DETERMINATION OF THE COEFFICIENT OF HEAT TRANSFER AT THE BED WALL BOUNDARY OF AN EXTERNALLY HEATED FLUIDIZED BED

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I. INTRODUCTION

A problem of special importance in reactor design has been the reduction of temperature gradients within the reaction mass. Attempts toward the solution of this problem have been made by the petroleum industry in the cracking of crude oils, where exact control of temperatures is essential. The Houdry unit developed in 1937 utilized finned tubes located throughout the fixed catalyst bed to maintain uniform temperature. Later, in the Thermofor process, reactor temperature was controlled by continuously moving beds, which, with their relatively high specific heat, carried out the excess heat. The most recent catalyst bed technique, fluidization, is a unit operation in which solid catalyst particles in violent agitation are suspended in a rising stream of gas. This process has practically eliminated bed temperature gradients.

Although fluidization received its first application in the cracking of petroleum, its principle has been applied in several other industries, including the synthesis of gasoline and alcohols from natural gas, the recovery of oil from shale and tar sands, and the gasification of coal for subsequent synthesis into liquid products. Numerous other industrial applications are now in the stage of research and development.

The practical applications of fluidization have been utilized far in advance of a fundamental study of its basic physical characteristics, particularly the effect of the various properties of the system on the local individual coefficients of heat transfer at the bed wall boundary. Limited studies have been undertaken which employed catalyst chambers of 2; 3; and 4-inch inside diameter, fluidizing media of air, carbon dioxide, and helium, and bed wall temperatures of 200 to 600 °F. The results of this work have shown the effective particle diameter and the fluidizing gas velocity to have the greatest effect on boundary coefficient of heat transfer.

It is the purpose of this investigation to evaluate the effects of superficial air velocity and temperature driving force on the coefficient of heat transfer at the bed wall of an externally heated, fluidized bed of Ottawa sand, utilizing dry air at approximately 80 °F, bed wall temperatures of 200, 400, and 600 °F, and mass superficial air velocities of 82.5 to 217.5 pounds per hour-square foot.

II. LITERATURE REVIEW

The use of finely powdered solids as catalysts was first successfully applied to large scale commercial operations in the cracking of petroleum in the early part of 1942. Its outstanding success in this field is attested by the rapid growth of fluid catalytic cracking capacity from 40,000 barrels per day at the end of 1942 to an estimated 1,000,000 barrels per day at the end of 1948. The reasons for this wide and immediate acceptance of the process by the industry lay in several unique characteristics inherent in the fluidized solids system. The most important of these characteristics is a substantial reduction in channeling as compared to fixed beds, and subsequent elimination of resulting "hot spots." Another advantage of the fluidized system is the ease with which the solids may be transported from vessel to vessel which permits the most efficient utilization of heat and the ready development of a continuous process. These characteristics make the use of fluidized solids particularly applicable to those processes where large amounts of heat must be transferred and/or large amounts of solids must be treated, as in catalyst regeneration. Accordingly, a great deal of work is in progress involving the application of the fluid-solids technique to processes such as hydrocarbon synthesis, coal gasification, and the reduction of ores.

The most efficient use of this new processing technique requires an understanding of the fundamental factors controlling the behavior of fluid-solid and gas-solid systems. This investigation pertains to one of the fundamental characteristics of a gas-solid system; the coefficient of heat transfer at the bedwall of a fluidized bed.

Principles of Fluidization

The practical applications of fluidization have so far preceded a complete scientific study of the phenomena involved that no exact or complete body of laws has yet been formulated. However, studies of a limited degree have been carried out on the basic qualitative and quantitative principles of fluidization within the last few years. The work thus far completed pertains, in a large part, to those variables that affect the fluidized state and which may be observed directly by eye. These variables include the effect of gas velocity, vessel dimensions, particle diameter, density, and shape, and mathematical correlations of these variables.

Definitions of Fluidization Terms. The field of fluidization has expanded so rapidly that the terminology used in papers by different authors has often been overlapping and confusing. Murphy, et al(71), have presented a glossary, which, it is hoped, will establish a basis for a permanent nomenclature in the field of fluidized solids techniques.

Fluidization. Patterson(76) has defined fluidization as "That unit operation in which a mass of solid particles, usually finely divided, is maintained by means of an upwardly moving liquid or gas stream, in a turbulent, dense state." Campbell(3,5,6) was comprehensive, indicating that fluidization is a new unit operation in which gases and solids can be contacted, with the solids in a turbulent pseudoliquid state, topped by a pseudogaseous phase, and by which the solids can be fed and withdrawn from a chamber by the use of the standpipe combination down- and up-leg principle.

Fixed Bed. A fixed bed(71) is a body of motionless particles supported by direct contact with each other and the retaining walls.

Moving Bed. In a moving bed(71), the particles remain in direct contact and are substantially fixed in position with respect to each other, but move with respect to the retaining wall.

Fluidized Mass. A fluidized mass(71) is a bed of solid particles that exhibits the mobility and hydrostatic pressure of a fluid.

Channeling. Channeling(71) is the establishment of flow paths within a bed of fluidized solids through which a disproportionate quantity of the fluid passes.

Slugging. Slugging(71) is a condition in which pockets or bubbles of the supporting fluid grow to the diameter of the containing vessel, and the mass of particles trapped between adjacent pockets moves upward in a piston-like fashion.

Mechanisms of Fluidization. It is necessary that the basic mechanisms that cause fluidization be understood before a study of the variables affecting fluidization be attempted. These mechanisms include the forces acting upon the fluidized bed, the manner in which a gas traverses a bed of fine particles, types of fluidized beds, and the various types of dense phase flow.

Manner in Which a Gas Traverses a Bed of Fine Particles. In this discussion, the manner in which a gas traverses a bed of fine particles is divided into three sections; gas rates lower than that necessary to cause fluidization, minimum gas rate for fluidization, and gas rates greater than the minimum for fluidization, but less than the rates required to blow the bed out of the reactor.

At very low gas velocities, the pressure drop across the bed is less than that equivalent to the weight of the bed, and the gas merely percolates through without agitation of the particles(38). Leva (28,30,33,34) and Morse(64) agree that the pressure drop increases with increasing gas velocity, and that when the pressure drop across a section of the bed equals the weight per unit of cross-sectional area, the bed will begin to expand. Beginning at this point, as shown by Leva, et al(12), the bed expands with increasing gas velocities, with the pressure drop remaining essentially constant.

Fluidization actually begins when the gas velocity is sufficient to expand the bed to a degree at which the particles become disengaged and internal particle motion is permitted(26,29). In order for the bed to become fluidized, it must reach a degree of expansion known as the "minimum fluid voidage," ϵ_{mf}. The limiting bed density, ρ_{mf}, is the density of the bed at the beginning of fluidization.

When the gas velocity is increased slightly above minimum fluidization velocity(29,64), the bed continues to expand, with intensified particle motion. The bed reaches its greatest expansion with stable configuration just as the gas velocity increases sufficiently to cause bubbling flow, or slugging. Continued increase in gas velocity causes larger portions of the gas to flow in a discontinuous phase through the bed, and ultimately reaches a point where a single dilute phase is formed. It has been shown by Matheson(38) that the velocity required for this single dilute phase is considerably greater than that predicted by Stoke's law. This deviation has been attributed to the hindered settling occurring because of the relatively small spaces between particles.

Types of Fluidized Beds. The types of fluidized beds have been classified according to density by Paterson(74) as follows: (1) high density, used in standpipes; (2) medium density, used in reactors; and (3) low density, used in lines transferring the fluid mixture. Another method often used in classification depends on the type of process used: (1) batch fluidization, in which the solids remain within the unit with practically no entrainment, and (2) continuous fluidization in which the solids flow continuously through the unit.

Types of Dense Phase Flow. Matheson(38,39) observed that channeling occurred in beds consisting of particles less than 25 microns in diameter. Vertical passages formed through which the gas passed upward with relatively little contact with the bulk of the solids in the bed. In the higher range of velocities, it was noted that a type of flow occurred in which the channels were broken and the bed moved about in large cohesive masses that continually broke and reformed.

Aggregative Fluidization. Morse(64) defined aggregative fluidization with a gas as that type in which the bed contains channels and large clumps of only slightly separated particles, that are continuously falling and being torn apart by high velocity gas. Wilhelm and Kwauk(85) stated it in another manner, saying that aggregative fluidization in a solid-air system more closely resembled a liquid than a gas. They also noted a dispersed "vapor phase," in evidence above the main aggregated "liquid phase." Leva(29) noted that pressure fluctuations for this type of fluidization were very low. Matheson(39) termed as aggregative fluidization the conditions existing when the gas percolates through the bed, and at higher gas velocities, the condition of bubble formation which causes turbulence, and carries the excess gas through the bed.

Particulate Fluidization. Morse(64) observed a type of fluidization in a liquid-solid system in which there was very little circulation of particles, known as particulate fluidization. Other investigators(85) noted that in this type of fluidization, the solid particles separated in the manner of a gas system, the mean free path increasing with fluid velocity.

Important Characteristics of a Normally Fluidized Bed. Included in the discussion below are the characteristics and description of a desirable, normally fluidized bed. These include a general description, quality, density, pressure drop, relation of pressure drop to velocity of fluid, viscosity, contact time, mixing, entrainment, heat transfer and temperature control, and ease of solids transfer.

General Description. Under normal conditions of fluidization, as noted by Matheson(39), the bed consisted of a gas phase rising in the form of bubbles through a fluid system composed of the solid particles, fluidized by a small fraction of the total amount of gas passing up through the bed. Campbell(4) described the system as a pseudoliquid phase in which the solid particles are completely air-borne, remaining relatively close to each other in a concentrated mixture, flowing about in eddying fashion. Murphree(67) and Patterson(73) reported the solid-gas mixture to be in extremely turbulent motion, resembling a boiling liquid, or as expressed by Campbell(4), the mass resembled water with air bubbling through it. As brought out by Ergun and Orning(12) and Matheson(38), the system contained two phases--a liquid-like dense phase and a relatively dilute suspension in which solids are totally entrained. Campbell(4) observed that a dilute phase of solids suspended in the gas usually existed immediately above the dense phase.

