Experimentation

The motivation of the present investigation on heat transfer in packed beds comes from waste incinerators. Initially, the effect of heat on the packed bed is not considered. The experimental setup, illustrated in Fig. 1, is custom-designed and built. It consists of a randomly packed bed of uniform ceramic spheres in a steel containment, a gas burner, a flow arrangement and a measurement system. The experimental setup is fully thermally insulated. The rectangular steel containment forms a chamber with cross section of $0.16 \times 0.16 m^2$ and 0.60m in height. The ceramic balls are 15mm in diameter. The bed has D_e/d_P of 10 and porosity of 0.41. The bed porosity was determined by water replacement in the void. A perforated grate was used to support the packed bed and allowed the gas through. A bottom to top gas flow through the packed bed is ensured by aspiration with a draught fan. Six type J thermocouples were used to measure the temperatures of the chamber, the interior of the packed bed and the bed exit. Thermocouples were carefully inserted through ceramic straws protruding into the bed. Particular attention was paid to minimize distur-

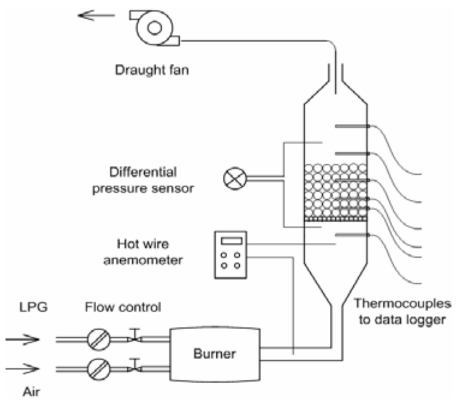


Fig. 1. Schematics of a packed bed experimental setup.

bance to the flow and temperature fields. Axial temperature profile was measured in the center of the chamber at six axial positions. The first one was set immediately upstream of the packed bed and the other five were set at of 30, 60, 120, 170 and 210mm above the grate. All thermocouples were connected to a data acquisition system inside a personal computer and temperature readings were taken every 10 seconds. Pressure drop was measured by a differential pressure sensor. Inlet airflow was measured by a hot wire anemometer and controlled by a regulator valve. Fuel gas flow rate was measured and controlled by means of a regulator valve and a variable area flowmeter. Two sets of experiments were carried out: cold and hot gas flows. In the cold flow experiment, steady state tests were undertaken without heating for a range of superficial velocities with Reynolds number between 60-300 and L/d_P in a range of 0, 6.7, 10.0 and 13.3, respectively. The second set of experiments was performed at a transient condition with hot gas flow through the packed bed. The hot gas was provided from combustion of liquefied petroleum gas – air mixture at 35% excess air. Constant heating rate was provided at about 5.0kW. The hot flow test was conducted by varying only L/d_P. The temperature evolutions were recorded from a cold start. The convective heat transfer coefficient was calculated from the transient operation and used to evaluate flue loss. Comparison was made between the combustors with and without the packed bed.

3. Results and Discussion

3.1 Pressure drop through the packed bed

It is well known that pressure drop through a ball-packed bed is a function of gas velocity over a range of flow rates, and can be predicted by the Ergun equation in the following form,

$$\frac{\Delta P}{L} = a\frac{(1-\varepsilon)^2}{\varepsilon^3}\frac{\mu \cdot u}{d_p^2} + b\frac{(1-\varepsilon)}{\varepsilon^3}\frac{\rho \cdot u^2}{d_p} \quad (1)$$

