Heat transfer from a horizontal bundle of finned tubes to an air fluidized bed
by Stephen John Priebe

A thesis submitted in partial fulfillment of the requirements for the degree of DOCTOR OF PHILOSOPHY in Chemical Engineering
Montana State University

Abstract:
Heat transfer coefficients were measured from a horizontal bundle of finned tubes in an air fluidized bed. Two types of tubes were studied. Discontinuous finned tubes were used to study' fin spacing and heat flux, while spined tubes were used to study spine height, spine material, and number of spines.

Results indicate that fin spacing has little effect when the spacing between fins is greater than 10 particle diameters. At less than 10 diameters, the coefficient begins to fall rapidly up to a point where the spacing is less than two particle diameters. From this point on, the curve falls very slowly, it was also found that the heat transfer coefficient increases slightly as the heat flux is increased.

Results with the spined tubes indicate that the heat transfer coefficient decreased with increasing fin height. Coefficients obtained with copper spines were much higher than those of stainless steel.

There was little difference in the coefficient with larger number of spines. However, the increased area with more fins yields a higher rate of heat transfer. Each type of tube led to a correlation relating heat transfer coefficient to the parameters of interest.

HEAT TRANSFER FROM A HORIZONTAL BUNDLE OF FINNED TUBES TO AN AIR FLUIDIZED BED

by

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INTRODUCTION

The fluidized bed has in recent years received considerable attention in a number of industrial applications. However, to enhance use and design, more knowledge of the physical processes occurring in the fluidized bed is needed. The purpose of this thesis is to better understand heat transfer from finned tubes to a fluidized bed.

Basically, a fluidized bed consists of a column, a bed of particles supported on a distributor plate, and a source of fluid to be used as the fluidizing medium. The column may range from a few inches to several feet in diameter. The particles can consist of a wide range of materials, as can the fluidizing medium.

The term fluidization arises from the nature of the bed of particles when sufficient fluid is passed through the bed. At low fluidizing velocities, the fluid, air for example, merely passes through the bed as in a packed bed.

As the velocity is increased, a point is reached when the pressure drop across the bed is equal to the weight of solids in the bed. At this point, the bed expands slightly and acts like a highly viscous fluid. This velocity is called the minimum fluidization velocity. There is very little particle motion at minimum fluidization.

With increased velocity, particulate fluidization may occur. Particulate fluidization occurs primarily in liquid fluidized beds where the density of particles is not very much greater than the liquid. Visually, the bed resembles minimum fluidization, with only a slight increase in particle motion.

In a gas fluidized bed, gas bubbles begin forming almost immediately after the minimum fluidization velocity is exceeded. This bubbling is known as aggregative fluidization. Small bubbles form in the region of the distributor plate and rise through the bed similarly to a boiling fluid. As the bubbles move upward through the bed, they coalesce to form larger bubbles which finally burst at the surface. The size and number of bubbles depend upon the nature of the distributor plate, and bed height. With vigorous bubbling, there is considerable motion with aggregative fluidization.

The final regime of fluidization, known as slugging, occurs when slugs of fluid, occupying the entire cross section of the column, pass through the bed. Maximum particle motion occurs during slugging, but physical problems may become important. Particles may be broken down, and smaller particles may be entrained and lost.

Finally, as the velocity is increased beyond slugging, all the particles are entrained and carried out of the column. The various regimes of fluidization are shown schematically in Figure 1.

Industrially, the fluidized bed has been known for quite a long time, but only in the last 40-50 years has it achieved any major importance. At about this time, the petroleum industry discovered that the fluidized bed was very useful in the catalytic cracking unit. This eliminated the need for shutdown and complete regeneration of the catlyst.