Quality. The quality of a fluidized bed, as defined by Morse(63), is the uniformity of particle dispersion and of gas velocity throughout the bed; the more uniform, the higher the quality. He observed that water-fluidized beds of sand approach perfect quality, whereas, a bed which slugs or channels is of poorest quality. It is his belief that the bed of intermediate quality, with small and uniform gas pockets is optimum as to mixing and consequently lends itself best to temperature control.

Morse(64) has pointed out that the degree of segregation is the dynamic balance between the tendency for a fluid to separate from solid particles and the tendency to remix. If the segregation rate is high, mostly aggregative fluidization will occur. If the segregation rate is low, the fluidization will be mainly of the particulate, desirable type.

Density. Matheson(40) pointed out that the density of the dense phase fluid-solid system is of interest from both fundamental and engineering considerations. It was also noted that maximum bed density was obtained when the pressure drop across the bed was a maximum with the bed aerated and few bubbles rising. The maximum bed density, therefore, represents the density of the continuous phase of the bed and is analogous to the density of a true liquid. Lewis(36) observed that at constant gas velocity there was considerable variation in bed height and consequently variation in average bed density or average fraction voids.

Pressure Drop. According to Morse(65), the pressure drop through the fluidized bed is less than that through the corresponding fixed bed at the same gas velocity. The foregoing notations have shown that knowledge of the pressure drop within a fluidized system is an important criterion in determining state or type of fluidization.

Relation of Pressure Drop to Velocity of the Fluid. Zenz(86,87) has plotted a schematic summary of particle-gas flow characteristics, where the pressure drop per unit-length-of-bed versus superficial fluid velocity has been plotted on a log-log scale. The curve representing the fraction voids in a densely settled fixed bed, Figure 1, designated as ϵ_s, was included as a comparison to the parallel curve representing the voidage in the loosest possible fixed bed configuration, designated as ϵ_{mf}. He noted that at velocities through ϵ_{mf}, where pressure drop equals weight of material per unit cross section, the bed becomes fluidized. He observed that as the bed expands, the voidage increases, thus reducing the pressure drop. The fluidization curve intersects the pressure drop line for the empty pipe at u_t, the terminal velocity of the particles. For a small change in velocity at high voidages, there is a large change in pressure.

Viscosity. Matheson(41,42,43) has shown, by comparison of fluidized solid systems with gas flow through liquids, that fluid-solid systems possess a property similar to that of viscosity in liquids.

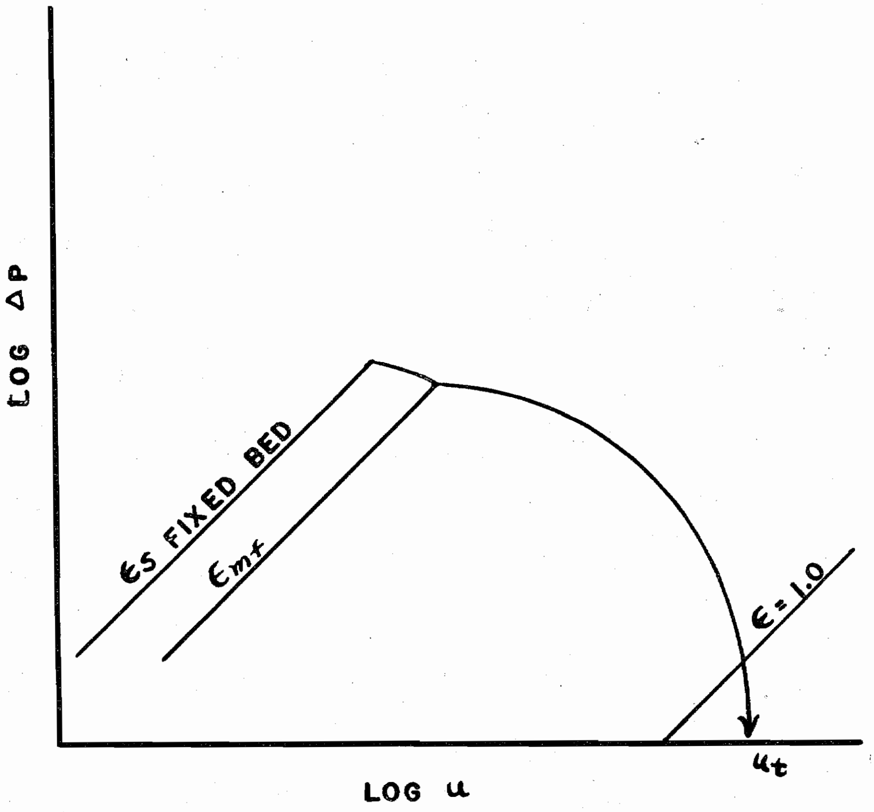


FIGURE 1. RELATION OF PRESSURE DROP TO FLUID VELOCITY IN A FLUIDIZED BED.

ZENZ, F.A.: TWO-PHASE FLUID-SOLID FLOW, IND. ENG. CHEM., <u>41</u>, 2804 (1949).

A normally fluidized bed was compared to upward flow of gas through a low viscosity liquid and slug flow was compared to gas flow through a viscous liquid. He defined the viscosity of the bed as "the Stormer net weight (grams) determined with a superficial air velocity through the bed just slightly higher than that at which the Stormer torque becomes independent of air velocity." He observed that the tendency for a solid to slug increased with increased viscosity of the system.

Contact Time. It was reported by Campbell(7) that the time of contact of solids in the dense fluid bed sometimes affects the degree of completion of the desired change. The time can be varied by changing the rate of solids input or by changing gas velocities within the limits of fluidization. However, it was brought out by Patterson(74,76) that gas flow rates are relatively low, usually between 1.5 and 2.0 feet per second; therefore, the process is inapplicable where short contact time is required.

Mixing. According to Patterson(73), mixing within the fluidized bed is complete and virtually instantaneous because of violent agitation. The property of instantaneous mixing has been shown by

Patterson and Campbell(6) by the fact that samples drawn from any section of the bed show little variation in particle distribution or residence time. As explained by Morse(62), the main advantage of this mixing action in fluidized reactors is the great reduction in the duration of hot spots within the reaction mass.

Particle Entrainment. Campbell(4) has shown that air velocity may be considerably greater than would be predicted by Stoke's law without excessive entrainment of particles from the system.

Heat Transfer and Temperature Control. Mickley and Trilling(55) have found that the fundamental advantage of a fluidized system is its tendency to maintain a uniform temperature distribution throughout the bed. Mickley and Trilling(55) and Murphree(68) concluded that it is the rapid circulation of solids which tend to hold a uniform temperature throughout the mass. Patterson(74) explained that the essentially instantaneous rate of heat transfer results from the very high surface to mass ratio as well as to the agitation of the solids. Campbell(7) deduced that heat flow by direct contacting is inherent within the bed. According to Murphree(68), the temperature variation from bottom to top in large reactor vessels is less than 5 °F. According to Campbell(6), the entering gas comes almost at once to bed temperature. Murphree(70) and Patterson(74) have pointed out that, as a result of this very uniform temperature distribution within the bed, temperature control is facilitated.

Matheson(38) has reasoned that the extreme turbulence of the solid-gas suspension allows excellent heat transfer to and from the system. Patterson(74) found that heat transfer rates between the fluidized bed and the vessel walls are about the same or better than are obtained with boiling liquids. Mickley and Trilling(59), in their studies of heat transfer in an externally heated fluidized bed, obtained coefficients of heat transfer ranging from 3 to 70 times as great as those obtained with the same gas flow rates in the absence of particles. Breckon(1) found the heat transfer coefficient for a fluidized bed of sova beads to be 13.7 times as great as for an empty bed at the same mass air flow rates. On the other hand, for low gas velocities, Parent(72) obtained coefficients of heat transfer 3 per cent lower for fluidized beds than for empty beds. However, at higher velocities, he obtained heat transfer coefficients which were about 5 per cent greater for fluidized solids than for empty beds.

According to Patterson(74), for endothermic reactions, heat may be supplied by preheating the solid to well above the reaction temperature with no danger of overheating the reacting gas. As pointed out by Patterson and by Ergun and Orning(11), for exothermic reactions, it is possible to operate at higher average temperatures and consequently higher reaction rates without excessive decomposition because of hot spots.

Campbell(6) and Murphree(70) pointed out that the solid mass within the vessel imparts heat capacity to the system, thus guarding against rapid temperature fluctuations. Mickley and Trilling(55) noted that this heat capacity is an important factor in close temperature control of the system.

Mickley and Trilling have summarized the methods for temperature control in a fluidized reactor as: (1) control of temperature and flow rate of entering fluid, (2) control of temperature and cycling rate of solid particles entering the system, (3) transfer of heat to heat exchanger surfaces within the bed, and (4) transfer of heat to the walls of the vessel.

Ease of Solids Transfer. According to Hariu and Molstad(13), Leva(28), and Matheson(38), the ease with which fluidized solids may be transported from vessel to vessel is an inherent advantage of the system which facilitates application to a continuous type process. Matheson has reasoned that such ease of transport permits the most efficient utilization of heat.

Variables Affecting the State of Fluidization. It is absolutely essential that the effects of variable conditions upon the fluidized state be understood in order that fluidized beds may be utilized as fully and efficiently as possible. These variables are discussed in three groups: characteristics of the solid, the gas, and the containing vessel.

Characteristics of the Fluidized Material. The important characteristics of the solid being fluidized are diameter, density, shape, surface tension, and electrostatic charge. The most important of these properties is particle diameter. Patterson(75) showed that as particle diameter increases, an increase in gas flow rate is required to maintain a given fluidized bed density. Matheson(40) noted that for a bed of solids of narrow size range, the Stormer viscosity increased with increasing particle diameter, and that the rate of bubble growth increased with diameter. Leva(27) observed that fine materials have the greatest tendency to channel, but on the other hand, Morse(66) found that an increase in particle size caused segregation.

It has been shown by Patterson(75) that for a given flow rate and particle size, bed densities are essentially constant, regardless of the absolute density of the solids. Matheson(41), however, found that maximum bed density increased with greater particle density. Leva(30) showed that the velocity of gas necessary for fluidization is proportional to the density of the solid. Morse(66) stated that the tendency of the particles to segregate increased as the particle density increased.