where a and b are 150 and 1.75, respectively. In this study, the pressure drop across the packed bed as a function of the superficial gas velocity is shown in Fig. 2 for a range of bed heights, L/d_P = 6.67, 10, and 13.33. The low gas velocities tested were between 0.07-0.3m/s, corresponding to sphere Re between 60 and 300. A line that best fit the measurement results is also included. Along with the best fitting curve, comparisons with the Ergun [14] equation, Macdonald et al. [15] Fand et al. [16] and Yu et al. [17] modified equations for pressure drop prediction are illustrated in Fig. 2. Revised Ergun coefficients were obtained from these subsequent experimental studies and shown in Table 1. Also shown in Table 1 are applicable ranges of Re and their corresponding flow regimes. It was found that a tendency for pressure drop in terms of superficial velocity to power two was readily captured. The increase of superficial velocity and bed height to particle diameter ratio was found to raise the pressure drop of flow through the packed bed. However, models from previous literature appeared to underestimate pressure drop per unit bed thickness. Two separate lines can be clearly identified. Within the range considered, it can be seen that qualitatively good agreement between the experimental results and prediction was obtained only for extremely low superficial gas velocity. A discrepancy was found to occur at relatively higher velocities. The finding from this study was similar to that from the work of Wen and Ding [6] but in contradiction to Montillet [11] for the similar range of superficial velocities. The observed differences between the measurements and the mathematical models may be attributed to the fact that the flows in this study (60 < Re < 300) fall in the transition domain which occur between the unsteady laminar flow regime and the chaotic, turbulent flow regime. Seguin et al. [18, 19] who studied the flow transition in packed beds observed that the end of purely laminar flow regime was at Re close to 100 and suggested that the turbulent flow regime was fully established at Re about 530. A difference in Re range between this study and the existing literature is believed to be the reason why there

Table 1. Ergun coefficients and applicable Re ranges to determine pressure drop across a packed bed.

References	*a*	*b*	*Re*
Ergun (1952)	150	1.75	> 80
Macdonald et al. (1979)	180	1.80	n/a
Fand et al. (1987)	225	1.61	120 – 408
Yu et al. (2002)	203	1.95	750 – 2500

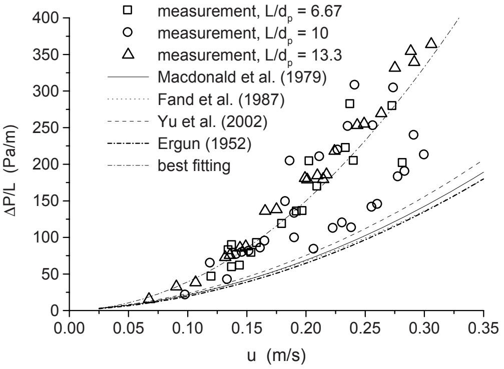


Fig. 2. Pressure drop as a function of superficial velocity. Comparison of experimental data with published correlations. (d_p = 15mm, ε = 0.41, T = 25°C).

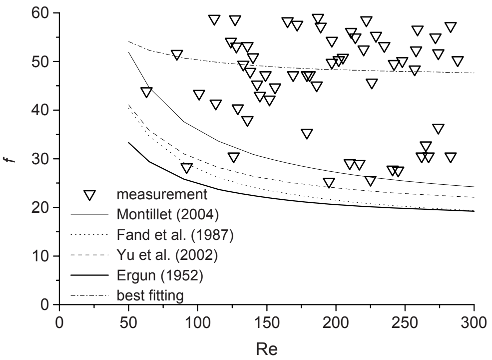


Fig. 3. Friction factor as a function of Reynolds number. Comparison of experimental data with published correlations. (d_p = 15 mm, ε = 0.41, T = 25°C)

are two separate lines seen in Figs. 2 & 3. It can be seen that bed heights influence the pressure drop as well. At *L/dp* = 13.33, the largest pressure drop is obtained while the lowest is reached at *L/dp*= 10 according to which the relation between flow pattern within packed bed and *L/dp* is yet to be determined. An attempt was made in this work to empirically obtain a mathematical expression for prediction of pressure drop evolution in the packed bed. For a correlation from best curve fitting developed here, the two constants were empirically determined to be 68.5 and 4.95 at low Re. The modified constants obtained appeared to differ greatly from Ergun and other formulations which stay approximately in similar ranges. Further analysis was carried out to compare friction factors. Based on the data from the present experimental investigation, the friction factor was calculated in similar fashion to Ha et al. [20] A comparison was made further for friction factor and its predictions from Fand et al. [16] and Yu et al. [17] respectively.

$$f = \frac{\Delta P}{L}\frac{d_p}{\rho \cdot u^2} = \frac{a}{\mathrm{Re}}\frac{(1-\varepsilon)^2}{\varepsilon^3} + \frac{b\,(1-\varepsilon)}{\varepsilon^3} \quad (2)$$