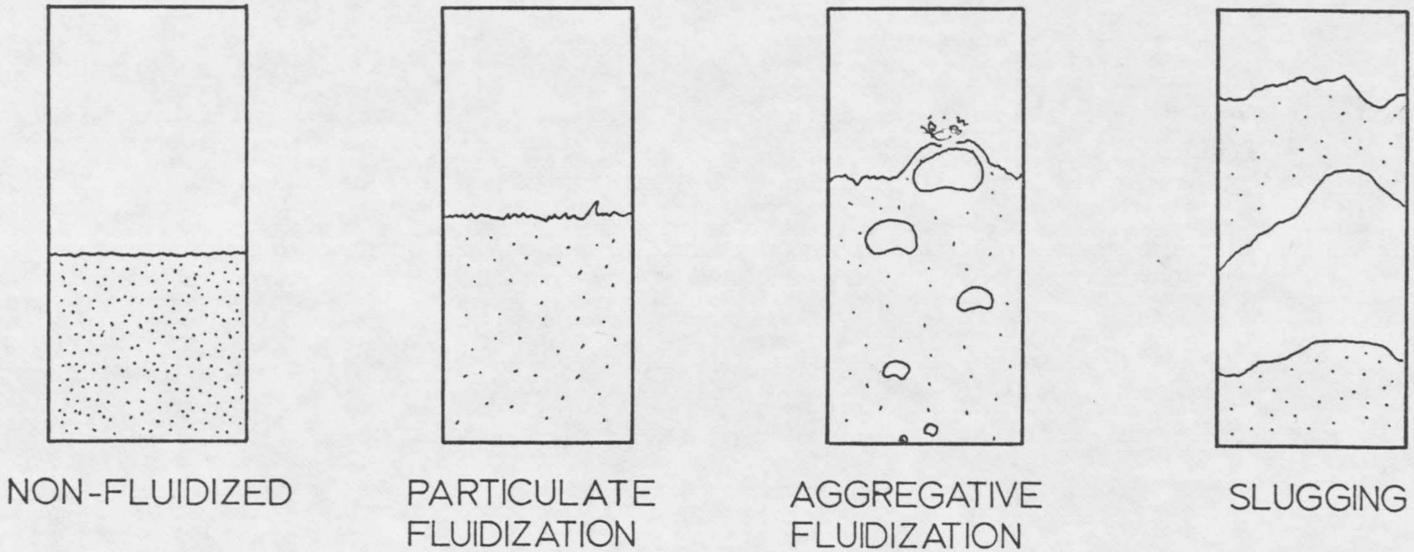

FIGURE 1. REGIMES OF FLUIDIZATION

The fluidized bed could be run with a continuous stream of particles being drawn off, regenerated, and recycled to the column.

Since the fluidized catalytic cracker became established, several other widespread applications of the fluidized bed have been made. In desalination of sea water, the sea water is sprayed into a hot bed. The water is evaporated leaving the salt to deposit on the particles.

Another application has been in the treatment of liquid nuclear wastes. Once again, solution is sprayed into a hot bed, allowing the radioactive material to deposit on the particles. The solid particles are then stored underground thus eliminating the bulk and potential leakage of liquid wastes.

In recent years, research has begun in several other areas. One area is fluidized bed combustion of coal for electrical generation plants, where limestone is used as the particulate. A prime advantage of this method is that sulfur in the coal reacts with the limestone thus eliminating SO_2 formation.

Two other recent uses of the fluidized bed have been as a shallow bed cooling tower and as a heat exchanger for geothermal energy.

As shown, there is a wide variety of possible applications for the fluidized bed. However, there are several disadvantages in its use. Listed below are some of the advantages and disadvantages of the fluidized bed.

Advantages

1. Due to extremely vigorous mixing in a fluidized bed, there are few concentration or thermal gradients. This would be important in many reactions which are strong functions of temperature or concentrations.

2. Continuous recycling of solids is much easier to implement than in a fixed bed.

3. Increased particle activity results in much higher heat transfer coefficients which would make temperature control easier and more efficient.

Disadvantages

1. Because relative fluid and particle motion is basically co-current, the driving force is not completely favorable, and the fluidized bed acts as a single stage. Multiple beds can overcome this problem, but might run into more expense.

2. Space velocity through the column is limited, because the bed fluidizes only in a relatively narrow range of velocities.

3. Particle degradation and elutriation may cause severe catalyst losses.

4. In some systems, particle agglomeration may plug a fluidized bed to the point where proper operation is not possible.

This shows that while a fluidized bed might be very useful in some cases, each application must be studied with care.

In most of the applications above, heat must be transferred to or from the fluidized bed. Originally this heat transfer was accomplished through the walls of the column. However, this severly limits the available surface area for heat transfer. To overcome this problem, either horizontal or vertical tubes are placed in the bed.

Since heat transfer coefficients have been shown to be much greater in a fluidized bed than in a similar column without the bed of particles, it is necessary to learn as much as possible about the heat transfer from immersed tubes. In many other heat transfer systems, finned surfaces have been shown to greatly increase the heat transfer. It seems probable then that finned tubes might also enhance heat transfer in a fluidized bed, thereby reducing the total number of tubes.

The purpose of this research was to study the effect of fin spacing on discontinuous finned tubes, (a type of tube similar to those used by Bartel [1]), on heat transfer coefficients. In addition, spined tubes were studied. The effects of fin height and fin material on the heat transfer coefficients were determined.

THEORY AND PREVIOUS RELATED RESEARCH

Mechanism of Fluidization

To better understand the process of heat transfer in a fluidized bed, some knowledge of the mechanism of fluidization is helpful. To simplify the process, a single gas bubble will be considered first.