According to Patterson(75), little is known about the effect of particle shape on the state of fluidization. It was reasoned, however, that from theoretical considerations, smooth spherical particles would require higher gas velocities for fluidization than would rough, irregular particles of the same size distribution and density. Seidel(77), studying particles of dissimilar shape, but with like bulk densities, however, concluded that there was no definite relationship between particle shape and the pressure drop over the fluidized bed.

Patterson also found that electrostatic charge is a very minor factor in the fluidized state, except that for very fine particles, an electrostatic charge tends to facilitate fluidization, as like charged particles tend to repel each other, thus requiring lower gas velocities.

Characteristics of the Fluidizing Medium. The effects of velocity, density, and viscosity of the medium on the fluidized state have been found to be the most important factors affecting the design of fluidization equipment.

Maximum bed density was found at minimum gas velocity for fluidization by all investigators. Patterson(75) and Matheson(40) observed that bed density decreased with increasing gas velocity. Part of this decreased density, however, is due to the presence of gas bubbles in the bed. Patterson(73) noted that pressure drop through a fluidized bed is nearly independent of gas velocity. Leva(25) observed that channeling was more pronounced at low flow rates. Violent agitation destroyed channels at higher velocities.

It is generally agreed(35,41,75) that the effect of gas density on the fluidized state is negligible, because the gas density is slight in comparison with the solid density.

Matheson(41) observed that maximum bed density increased with decrease in gas viscosity, while Lewis(35) found that the velocity of the fluidizing gas was inversely proportional to the gas viscosity.

Characteristics of the Vessel. The only criteria of vessel design yet established is that the greater the diameter-to-length ratio, the less the tendency for slugging to occur, because the bubbles which form are given more time to expand and combine in the longer vessels.

Mathematical Relationships. As explained by Campbell(10), many of the empirical relations that have been developed are reasonably satisfactory for certain commercial uses, but are unsafe to extrapolate for universal application. Some of the methods of calculating pressure drop, gas velocity, and bed density are discussed.

Pressure Drop. It has been shown by Patterson (74,76) that, within practical limits, the pressure drop through the fluidized bed is approximately equal to the static head of the bed:

$$\Delta P = \frac{4\,W}{\pi\,D^2} = \rho_B L \qquad (1)$$

where

ΔP = pressure drop through bed, lb/sq ft

W = weight of solids in bed, lb

D = diameter of bed, ft

ρ_B = density of bed, lb/cu ft

L = depth of bed, ft.

Lewis(35) likewise concluded that the pressure drop is essentially equal to the weight of the solids, only for beds with small ratios of length to diameter. Beds with large ratios of length to diameter are likely to be in considerable error. He found that Equation 1 was satisfactory for particles of small diameter, 0.0004 to 0.0006-inch, with good fluidization characteristics and no slugging; while particles of large diameter, 0.01 to 0.022-inch, which gave slugging conditions, showed larger pressure drops than would be indicated by the static head. Only at the lowest fluidization velocities, under conditions of slug flow, did observed pressure drops approximate that represented by Equation 1.

According to Hariu and Molstad(14), the total pressure drop can be considered equal to the sum of two pressure drops; that resulting from the gas flowing through the empty tube, and a solids pressure drop consisting of the static head and a friction loss resulting from particle-to-particle and particle-to-wall contact.

Wilhelm and Kwauk(84) derived the following equation, based upon the concept that fluidization takes place when the weight gradient of the solid bed is equal to the pressure gradient through the bed:

$$\frac{\Delta P}{L} = (1 - \epsilon)\,(\rho_s - \rho_P) \qquad (2)$$

where

$\frac{\Delta P}{L}$ = pressure gradient, lb/sq ft-ft

$1 - \epsilon$ = fraction solid at height L, ft

ρ_s = density of particles, lb/cu ft

ρ_P = density of fluid, lb/cu ft.

Leva(32) has developed a formula for pressure drop through the expanded bed which is essentially the same as Equation 2 above:

$$\Delta P = \frac{V_T}{A_T} (1 - \epsilon) (\rho_s - \rho_F) \qquad (3)$$

where

ΔP = pressure drop through bed, lb/sq ft

V_T = volume of dumped bed, cu ft

A_T = cross-sectional area of bed, sq ft

ϵ = fraction voidage in dumped bed

ρ_s = density of particles, lb/cu ft

ρ_P = density of fluid, lb/cu ft.

Gas Velocity. Leva(32) has derived the following equation for the mass flow of gas required for fluidization:

$$G = \frac{0.005\, D_p^2\, \epsilon^3\, (\rho_s - \rho_F)\, \rho_F\, g}{\lambda^2\, (1 - \epsilon)\, \mu} \qquad (4)$$

where

G = mass superficial gas velocity, lb/hr-sq ft

D_p = average effective particle diameter, ft

ϵ = fraction voids in dumped bed

ρ_s = density of particles, lb/cu ft

ρ_F = density of fluid, lb/cu ft

g = acceleration due to gravity, 4.18×10^8 ft/hr/hr

λ = particle-shape factor, dimensionless

μ = viscosity of fluid, lb/hr-ft.

Leva(27) stated that the mass velocity required for beginning of fluidization may be predicted by substituting the fraction voidage of bed at minimum fluidization, ϵ_{mf}, for fraction voidage in dumped bed, ϵ, in Equation 4.

The general expression, given by Leva(25), for gas flow rate at the beginning of fluidization is:

$$G_{mf}^2 = \frac{D_p \, g \, \rho_F \, (\rho_s - \rho_F) \, \epsilon_{mf}^3}{2 \, f \, \lambda^{3-n} \, (1 - \epsilon_{mf})^{2-n}} \qquad (5)$$

where

G_{mf} = minimum mass superficial fluid velocity required for fluidization, lb/hr-sq ft

D_p = average effective particle diameter, ft

g = acceleration due to gravity, 4.18×10^8 ft/hr/hr

ρ_F = density of fluid, lb/cu ft

ρ_s = density of particles, lb/cu ft

ϵ_{mf} = fraction voidage of bed at minimum fluidization

f = modified friction factor, dimensionless

λ = particle shape factor, dimensionless

n = state-of-flow-factor, dimensionless.

Bed Density. Ergun and Orning(12) found that the lowest bulk density of any type of particles could be obtained by fluidizing the particles with a gas and then slowly reducing gas flow. Knowledge of the lowest bulk density is sufficient to estimate flow rates corresponding to the incidence of two-phase fluidization.

Matheson(40) found that for irregular particles, ranging in size from 28 to 456 microns diameter and with densities ranging from 72 to 490 pounds per cubic foot, the maximum bed density, $_{MB}$, may be represented by the following equation:

$$\log D_p = 2.81 \frac{\rho_{MB}}{\rho_s} + 1 \qquad (6)$$

where

D_p = average effective particle diameter, ft

ρ_{MB} = maximum bed density, lb/cu ft

ρ_s = density of particles, lb/cu ft.

He noted that smaller microspheres had maximum bed densities considerably greater than those predicted by Equation 6, but established no quantitative relationships.

Utilization of Fluidization Processes

The advent of the fluidized technique has opened up a whole new field within the chemical industry. Considering the petroleum industry as an example, this field of application has increased 25 fold in eight years. In 1942, the first commercial scale fluid catalytic cracking unit was put on stream, and the total fluid cracking capacity at the end of that year was 40,000 barrels per day. At the end of 1949, the extimated total capacity of fluid cracking units was in excess of 1,000,000 barrels per day.

Commercial Applications. By far the greatest advances in commercial application of fluidization have been made in the petroleum industry. However, there are many other practical applications of the fluidized bed now in operation. These include hydrocarbon synthesis, coal gasification, ore reduction, drying of dolomite, heating of air and steam to high temperatures, and alkylation.

Future Applications. The most promising field for future application of fluidization appears to be the carrying out of Fischer-Tropsch reactions, previously too complex and costly for commercial scale processing. Other fields in which fluidization seems to offer many advantages over present methods include: (1) devolatilization; (2) calcination; (3) mixing of solids; (4) drying of solids; (5) seperation and purification of gases; and (6) recovery of vapors from gases.

<u>Limitations</u>. Patterson(76) listed the present limitations of fluidization as: (1) short gas-to-solid contact time; (2) difficulty in controlling average retention time of solids within the bed because of violent agitation; and (3) limited flexibility of operation because of necessity of keeping gas flow within narrow limits.

<u>Advantages of Fluidized Bed Reactors</u>. The main advantage of a fluidized bed reactor over any other type of reaction vessel is the uniform temperature distribution throughout the fluid bed. Channeling is greatly reduced by the extremem turbulence of the fluidized solids, thus eliminating hot spots. Other advantages over fixed bed reactors include: (1) better heat transfer within the reactor; (2) lower mechanical energy requirements; and (3) simple equipment construction.

General Theory of Heat Transmission

The problem of heat transmission is encountered in almost every industry, and in an almost infinite variety of special cases and applications. However, the principles underlying the problem are everywhere the same, and it is the intent of the following topics to present the basic fundamentals of heat transfer, so that the next section, on its applications to fluidization, may be more thoroughly understood. The basic types of heat transmission are defined, the mechanism of heat transfer between fluids and solids is included, and the film concept and development of equations for the local overall and local individual coefficients of heat transfer are covered.

Modes of Heat Transmission. Heat may flow by three distinct mechanisms, conduction, convection, and radiation.

Conduction. Conduction, according to McAdams (44,52), is "The transfer of heat from one part of a body to another part of the same body, or from one body to another in physical contact with it, without appreciable displacement of the particles of the body."

Mathematically expressed, Fourier's law for the conduction is as follows:

$$\frac{dQ}{d\theta} = -kA\frac{dt}{dx} \qquad (7)$$

where

$dQ/d\theta$ = rate of heat flow, Btu/hr

k = proportionality factor or the thermal conductivity, Btu/hr-sq ft-°F/ft

A = area of section taken at right angles to the direction of heat flow, sq ft

$-dt/dx$ = rate of change of temperature, t, with respect to length of path, x, °F/ft.