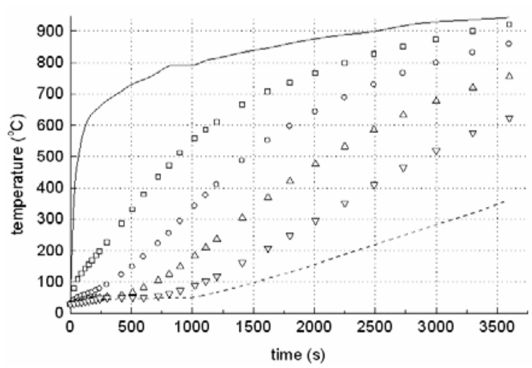
A modified friction factor was also adopted from Montillet [11] and included in this study.

$$f = \frac{1410}{\mathrm{Re}} + 16 + \frac{45}{\mathrm{Re}^{0.45}} \quad (3)$$

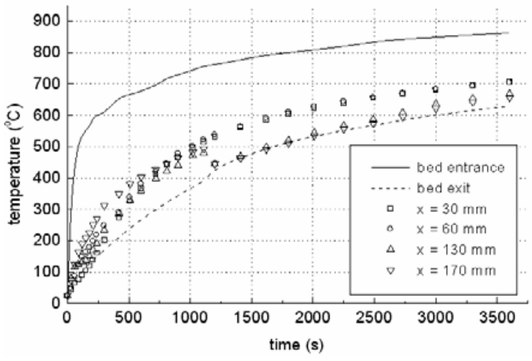
Fig. 3 illustrates the friction factor as a function of Reynolds number. Results from measurements, the best fitting line, and models from existing literature were plotted. Again, two distinct groupings may be observed, one for experimental results and other for numerical modeling. All models were found to underpredict flow resistance in the packed bed in this range of Reynolds number (60-300). It was suggested that Re achieved in this investigation may fall in the laminar/turbulent transition regime. Even though the results were observed to be scattered, the average friction factors were in similar magnitude to those reported in the literature for the same range of Reynolds number. Nonetheless, in this representation, the models were found to significantly underestimate the experimental measurements, inline with pressure drop results. Differences may be due to the effect of transition and flow unsteadiness, resulting from the finite character of the bed, the roughness of the balls' surface, the wall effect and the flow direction variation in the packed bed. The model from Montillet [11] derived from a study of transition flow in packed beds seemed to offer a better performance than others.

3.2 Temperature profiles in the packed bed

Fig. 4(a) and (b) show time evolution of temperature profiles with and without the packed bed. The detailed temperature distribution in the interior of the packed bed and the temperature difference between the packed bed and flowing gas can be derived. Temperatures at the bed entrance and bed exit may be used to represent the incinerator chamber temperatures and temperature of the gas exhaust to stack, respectively. It was found that initially the temperatures increase rapidly, but the rate of increase declines with an increase in temperature, approaching steady state values. The presence of the packed bed was found to affect the temperature distribution greatly. One can see that without the bed, the temperatures at different axial positions vary slightly, but with the bed, a large variation of axial temperatures was observed. Larger temperature drop was mainly due to heat transfer from the hot gas to the packed bed. A comparison between Fig. 4(a) and (b) indicates that higher chamber temperature was achieved at 7-10% faster ascending rate for the incinerator with the bed, compared to that without. During the startup period, the packed bed acted as thermal energy storage where heat was transferred to the bed material. From Fig. 5, the temperature in the incinerator chamber upstream of the packed bed was found to be higher with the presence of the packed bed, reaching 900°C more rapidly. For a given axial position downstream of the grate, initial temperatures in the chamber with the packed bed were lower than those without the packed bed. However, after about 3600s, the packed bed appeared to exhibit higher values than those can be achieved with the void chamber. The flue gas temperature exhaust to stack was always lower at all times. This resulted in less flue loss, as shown in Table 2, which was inversely proportional to the bed thickness. Flue loss is defined as a ratio between flue gas thermal energy and heat input to the bed.

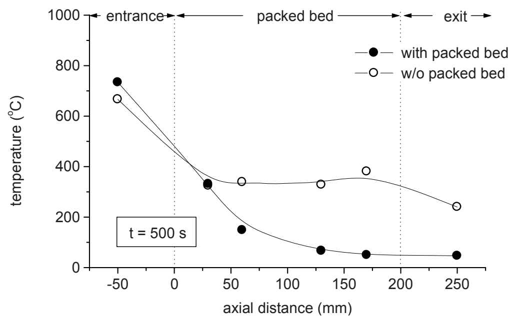


(a)