A model proposed by Davidson and Harrison (3) contains three major conditions:

1. The gas bubble is spherical and particle free.
2. The particles around the bubble move as an inviscid fluid.
3. In the bulk phase, the gas flows incompressibly through the bed.

In addition to these three conditions, they also assumed that the pressure gradient far from the bubble is unaffected by the bubble passage, and that the pressure in the bubble is constant.

This model has been shown to be mostly correct with only minor changes, such as the shape of the bubble. Visual observations (14) have shown the bubble to have a concave bottom into which flows gas and some particles.

Extending this model to include multiple bubble formation and bubble interactions leads to a model proposed by Kunii and Levenspiel (7).

1. A wake of particles and gas follows each bubble in its upward travel.
2. Solids in the wake are constantly being renewed and move

upward at a velocity, U_b. At the top of the bed, the solids begin to move downward at a velocity, U_s.

3. The following relation between the bulk phase velocity, U_e, and the downward solids was proposed:

$$U_e = \frac{U_{mf}}{\varepsilon_{mf}} - U_s$$

where,

U_{mf} = velocity of fluid at minimum fluidization

ε_{mf} = void fraction at minimum fluidization

This model has been shown to do a reasonably good job describing the action of the fluidized bed. However, it must be remembered that this model applies only to aggregative fluidization.

A study was conducted by Hager and Thomson (5), on the effect of immersed tubes on bubbles traveling through a fluidized bed. Visual and x-ray techniques were used to study the motion of bubbles around bare and finned tubes.

With horizontal bare tubes, about 50 percent of the bubbles divided when passing over the tube. These bubbles quickly coalesced above the tube. The remainder of the bubbles were diverted around the tube.

With finned tubes, the action was somewhat different. Here, most bubbles were either diverted around the tube, or entered the fin structure, eventually emerging at the top of the tube. When a bubble did divide, the two halves passed around the outside of the fin structure. When this occurred, there was little particle motion between the fins.

Mechanism of Heat Transfer Between Fluidized Bed and Heat Surface

When studies showed that heat transfer coefficients were much higher in fluidized systems than in empty columns, interest was generated in developing some idea as to the reason for the increase. Several explanations have been made and most of them probably contribute to the overall picture.

A number of people, including Leva, et al. (9) and Levenspiel and Watson (10), proposed a scouring mechanism. In this model, the normal boundary layer on the heat transfer surface is partially scrubbed away by the action of the particles. When the boundary layer is thinned, the resistance is reduced thereby increasing heat transfer. Levenspiel developed a relation to describe the effective film thickness in both laminar and turbulent flow conditions. The relation describes the scouring and the growth rate.

A different model was proposed by Mickley and Fairbanks (11). According to this model, packets of particles were viewed to be the heat transfer agent. These packets were said to move from the bulk phase to the heater surface. After remaining at the surface only a short time, the packet returned to the bulk phase where the packet was dispersed.

Mickley, et al. found the heat transfer coefficient to be a function of the quiescent bed thermal conductivity as follows:

$$h = \sqrt{k_m \rho_m \, c \, S}$$

where

k_m = thermal conductivity of the packet, BTU/hr-ft-°F

ρ_m = density of packet, lb_m/ft^3

c = heat capacity of packet, BTU/lb_m-°F

S = stirring factor

The stirring factor is basically a function of the hydrodynamics of the system, but is not understood well.

About ten years ago, this model was modified by Ziegler, Koppel, and Brazelton (16). This new model was then extended by Genetti and Knudsen (4). Under this new model, a single particle rather than a packet moves from the bulk phase to near the heater surface. The fluid near the surface is assumed to be at the temperature of the heater. The particle is heated by convection from the hot fluid and then returns to the bulk phase where the heat is dissipated.

The hot gas near the surface is assumed to be a constant temperature, and the contact between the surface and the particle is neglected. The particles are packed rectangularly, and a gamma function is assumed for the residence time of the particles.

With these assumptions, the following expression for the particle Nusselt number was developed.

$$Nu_p = \frac{7.2}{\left[1 + \frac{.6\ k_{air}\bar{t}}{\rho_s C_{p_s} D_p^{\ 2}}\right]^2}$$

where,

$Nu_p = hD_p/k_{au}$ = particle Nusselt number

k_{air} = thermal conductivity of air, BTU/hr-ft-°F

$\bar{t}$ = mean residence time of particles, hr

ρ_s = particle solids density, lb_m/ft^3

C_{p_s} = heat capacity of particles, BTU/lb_m-°F

D_p = particle diameter, ft

h = heat transfer coefficient, BTU/hrft^2-°F

As can be seen from this expression, the particle Nusselt number is not a function of the thermal conductivity of the particles.