The differential form given above is general for unidirectional conduction and may be applied to cases in which the temperature gradient $-dt/dx$ varies with time as well as with the location of the point considered. In every case of heat flow by conduction, a temperature gradient must exist.

The process of heat flow in a case in which temperature varies with both time and position is called heat conduction in the unsteady state(45).

As contrasted with heat conduction in the complicated unsteady state, heat conduction in the steady state refers to those cases in which the temperature at any given point in the system is independent of time.

McAdams(45) gives the basic equation for thermal conduction in the steady state as follows:

$$q = -kA\frac{dt}{dx} \qquad (8)$$

where

q = steady rate of heat flow, Btu/hr

k = thermal conductivity at temperature, t, Btu/hr-sq ft-°F/ft

A = area of section taken at right angles to the direction of heat flow, sq ft

$-dt/dx$ = temperature gradient, °F/ft.

Convection. Convection, as defined by McAdams(44,52), is "The transfer of heat from one point to another within a fluid, gas or liquid, by the mixing of one portion of the fluid with another. The motion of the fluid may be entirely the result of differences of density resulting from the temperature differences, as in natural convection; or the motion may be produced by mechanical means, as in forced convection."

Radiation. Thermal radiation, as expressed by Hottel(15,16,17), consists of that radiant energy emitted by a substance which is entirely dependent on the temperature level of the substance. Conduction and convection are controlled by temperature difference alone; in contrast, radiation depends on temperature difference and on the absolute temperatures of the emitting and receiving substances. Thus, while conduction and convection may be the limiting factors of heat transfer at lower temperatures, radiation becomes the controlling factor at some higher temperature.

Heat Transfer Between Fluids and Solids. McAdams(50) indicated that many types of industrial heat-transfer equipment involve heat transfer between a surface and a fluid without evaporation or condensation. He listed such equipment as fire-tube boilers, superheaters, economizers, preheaters, and condensers in the power-plant field. Other industries include heat transfer to air, flue gases, water, steam, products ranging from fixed hydrocarbon gases to the very viscous liquids such as lubricating oils and asphalts in the petroleum industry, molten metals, slags, broken solids, acids, and organic solvents. The following discussion includes the film concept of resistance to heat flow at the boundary of a fluid and solid with especial reference to turbulent motion and the development of basic equations for the local over-all and local individual coefficients of heat transfer.

Film Concept. As explained by Walker(80), a large resistance to heat flow is found at the boundary of a fluid and a solid because of a thin film of fluid at the interface through which heat can be transmitted by conduction only. He also noted that the capacity of heat transfer apparatus is frequently limited by the thermal resistance of the fluid films rather than by the retaining wall of the vessel, hence individual film resistances become controlling factors in equipment sizing. According to McAdams(51,53), the mechanism of heat transfer by conduction and convection is complicated for the case of turbulent flow. He pointed out that the velocity gradient across a stream apparently involves, in addition to the laminar film, a buffer layer between the film and the turbulent core. Walker(80) brought out that the most effective means of reducing boundary resistance is to reduce film thickness, usually by increasing fluid velocity. As pointed out by McAdams(50), certain factors, of which average velocity of the fluid past the heat-transfer surface is the most notable, will generally have a greater effect on resistance for the case of flow in the turbulent range than for streamline flow.

Local Over-all Coefficient of Heat Transfer. According to McAdams(46) and Walker(81), in most instances of industrial heat transfer, the flow of heat is from one fluid through a solid wall to a second fluid.

The equation given by Kirkbride(19) for the local over-all coefficient of heat transfer is based on the general law for the rate of flow of energy:

$$\frac{dQ}{d\theta} = C\,\frac{\Delta F}{R'} \qquad (9)$$

where

Q = quantity of mass or energy transferred

θ = time corresponding to transfer of amount Q

C = a constant

ΔF = difference in driving forces

R' = resistance to transfer.

According to Kirkbride(18), in chemical engineering work, transfer is usually carried out through a

X

known area which is perpendicular to the direction of energy flow. The rate of energy transfer is directly proportional to this area:

$$\frac{dQ}{d\theta} = C\,\frac{\Delta F\ A}{R} \qquad (10)$$

where

A = area perpendicular to direction of transfer

R = resistance to transfer per unit area.

McAdams(47,54), Kirkbride(19), and Walker(82) have shown the basic relation for heat transfer between two fluids separated by a retaining wall to be represented by the following equation:

$$dq = U\ dA\ \Delta t \qquad (11)$$

where dq

dq = differential rate of heat transfer, Btu/hr

U = local over-all coefficient of heat transfer, Btu/hr-sq ft-°F

dA = differential area through which heat is transferred at right angles to direction of heat flow, sq ft

Δt = over-all difference in temperature between the warmer and colder fluids, °F.

It must be remembered, however, that the use of the over-all coefficient of heat transfer, U, is allowable and very convenient, but it is a very complex function of the operating conditions. As the individual coefficient depends on fewer variables than does the over-all coefficient, the correlation of data would be simplified by studying individual rather than over-all coefficients. The resistance of the intervening solids may be determined by use of their known physical properties.

Local Individual Coefficient of Heat Transfer. According to McAdams(48) and Walker(81), the heat flow rate is proportional to the difference in temperature between the solid wall and the fluid, and to the heat transfer surface:

$$dq = h'' \, dA'' \Delta t'' \qquad (12)$$

where

dq = differential rate of heat transfer, Btu/hr

h'' = local individual coefficient of heat transfer, Btu/hr-sq ft-°F

dA'' = differential, surface area of wall directly beneath the fluid film, sq ft

$\Delta t''$ = difference in temperature between the solid wall and the bulk temperature of the fluid, °F

McAdams(48) and Walker(80) have pointed out that the individual coefficient of heat transfer is not constant even for a given fluid, but is a complex function of such variables as the physical properties of the fluid, nature and shape of the solid surface, and the fluid velocity past the solid boundary.

Heat Transfer in Fluidized Beds

Less basic research has been done on heat transfer in the fluidized state than on any other phase of fluidization. Mickley and Trilling(55) stated that when the factors controlling the rate of heat transfer to and from a fluidized system are determined and properly correlated, a more efficient utilization of the temperature control possibilities, and hence a wider application of the fluidized technique will be possible. The following discussion will cover these topics; (1) bed temperature distribution; (2) mechanism of heat transfer; (3) variables affecting heat transfer; and (4) correlation of variables.

Fluidized Bed Temperature Distribution. Mickley and Trilling(58) found the temperature variation in fluidized beds to be very slight. Their data showed that the maximum longitudinal temperature difference in a 4-inch diameter, 25-inch tube, externally heated to 500 °F, was 13 °F. The maximum horizontal temperature gradient in the same test was 6 °F, measured from the tube axis to $\frac{1}{4}$-inch from the heating wall. Other tests under varying conditions gave smaller temperature variations.

Mechanism of Heat Transfer in the Fluidized State. Mickley and Trilling(59,60) concluded, "The elimination of temperature gradients in the bulk portion of the stream by virtue of transport of heat by the particles localizes the temperature gradient to a thin layer near the heated wall. The effective thickness of this layer is probably reduced by the motion of the particles. The temperature difference across this layer is essentially the difference between the bulk stream temperature and the wall temperature. Heat flows through this layer by means of conduction and, because of the disturbing influence of the solid particles, probably by convection. In addition, it is reasonable to assume that heat is transferred across the layer also by the movement of the heat-carrying solid particles."

Variables Affecting Heat Transfer in the Fluidized State. Several investigators(20,55) have listed the following variables as being those which affect heat transfer in the fluidized state: (1) properties of the materials, such as thermal conductivity, density, and viscosity of the fluidizing gas, density, specific heat, and thermal conductivity of the fluidized solids; (2) operating conditions, such as size, size distribution, and shape of the solid particles, concentration of the solids in the bed, superficial velocity of the gas, feed or recycle rate of the solids, and temperature level and magnitude of the temperature driving forces; (3) equipment design.

Solids Concentration. Probably the greatest influence on heat transfer in a fluidized bed is the concentration of solids present. Mickley and Trilling(58,59) found that the logarithm of the coefficient of heat transfer increased proportionately with the logarithm of the solids concentration, up to the transition region in the curve of solids concentration versus mass air flow rate. It should be noted, however, that the gas flow rate is a function of the solids concentration, and is therefore indirectly responsible for the change in heat transfer properties.

Particle Density. Leva(20) observed that particle density exerted no effect on the coefficient of heat transfer when particles whose densities varied by as much as 100 per cent were used.

Particle Diameter. All data so far available show conclusively that the heat transfer coefficient increases with decrease in particle diameter, other factors being equal. However, Leva(21) has shown that heat transfer depends on fluidization efficiency, which also increases with decreasing particle diameter.

Vessel Size. Leva(22,23) observed that bed height and diameter had insignificant effects on the coefficient of heat transfer in 2- and 4-inch diameter columns. It was his belief that in much larger tubes, where the horizontal temperature gradients are likely to become much steeper, the diameter would have more effect on heat transfer.

Correlation of Variables Affecting Heat Transfer Coefficients in Fluidized Systems. Mickley and Trilling(61) found, as a result of their work on heat transfer in externally heated fluidized beds, that when the coefficient of heat transfer at the bed-wall boundary, h'', was plotted against the $\rho_B G/D_p^3$, on a logarithmic scale, a straight line could be drawn which would fit all of their data within 25 per cent. The equation is given below:

$$h'' = 0.0118 \frac{\rho_B G^{0.263}}{D_p^3} \qquad (13)$$

where

h'' = local individual coefficient of heat transfer at bed-wall boundary, Btu/hr-sq ft-°F

ρ_B = density of bed, lb/cu ft

G = mass superficial gas velocity, lb/hr-sq ft

D_p = average effective particle diameter, ft.

They believed that this preliminary correlation does not include all the important variables, based on the postulated mechanism of heat transfer in the fluidized state.

Leva(25) obtained a line of slope equal to 1 when log h''/k versus log $G E_\phi/\mu$ was plotted. The following equation was obtained:

$$h'' = 0.64 \frac{k}{\mu} G E_\phi \qquad (14)$$

where

h'' = local individual coefficient of heat transfer at bed-wall boundary, Btu/hr-sq ft-°F

k = thermal conductivity of the fluid, Btu/hr-sq ft-°F/ft

μ = viscosity of fluid, lb/hr-ft

E_ϕ = efficiency of fluidization, no dimensions.