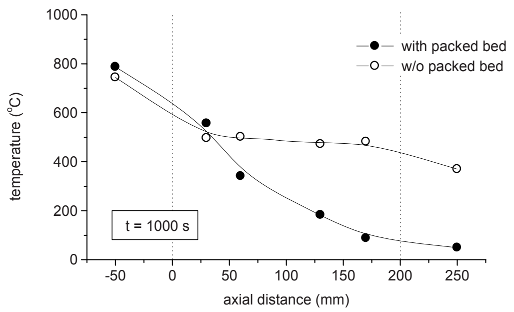


(b)

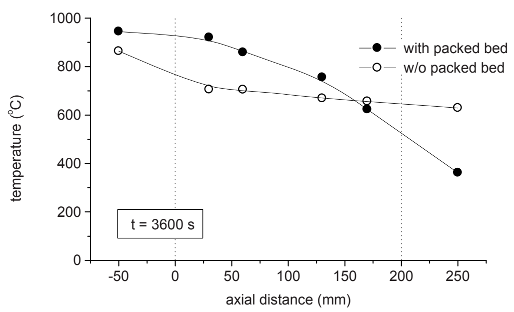
Fig. 4. Temperature evolution along the incinerator chamber (a) with the packed bed and (b) without the packed bed.



(a)



(b)



(c)

Fig. 5. Axial temperature distribution in the packed bed during the start-up period.

Table 2. Thermal energy loss from flue gas exhaust to stack as a function of the bed thickness.

	w/o packed bed	with packed bed		
	$L/d_p = 0$	$L/d_p = 6.67$	$L/d_p = 10$	$L/d_p = 13.3$
Flue loss (%)	21.7	15.1	4.9	4.5

Transient heat transfer modeling was also performed to derive the mean heat transfer coefficient. The thickness of the bed was divided and each level thickness was set to d_P so that a ball can represent a temperature distribution at each level. Inlet gas flow transfers heat to the 1st row of the packed bed and exits. Gas temperature was calculated prior to entering the successive row. This step was repeated until reaching the last row and the destination time was met. The heat balance of the ball is in the general form of transient heat conduction in spherical coordinate [17].

$$\frac{1}{\alpha_m}\frac{\partial\theta_m(t,x,r)}{\partial t} = \frac{\partial^2\theta_m(t,x,r)}{\partial^2 r} + \frac{2}{r}\frac{\partial\theta_m(t,x,r)}{\partial r} \tag{4}$$

The proper initial condition for all spheres in the chamber is listed according to its room temperature at 298K such that

$$\theta_m(0, x, r) = 298.15 \tag{5}$$

Two boundary conditions are needed to solve this equation. As the temperature at the center of the spherical ball is bounded and heat is assumed to be transferred from the ball from the heat convection at ball's surface, the boundary conditions are,

at $r = 0$

$$\alpha_m \frac{\partial\theta(t,x,r)}{\partial r} = 0 \tag{6}$$

at $r = d_p/2$

$$\alpha_m \frac{\partial\theta(t,x,r)}{\partial r} = h_c[T(t,x) - \theta_m(t,x,\tfrac{d_p}{2})] \tag{7}$$

where α is thermal conductivity (W/m.K) of the spherical ball. Since a packed-bed is divided into layers; each is represented by a row of spheres and Eq. (4) is solved for each time step. The temperature of the layers of spheres is then used to determine the fluid temperature change via the energy balance between the layer of spheres (with the porosity of ε) and hot fluid from the equation

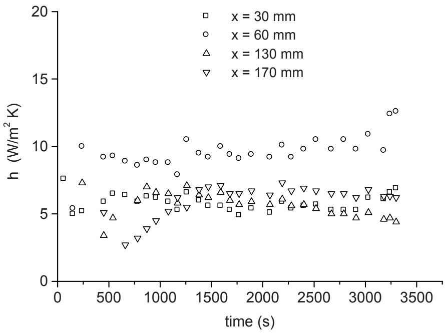


Fig. 6. Estimated convective heat transfer coefficient along the packed bed.

$$A_o(1-\varepsilon)\rho\left(C_{p,b}\frac{\partial\theta_m}{\partial t}\right)_{ball} = \dot{m}c_p\frac{\partial T}{\partial x} \tag{8}$$