Genetti and Knudsen proposed that the constant, 7.2, be replaced by a function of the square root of the particle fraction, $(1-\varepsilon)^{0.5}$. It has been proposed that the arithmetic average of the surface temperature and the bulk temperature be used as the film temperature. This model, shown in Figure 2, will be used to correlate the results of this research.

To check the validity of this model, Ziegler and Brazelton (15) conducted an experiment with both heat and mass transfer. The system they used was an absorbent clay sphere saturated with water. This sphere was suspended in an empty column and then in a fluidized bed. The fluidized bed particles were non-absorbent, so that under the assumptions of the model, heat transfer rates should increase much more than mass transfer rates.

Results of the experiments showed an increase of 1.5 to 2 times for the mass transfer whereas heat transfer increased 10 to 20 fold. From these results, they concluded that 85-90 percent of the heat transfer can be accounted for by a particle mode of transfer.

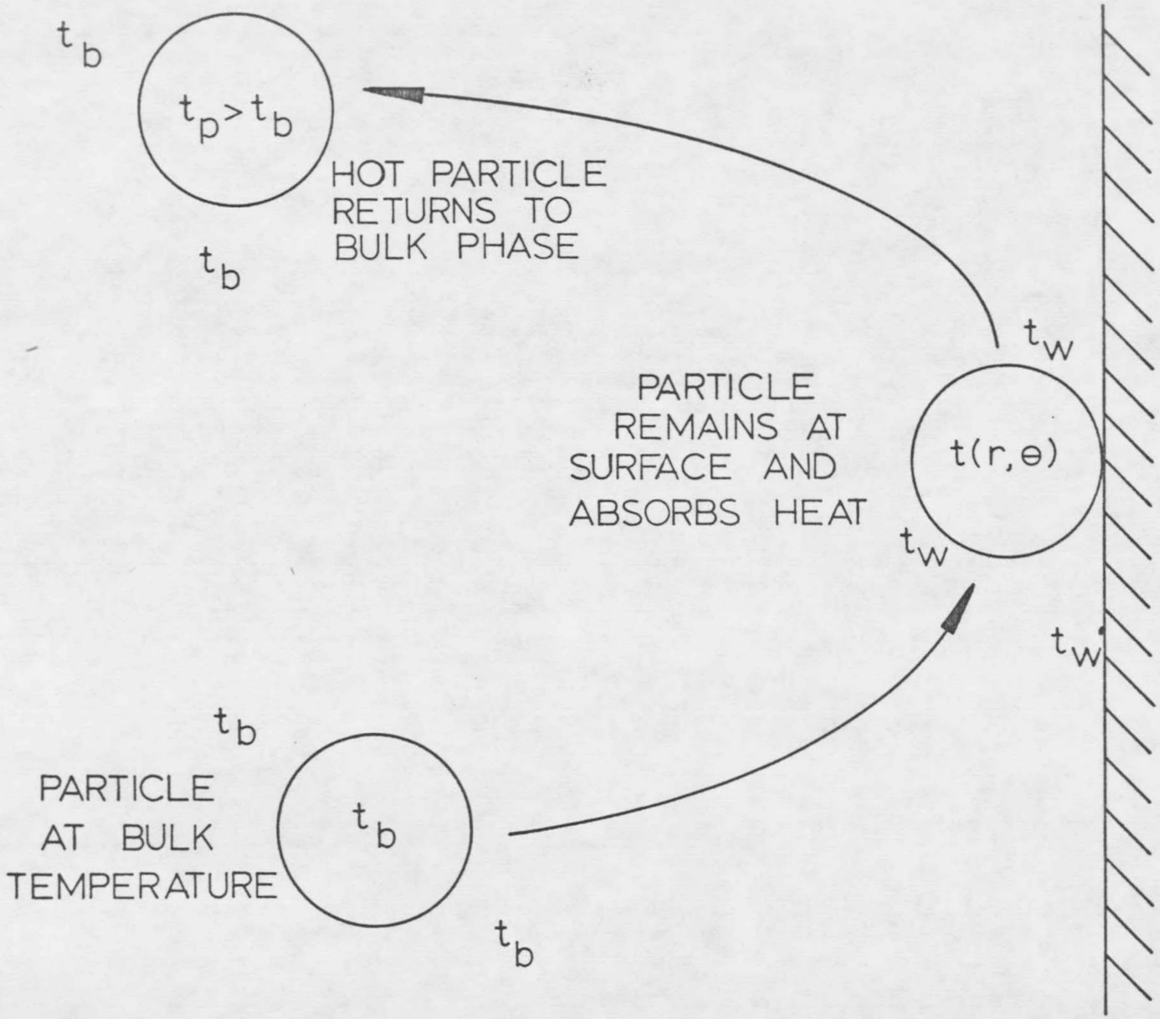

FIGURE 2. PROPOSED MECHANISM OF HEAT TRANSFER