For gases where $C_p \mu/k$ equals 0.74, Equation 14 becomes:

$$h'' = 0.86\, C_p\, G\, E_\phi \qquad (15)$$

where

h'' = local individual coefficient of heat transfer at bed-wall boundary, Btu/hr sq-ft-°F

C_p = specific heat of gas at constant pressure, Btu/lb-°F

G = mass superficial gas velocity, lb/hr-sq ft

E_ϕ = efficiency of fluidization, dimensionless.

Leva found that Equation 15 represented all data with an average deviation of ± 22 per cent.

III. EXPERIMENTAL

The experimental procedure in the investigation of heat transfer to a fluidized bed includes the purpose of investigation, the plan of experimentation, the materials and apparatus used, the method of procedure, the data and results, and the sample calculations.

Purpose of Investigation

It was the purpose of this investigation to evaluate the effects of superficial air velocity and temperature driving force on the coefficient of heat transfer at the bed wall of an externally heated, fluidized bed of Ottawa sand, utilizing dry air at approximately 80 °F, bed wall temperatures of 200, 400, and 600 °F, and mass superficial air velocities of 82.5 to 217.5 pounds per hour-square foot.

Plan of Experimentation

The plan of experimentation followed in this investigation consisted of a survey of the literature, modification of the Breckon(1) fluidization unit, preliminary tests on the Seidel(77) catalyst test apparatus, experimental procedure, and evaluation of the data obtained.

Literature Survey. A literature search was made to obtain a general introduction to the field of fluidization and to determine the applications of the fluid-solids technique. A review was made of the principles of heat transmission applicable to the study of boundary coefficients, and of the specific field of heat transfer in the fluidized state.

Determination of Fluidization Properties of Bed Material. A catalyst test apparatus, designed and constructed by Seidel(77), was used to determine the fluidization properties of the Ottawa sand employed as the bed material. The primary purpose of the tests conducted on the Seidel unit was to determine the range of mass superficial air velocities to be used in the investigation on heat transfer to a fluidized bed.

Modification of the Breckon Fluidization Unit. The apparatus used in this investigation was essentially the unit designed and constructed by Breckon(1). The only modification made in this unit is described below.

The gas exit of the fluidization column was changed from a horizontal position to an upward inclined position (Figure 2), making an angle of 45° with the vertical axis of the column. The change was made to reduce carry-over of the dispersed solids, and their consequent entrapment in the exit line. The inclined exhaust pipe permitted the solid particles to fall back into the fluidization chamber.

Experimental Procedure. The fluidized bed consisted of Ottawa sand as the solid and air as the fluidizing medium. Tests were made at bed wall temperatures of 200, 400, and 600 °F, and at four mass superficial air velocities ranging from 82.5 to 217.5 pounds per hour-square foot at each wall temperature.

In making a test under specific conditions, the air flow through the fluidized bed was adjusted to the desired rate as indicated by the air inlet orifice manometer. The desired bed wall temperature was maintained by means of the "capacitrol" temperature controller. When it was apparent that steady state conditions had been reached, data were recorded of bed and bed wall temperatures, pressure drop across the fluidized bed, energy input to the bed heating unit, and temperature and flow rate of the entering air.

Evaluation of Results. The operating characteristics of the fluidization unit were observed and discussed in conjunction with the data obtained from the experimental tests. The calculated results were based on the average of at least three readings taken at each point over a recorded period of time, under steady state conditions. The coefficients of heat transfer were based on the internal area of the pipe wall and the integrated mean temperature-difference between the wall and the bulk bed. The effects of mass superficial air velocity and temperature driving-force on the coefficient of heat transfer were evaluated. Bed-temperature gradients were determined in an effort to substantiate previously described heat-transfer characteristics in the fluidized state.

Materials

The following materials were used in this investigation:

Air. Compressed air, humidity, 0.01 pounds water vapor per pound dry air, supplied from the Nash Hytor compressor, Department of Chemical Engineering, Virginia Polytechnic Institute, Blacksburg, Virginia. Used as the fluidizing medium.

Sand. Ottawa, 99.98% SiO_2; 100% through U. S. Standard sieve #20, 100% retained on U. S. Standard sieve #30; meets specifications of A. S. T. M. Designation C-109; 0.026-inch average particle diameter. Obtained from Department of Ceramic Engineering, Virginia Polytechnic Institute, Blacksburg, Virginia. Used as the fluidized solid.

Apparatus

The following pieces of apparatus were used in this investigation:

Battery, Dry Cell. "Eveready," 1.5 volt, number 6. Manufactured by the National Carbon Company, Cleveland, Ohio. Used in conjunction with the potentiometer.

Cell, Standard. Number 392884, 1.0184 volts, internal resistance not over 500 ohms. Manufactured by Epperly Laboratory, Inc., Newport, R. I. Obtained from Fisher Scientific Co., Pittsburgh, Pa. Used in conjunction with the potentiometer.

Fluidization Unit and Accessory Equipment. All materials required for the construction of the fluidization unit and accessory equipment are included in Table I.

Galvanometer. Center zero type, catalog number 570-201. Manufactured by G-M Laboratories, Inc., Chicago, Ill. Used in conjunction with the potentiometer.