The temperature of the next layer is then calculated by using the fluid temperature passing from the previous layer as the boundary condition. The repetition continues along both the longitudinal axis and time domain. The required variable inputs were inlet flow temperature, initial temperature of the packed bed, thermal properties, destination time, bed thickness, and convective heat transfer coefficient as boundary conditions. The average convective heat transfer coefficient during transient state depending on time-temperature gradient was derived from heat exchange at the ball surface between the bed and hot air during each short period by using the following expression [5]

$$h = \ln\left(\frac{\theta - T}{\theta_0 - T}\right)\frac{d_p\rho\ c_p}{6(t-t_0)} \tag{9}$$

Fig. 6 shows the mean heat transfer coefficient as a function of the axial position. Heat transfer was highly influenced by inhomogeneities in ball distribution and arrangement. Furthermore, a nonuniform axial as well as radial dispersion in the packed bed may have played an important role. This led to substantial scattering of data. Nonetheless, the values appeared to vary markedly with axial position but not with time. The mean heat transfer coefficients were between about 5-12W/m^{2}K, having higher values inside the bed than both ends of the bed, as observed by [5]

3.3 Uncertainty analysis

Experimental uncertainty in measurement catego-

Table 3. Inaccuracy margin of measurements and properties.

	Symbol	Nominal value	Measurement uncertainty
Pressure drop	ΔP	30Pa	10.0%
Bed length	L	200mm	5.0%
Ball diameter	d_p	15mm	6.7%
Gas velocity	u	0.18m/s	6.6%
Temperature difference	ΔT	20K	25.0%
Time interval	t	10s	2.0%
Gas specific heat	c_p	1.06kJ/kgK	1.0%
Gas density	ρ	0.32kg/m^3	1.0%
Gas viscosity	μ	4.50×10^{-5}Ns/m^2	1.0%

ries is given in Table 3. Apart from gas property data, the measuring errors of the time and geometric dimensions were quite small compared to other sources of errors. The gas properties were taken from the literature with estimated error of 1.0%. In the time interval used, the error of the time derivation of temperature was not large, estimated to be on the order of 2.0%. The biggest measurement uncertainty was acquired from determining the temperature difference between the gas and the bed material. Relative measurement error was assumed from the literature [21, 22] to be about 15-25%. With this estimation of measuring errors, the uncertainty of Reynolds number, friction factor and convective heat transfer coefficient can be calculated [23] to be about 9.5%, 16%, and 26%, respectively. These estimated uncertainties seem to be acceptable.

4. Conclusions

The present study investigated and utilized a packed bed of uniform ceramic spheres in an incinerator. Cold flow experimental determination of pressure drop and corresponding friction in the packed bed was undertaken at low gas velocities and particle Reynolds numbers. It was found that flow resistance increased with the gas superficial velocity and the bed thickness. An attempt was made to derive Ergun constants for pressure drop evolution and compare them with the existing literature. Based on the present experimental data, revised Ergun constants were 68.5 and 4.95 for the Reynolds numbers of 60-300. The observed differences and limitation of the present empirical model may be attributed to the concomitant effects of the finite characters of the packed bed and of the transition of flow regimes near the beginning of the fully developed turbulent flow regime. Temperature evolution characteristics and temperature distributions in the incinerator with and without the packed bed were also investigated. Results from thermal flow experiments indicated that an incinerator with a packed bed had faster rate of temperature rise than the void chamber. Mean heat transfer coefficients were higher inside the packed bed than at both ends of the bed. It was also found that the packed bed helped improve heat recovery from the hot flue gas. The presence of the packed bed, acting as thermal energy storage, allowed the incinerator chamber to reduce flue loss and achieve higher peak temperature, hence enhancing the conditions for destruction of pollutants.

Acknowledgments

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Nomenclature

A_o : Bed section area [m^2]
a : First Ergun coefficient
b : Second Ergun coefficient
c_P : Heat capacity [m^2/s^2-K]
D_e : Bed equivalent diameter [m]
d_P : Ball diameter [m]
f : Friction factor
h : Convective heat transfer coefficient [W/m^2-K]
L : Bed thickness [m]
m : Mass flow rate [kg/s], radial positions in a sphere [m]
r : Radial distance [m]
Re : Reynolds number
T : Gas temperature [K]
t : Time [s]
u : Superficial gas velocity [m/s]
α : Thermal conductivity [W/m-K]
ΔP : Pressure drop [Pa]
ΔQ : Heat transfer rate [kW/m^2]
ε : Porosity of the packed bed
μ : Viscosity [kg/m-s]
ρ : Density [kg/m^3]
θ : Ball temperature [K]

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