TABLE L

Bill of Materials for Fluidization Unit

TABLE I

Bill of Materials for Fluidization Unit and Accessory Equipment

Number Required	Unit	Name	Description	Supplier
EQUIPMENT SUPPORTS				
50	ft	Angle Iron	1-½ x 1-½ x 1/8 inch	Noland Co., Inc., Roanoke, Va.
10	ft	Angle Iron	1 x 1 x 1/8 inch	Noland Co., Inc., Roanoke, Va.
5½	ft	Angle Iron	3/4 x 3/4 x 1/16 inch	Noland Co., Inc., Roanoke, Va.
4	ft	Lumber	2 x 4 inch	New River Lumber Co., Blacksburg, Va.
14	ft	Lumber	¾ x 4 inch	New River Lumber Co., Blacksburg, Va.
HEATER AND ELECTRICAL EQUIPMENT				
115	ft	Nichrome Resistance	14 B & S gage	Driver-Harris Co., Harrison, N. Y.
1	ea	Wire Switch Box	Murray Safety Switch, No. 242, 30 amp, 125/250 v	Metropolitan Device Corp., Brooklyn, N. Y.
2	ea	Insulators	3 x ½-inch diameter, porcelain	Electric Service Departmen Virginia Polytechnic Institute, Blacksburg, Va.
35	ft	Cable	Two wire, single strand, copper, 15 amp	Electric Service Department, Virginia Polytechnic Institute, Blacksburg, Va.
15	ft	Wire	Single strand, fabric insulated, 15 amp	Electric Service Department, Virginia Polytechnic Institute, Blacksburg, Va.
MANOMETERS				
18	ft	Glass Tubing	6-mm bore, pyrex	Fisher Scientific Co., Pittsburgh, Pa.
5	ft	Tygon tubing	¼-inch bore	Fisher Scientific Co., Pittsburgh, Pa.
12	ea	Graph Paper	20 x 20 divisions per inch, 8-½ x 11 inch	Everett Waddey Co., Richmond, Va.
10	gm	Glyptal	No. 1201, red enamel	General Electric Co., Schenectady, N. Y.
PIPE AND FITTINGS				
1	ft	Pipe	5-inch nominal diameter, black iron	Noland Company, Inc., Roanoke, Va.
4	ft	Pipe	4-inch nominal diameter, blacks iron	Noland Company, Inc., Roanoke, Va.
5	ft	Pipe	3-inch nominal diameter, black iron	Noland Co., Inc., Roanoke, Va.
18	ft	Pipe	1-½-inch nominal diameter, black iron	Noland Co., Inc., Roanoke, Va.
6	ft	Pipe	1-inch nominal diameter, black iron	Noland Co., Inc., Roanoke, Va.
1	ft	Pipe	½-inch nominal diameter, black iron	Noland Co., Inc., Roanoke, Va.
40	ft	Pipe	1/8-inch nominal diameter, black iron	Noland Co., Inc., Roanoke, Va.
10	ea	Short Nipple	1-½ inch nominal, black	Rose-Stevens Supply Corp., Radford, Va.
4	ea	Union	1-½ inch ground joint, black iron	Rose-Stevens Supply Corp., Radford, Va.
3	ea	Tee	1-½ inch, standard, black iron	Rose-Stevens Supply Corp., Radford, Va.
6	ea	Ell	1-½ inch, 90°, black iron	Rose-Stevens Supply Corp., Radford, Va.
4	ea	Reducing Tee	1-½ x 1-½ x 1-inch, black iron	Rose-Stevens Supply Corp., Radford, Va.
1	ea	Reducing Ell	1-½ x 1-inch, 90°, black iron	Rose-Stevens Supply Corp., Radford, Va.
1	ea	Reducing Coupling	3 x 1-½-inch, black iron	Rose-Stevens,Supply Corp., Radford, Va.
1	ea	Coupling	3-inch, black iron	Noland Co., Inc., Roanoke, Va.
1	ea	Coupling	1-½-inch, black iron	Noland Co., Inc., Roanoke, Va.
2	ea	Plug	1-½-inch, black iron	Noland Co., Inc., Roanoke, Va.
1	ea	Flange	1-inch, black iron	Noland Co., Inc., Roanoke, Va.
17	ea	Close Nipple	1-inch, black iron	Noland Co., Inc., Roanoke, Va.
6	ea	Union	1-inch, ground joint, black iron	Noland Co., Inc., Roanoke, Va.
3	ea	Ell	1-inch, 90°, black iron	Noland Co., Inc., Roanoke, Va.
3	ea	Tee	1-inch, standard, black iron	Rose-Stevens Supply Corp., Radford, Va.
1	ea	Reducing Tee	1 x 1 x 3/4-inch, black iron	Noland Co., Inc., Roanoke, Va.
1	ea	Reducing Coupling	1 x 3/4-inch, black iron	Noland Co., Inc., Roanoke, Va.
2	ea	Short Nipple	3/4-inch, black iron	Noland Co., Inc., Roanoke, Va.
2	ea	Union	3/4-inch, standard, blacks iron	Noland Co., Inc., Roanoke, Va.
2	ea	Ell	3/4-inch, 90°, black iron	Noland Co., Inc., Roanoke, Va.
1	ea	Reducing Tee	3/4 x 3/4 x ½-inch, Black iron	Noland Co., Inc., Roanoke, Va.
1	ea	Union	½-inch, ground joint, black iron	Noland Co., Inc., Roanoke, Va.
1	ea	Ell	½-inch, 90°, black iron	Noland Co., Inc., Roanoke, Va
2	ea	Cap	½-inch, standard, black iron	Noland Co., Inc., Roanoke, Va.
3	ea	Busing	½ x ¼-inch, black iron	Noland Co., Inc., Roanoke, Va.
1	ea	Bushing	1 x ½-inch, black iron	Noland Co., Inc., Roanoke, Va.
1	ea	Bushing	3/4 x ½-inch, black iron	Noland Co., Inc., Roanoke, Va.
3	ea	Close Nipple	1/8-inch, black iron	Noland Co., Inc., Roanoke, Va.
5	ea	Short Nipple	1/8-inch, black iron	Noland Co., Inc., Roanoke, Va.
14	ea	1-½-inch Nipple	1/8-inch, black iron	Noland Co., Inc., Roanoke, Va.
14	ea	2-inch Nipple	1/8-inch, black iron	Noland Co., Inc., Roanoke, Va.,
8	ea	Union	1/8-inch, ground joint, black iron	Noland Co., Inc., Roanoke, Va.
14	ea	Tee	1/8-inch, standard, black iron	Noland Co., Inc., Roanoke, Va.
5	ea	Short Nipple	1/8-inch, black iron	Noland Co., Inc., Roanoke, Va.
14	ea	1-½-inch Nipple	1/8-inch, black iron	Noland Co., Inc., Roanoke, Va.
14	ea	2-inch Nipple	1/8-inch, black iron	Noland Co., Inc., Roanoke, Va.,
2	ea	Ell	3/4-inch, 90°, black iron	Noland Co., Inc., Roanoke, Va.
1	ea	Reducing Tee	3/4 x 3/4 x ½-inch, Black iron	Noland Co., Inc., Roanoke, Va.
1	ea	Union	½-inch, ground joint, black iron	Noland Co., Inc., Roanoke, Va.
1	ea	Ell	½-inch, 90°, black iron	Noland Co., Inc., Roanoke, Va
2	ea	Cap	½-inch, standard, black iron	Noland Co., Inc., Roanoke, Va.
3	ea	Busing	½ x ¼-inch, black iron	Noland Co., Inc., Roanoke, Va.
1	ea	Bushing	1 x ½-inch, black iron	Noland Co., Inc., Roanoke, Va.
1	ea	Bushing	3/4 x ½-inch, black iron	Noland Co., Inc., Roanoke, Va.
3	ea	Close Nipple	1/8-inch, black iron	Noland Co., Inc., Roanoke, Va.
5	ea	Short Nipple	1/8-inch, black iron	Noland Co., Inc., Roanoke, Va.
14	ea	1-½-inch Nipple	1/8-inch, black iron	Noland Co., Inc., Roanoke, Va.
14	ea	2-inch Nipple	1/8-inch, black iron	Noland Co., Inc., Roanoke, Va.,
8	ea	Union	1/8-inch, ground joint, black iron	Noland Co., Inc., Roanoke, Va.
14	ea	Tee	1/8-inch, standard, black iron	Noland Co., Inc., Roanoke, Va.
38	ea	Ell	1/8-inch, 90°, black iron	Noland Co., Inc., Roanoke, Va.
VALVES				
5	ea	Gate Valve	1-½-inch, bronze	Rose-Stevens Supply Corp., Radford, Va.
4	ea	Gate Valve	1-inch, bronze	Rose-Stevens Supply Corp., Radford, Va.
1	ea	Needle Valve	1-inch, bronze	Rose-Stevens Supply Corp., Radford, Va.
1	ea	Gate Valve	3/4-inch, bronze	Rose-Stevens Supply Corp., Radford, Va.
1	ea	Globe Valve	½-inch, bronze	Rose-Stevens Supply Corp., Radford, Va.
11	ea	Globe Valve	1/8-inch, bronze	Rose-Stevens Supply Corp., Radford, Va.
THERMOCOUPLES				
125	ft	Thermocouple Wire	Iron-constantan, fiber-glass braid over glass wrap, 20 B&S gage, No. 9B3C2	Minneapolis-Honeywell Regulator Co., Brown Instruments Division, Washington, D. C.
10	ft	Ceramic Insulator	3-inch lengths of porcelain 2-hole tube, Catalog No. 42347-3	Minneapolis-Honeywell Regulator Co., Brown Instruments Division, Washington, D. C.
1	ea	Commutating Switch and Junction Box	Rectangular, steel constructed box with 12 poles	Brown Instrument Co., Philadelphia, Pa.
6	ea	Lead wire	Two wire, multiple strand, copper, rubber insulation	Brown Stores Co., Blacksburg, Va.
1	ea	Thermos Bottle	Model No. 2220, filler No. 22F	The American Thermos Bottle Co., Norwich, Conn.
4	ea	Switch	Toggle type, 3 amp, 250 v	Electric Service Dept. Virginia Polytechnic Institute, Blacksburg, Va.
MISCELLANEOUS				
30	lb	Insulation	"Celite," Standard Super-cel, powdered diatomaceous earth, filter aid	Johns-Manville New York, N. Y.
3.5	ft	Air Duct	10-inch diameter, galvanized steel	Noland Co., Inc., Roanoke, Va.
30	ea	Stove Bolts	3/13 x 1-½-inch,round head, steel	Noland Co., Inc., Roanoke, Va.
12	ea	Machine Bolts	¼ x 1-½-inch, square head, steel	Noland Co., Inc., Roanoke, Va.
5	sq ft	Asbestos Cloth	Woven fabric, 1/8-inch thick	Noland Co., Inc., Roanoke, Va.

Meter, Gas. Number 4354740, capacity 30 to 600 cubic feet per hour, graduated in cubic feet and tenths of cubic feet. Manufactured by the American Meter Company, Albany, N. Y. Used to measure the air flow in the orifice calibration.

Glassware. Miscellaneous beakers, funnels, and flasks. Obtained from Fisher Scientific Company, Pittsburgh, Pa. Used for handling the sova beads and the manometer fluids.

Potentiometer. Type s, range 0 to 0.017 and 0 to 1.70 volts. Manufactured by Fisher Scientific Co., Pittsburgh, Pa. Used to measure electromotive force from thermocouples.

Temperature Controller. "Capacitrol," model 224, range 0 to 1000 °C, rating 110 to 220 volts, 35 ampere. Manufactured by Wheelco Instrument Co., Chicago, Ill. Used to control the temperature of the fluidized bed wall.

Timer. "Precision Time-it," 0 to 9999.0 seconds, graduated in tenths of a second, 115 volt, 60 cycle, 5 watt. Manufactured by the Precision Scientific Co., Chicago, Ill. Used in conjunction with the gas meter to calibrate the orifice.

Meter, Watt-hour. Type OB, single phase, 60 cycle, 115 to 220 volts, 25 ampere, serial number 12346738. Manufactured by Westinghouse Electric and Manufacturing Co., Newark Works, Newark, N. J. Used to measure energy input to the fluidized bed heating unit.

Method of Procedure

The procedure followed in carrying out this investigation included the calibration of orifice manometer, calibration of the "capacitrol" temperature controller, determination of optimum operating conditions, and the operation of the fluidization unit.

Calibration of Orifice Manometer. The orifice manometer on the air inlet line was calibrated in the following manner. The one-inch air line directly below the orifice was connected to a dry gas meter of 30 to 600 cubic feet per hour capacity. Calibration readings were taken at a constant orifice pressure drop by obtaining the volume of air flow through the meter for a measured length of time. Three readings were made for each pressure drop, and the average of each set of readings was plotted versus the pressure drop to give the calibration for the orifice.

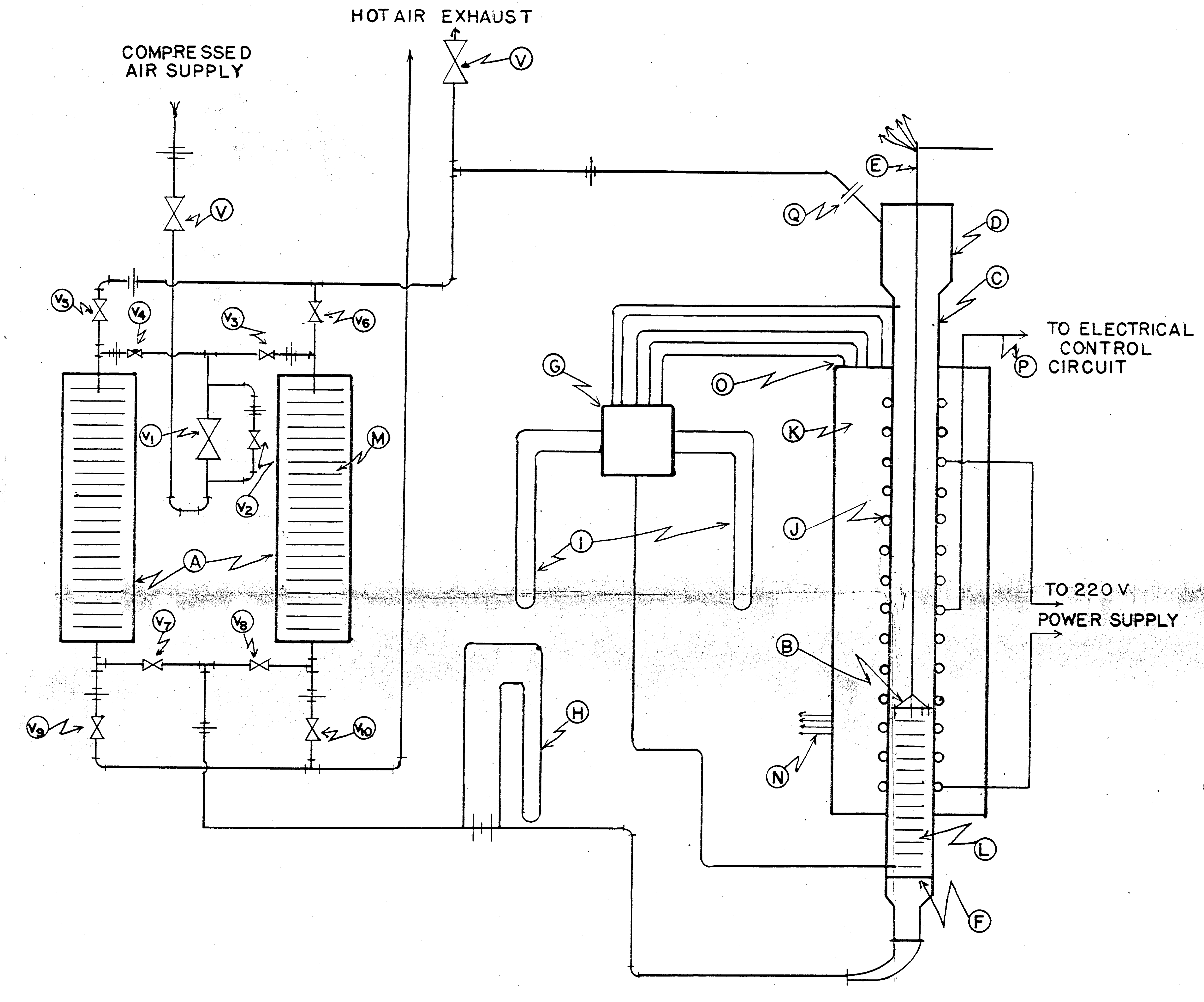
Calibration of "Capacitrol" Temperature Controller. The "capacitrol" was calibrated as follows: The wire lead from the control thermocouple was attached to the central poles of a double-pole, double-throw toggle switbh. A set of leads from one side of the switch was connected to a potentiometer. Another set of leads from the other side of the switch was connected to the temperature indicator of the "capacitrol." Thus, the switch could be used to direct the current generated in the thermocouple to either the potentiometer or the "capacitrol." The main heating element of the column was then energized, and the column wall temperature was allowed to reach approximately 200 °F. The thermocouple was connected to the potentiometer and the emf observed and recorded. As soon as the potentiometer reading was determined, the thermocouple was switched to the "capacitrol," and its indicated temperature was recorded. This procedure was repeated for temperatures of 400 and 600 °F, which were the temperatures to be used in operation of the unit. The calibration was accomplished by determining the temperature corresponding to the potentiometer reading, and comparing it with the temperature indicated by the "capacitrol." A suitable correction was made to the "capacitrol" reading.

Determination of Optimum Operating Conditions. The material to be used as the fluidized solid was placed in a transparent lucite plastic fluidization test apparatus, constructed by Seidel(77). Air was passed through the solid particles at known mass superficial air velocities. The range of velocities at which the bed was fluidized was determined and later used as the range of operation in the investigation of heat transfer characteristics of the solid in the fluidized state. The physical appearance and characteristics of the solid were observed and recorded over the range of fluidization velocities.

Operation of the Fluidization Unit. Preliminary to obtaining test data, the fluidization column was filled to a level 29 inches above the bottom of the heating unit with 20.5 pounds of Ottawa sand. The sand was introduced to the column through the air exhaust pipe of the disengaging section (Figure 2). The exhaust pipe was then connected by means of a flange to the system of outlet pipe.

Each test resulted in a series of experimental data obtained under steady state conditions of operation. Each test was conducted at a constant bed wall temperature and mass superficial air velocity. One test varied from another in that the wall temperature and/or the air

Figure 2. Schematic Diagram of Fluidization Unit



LEGEND

Ⓐ	AIR DRYING CHAMBERS, 4-IN. DIAMETER
Ⓑ	BED TRAVERSING THERMOCOUPLE AND MOUNT
Ⓒ	FLUIDIZATION CHAMBER, 3-IN. DIAMETER
Ⓓ	DISENGAGING SECTION, 5-IN. DIAMETER
Ⓔ	ADJUSTING ROD FOR BED THERMOCOUPLE, 3/8 IN. DIAMETER
Ⓕ	AIR DISTRIBUTION PLATE
Ⓖ	MANOMETER CONTROL BOARD
Ⓗ	MANOMETER FOR MEASURING INLET AIR FLOW RATE
Ⓘ	MANOMETER FOR MEASURING BED PRESSURE DROP
Ⓙ	BED NICHROME HEATING WIRE
Ⓚ	INSULATION, MAGNESIA-CELITE MIXTURE
Ⓛ	CATALYST BED, 30 IN. DEPTH
Ⓜ	DESICCANT, SOVA BEADS
Ⓝ	THERMOCOUPLE FOR MEASURING WALL TEMPERATURE
Ⓞ	MANOMETER LEADS FOR MEASURING BED PRESSURE DROP
Ⓟ	CONTROL THERMOCOUPLE FOR HEATING WIRE
Ⓠ	ENTRAINMENT PREVENTOR, COUPLE, 325-MESH SCREEN

V_1 THROUGH V_{10} – CONTROL VALVES

FIGURE 2 SCHEMATIC DIAGRAM OF FLUIDIZATION UNIT flow rate were changed. Tests at five air velocities were made at each of three bed-wall temperatures.

At the start of a test, the bed traversing thermocouples were placed at the L-30 (30 inches above the bottom of the heating unit) position in the fluidization chamber. The traversing mechanism was then held in place with a rod extending through the two matching holes on the rod and the unit support for that particular position. The temperature controller was then set at the desired bed-wall temperature. The main line switch and then the secondary switch for the column heating unit were closed. When the fluidization chamber wall reached the desired temperature, as indicated on the temperature indicator of the temperature controller, the air flow through the fluidization chamber was adjusted to the desired rate by manipulation of valves number V_1 and V_2, Figure 2. Valve number V_2 in the ¼-inch by-pass air line was used for fine adjustment of the air flow rate. The air-flow control valves were constantly adjusted to assure a constant rate of flow of air.

Valves number V_3 through V_{10}, Figure 2, were used to direct the flow of entering air through the left chamber, A. In the original design of the unit, the two chambers, *A*, were used singly for drying the inlet air. However, the air velocities used in this investigation were so low (4.24 to 9.00 pounds dry air per hour) that the exhaust gases from the fluidization unit were at ambient temperature by the time they had reached the drying chamber through which the inlet air was not passing, and thus were useless for regeneration of its absorbing medium. For this reason, the air used in all the tests was directed through the left chamber, and a constant check was made to ensure that its humidity remained both constant and low (0.01 pound water vapor per pound dry air). To thus direct the entering air, valves number $\underline{V_4}$ and $\underline{V_8}$ were opened and valves number $\underline{V_5}$ and $\underline{V_9}$ were closed. To pass the exhaust gases through the right chamber, valves number $\underline{V_6}$ and $\underline{V_{10}}$ were opened and valves number $\underline{V_3}$ and $\underline{V_7}$ were closed.

The manometer control board shown in Figure 2 was used in conjunction with two manometers to measure the pressure drop through the fluidized bed. Proper manipulation of the 1/8-inch valves located on the control panel allowed the pressure drop between any manometer taps in the fluidization chamber to be measured.

After about a two-hour heating period, preliminary temperature readings were made at 15-minute intervals to ascertain when steady-state conditions were reached.

TABLE II

Location of Thermocouples Employed in the Fluidization Unit

Thermocouple No.	Position in Fluidization Unit
	Fluidization chamber wall[a]:
1	L-31
2	L-30
3	L-24
4	L-18
5	L-12
6	L-6
7	L-0
8	L-(-1)
	Fluidized bed:
9	1-1/2 inches from inner chamber wall
10	7/8 inches from inner chamber wall
11	3/8 inches from inner chamber wall
12	1/8 inches from inner chamber wall
	Fluidization Chamber Insulation:
13	Outer position
14	Inner position
	Entering air line:
15	Air entering fluidization unit

[a] The code numbers refer to the position of the thermocouple with reference to the lowest level, L-0, of the heating section.

The preliminary temperature readings were taken at thermocouple positions number 4, 9, 13, and 15(Table II). When successive readings were essentially constant, the series of readings constituting the test were begun.

The data observed during a test consisted of the energy input to the bed heating unit, orifice pressure differential (held constant), air line pressure above the orifice, the pressure drop over various sections of the fluidized bed, insulation temperature, temperature of the air entering the unit, the bed wall temperature at six-inch intervals along the 30-inch heating unit, bed temperatures at each six-inch level, wall temperature one inch above and one inch below the heating unit, room temperature, and barometric pressure. Necessary time intervals were recorded during each series of readings. The indicator on the temperature controller was checked constantly to ensure that the wall temperature of the column was maintained constant. The potentiometer used for indicating thermocouple data was re-standardized several times during each set of readings.

Data were obtained with Ottawa sand as the fluidized solid and air as the fluidizing medium for a series of 15 tests. The variables considered were three bed wall temperatures (200, 400, and 600 °F) and four mass, superficial air velocities (82.5, 123.2, 170.3, and 217.5 pounds per hour-square foot). An additional test was made at each bed wall temperature at a mass superficial air velocity of 175.0 pounds per hour-square foot with no solids present in the unit to determine the relative value of the presence of fluidized solids in increasing the coefficient of heat transfer. Each temperature recorded in a test was checked at least three times during the test over a period of about two hours, and the values presented herein are averages of the temperatures thus recorded.

Data and Results

The data and results of this investigation on heat transfer to a fluidized bed are arranged in the following order.

Observation of Fluidization Characteristics of a Bed of Ottawa Sand. Bed expansion, degree of fluidization, and physical appearance of the bed are listed as functions of mass superficial air velocity in Table III, as determined in a transparent lucite plastic fluidization test apparatus.

Calibration Data for Orifice in Inlet Air Line. Volume flow per unit time is listed with corresponding pressure drops as observed on inlet air line orifice manometer, in Table IV.

Air Inlet Orifice Calibration Curve. The data listed in Table IV is plotted in Figure 3.

Summary of Data and Results of Heat Transfer to an Air Fluidized Bed of Ottawa Sand at Bed Wall Temperatures of 200, 400, and 600 °F. Included in Table V are air flow rates, insulation temperatures, bulk bed and bed-wall temperatures, heat flow rates, mean temperatures differences, bed-wall temperature drops, and coefficients of heat transfer for 15 tests conducted on the Breckon(1) fluidized bed heat transfer apparatus.

TABLE III

Observation of Fluidization Characteristics of a Bed of Ottawa Sand[a]

Observation No.	Mass Air Velocity lb/hr-sq ft	Bed Expansion[b] %	Notes on Condition of Bed
1	65.0	2.4	No motion at bed wall. Only surface of bed shows sign of movement. Not fluidized.
2	91.0	5.4	Slight movement at bed wall. About one small bubble per second may be seen rising through bed. Fluidized.
3.	108.0	9.0	Increased movement at bed wall. About 3 bubbles per second pass through bed.
4	123.0	12.5	Bubbles have nearly reached column diameter as they approach bed surface.
5	145.0	17.3	Large bubbles visible throughout entire bed, but no channeling or slugging exist.
6	186.0	22.1	Slugging occurs in upper half of bed. Sand is thrown as much as 3 inches into the air by bursting slugs.
7	210.0	36.5	Slugging occurs throughout entire bed.

[a] As determined in transparent lucite plastic fluidization test unit, 2.75 inches in diameter with an initial bed depth of 10.5 inches.

[b] Based on height of bed at zero air velocity.

TABLE IV

Calibration Data for Air Inlet Orifice

Orifice Pressure Drop, in. H_2O	Air Flow* Rate, cu ft/min
0.25	1.20
0.70	1.62
1.21	2.09
2.18	2.90
4.12	3.87
6.10	4.82
8.00	5.58

* Measured at 77 °F and 28.1 inches of mercury.

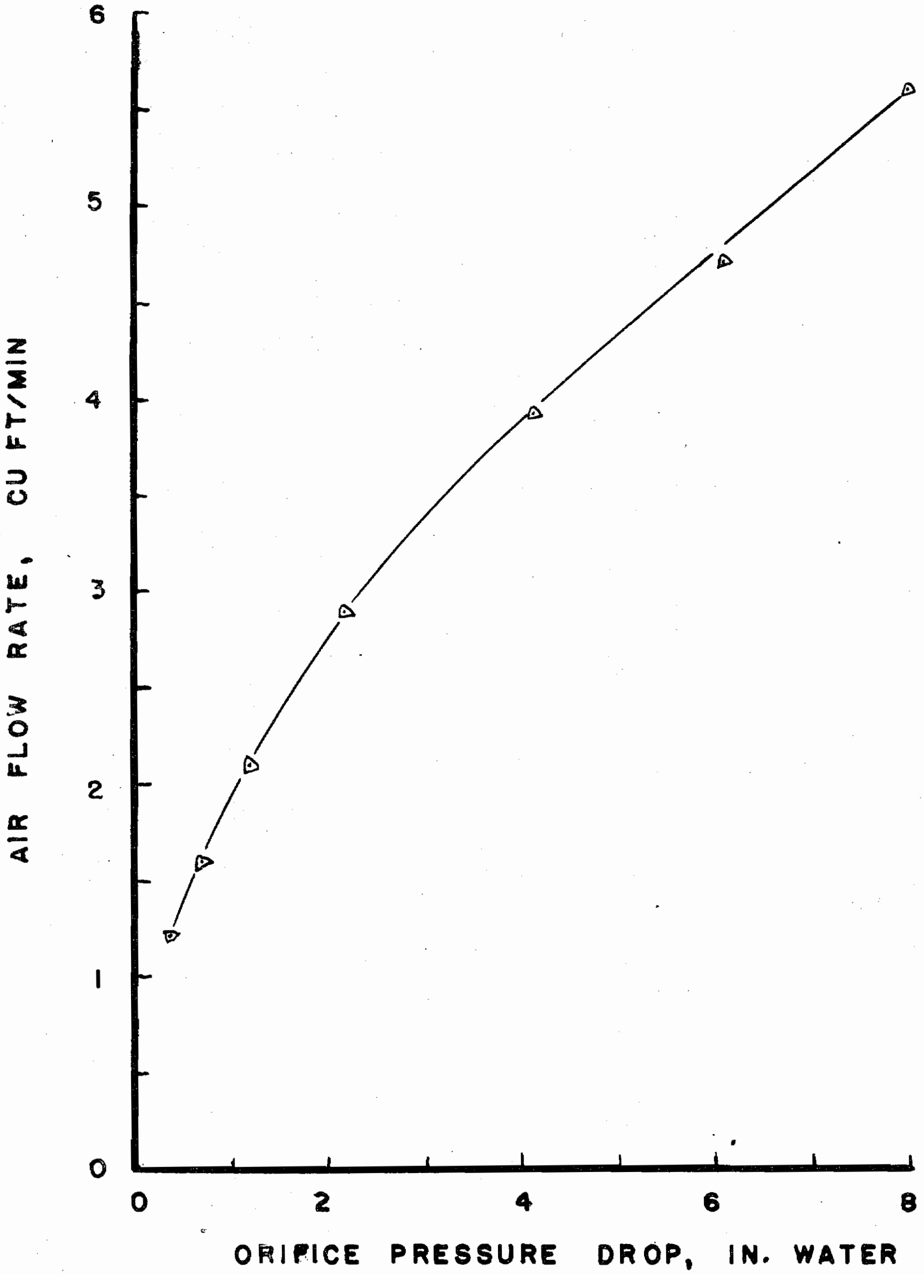


FIGURE 3. CALIBRATION CURVE FOR AIR INLET ORIFICE.

TABLE V

Summary of Data and Results of Heat Transfer to an Air Fluidized Bed of Ottawa Sand at Wall Temperatures of 200, 400, and 600 °F

TABLE V

Summary of Data and Results of Heat Transfer to an Air Fluidized Bed of Ottawa Sand at Wall Temperatures of 200, 400, and 600 °F

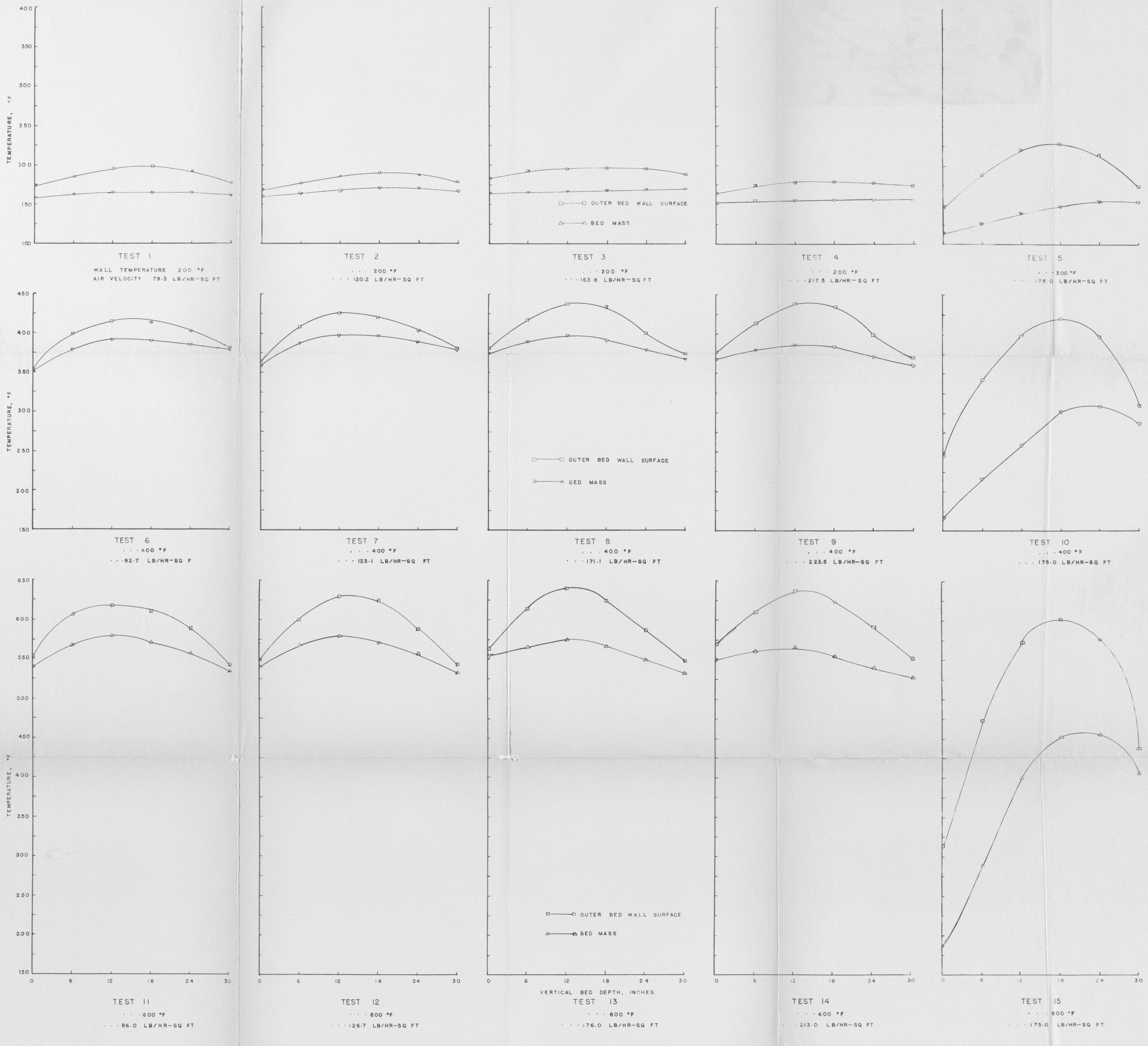
Test No.	Bed Wall Temperature, °F	Air Inlet Orifice Pressure Drop, in. water	Air Pressure (gage), in. Hg	Electric Energy Input Rate, kw-hr/hr	(Dry Basis) Air Flow Rate, lb/hr	Mass Superficial Air Velocity, lb/hr-sq ft	Air Humidity, lb water/lb dry air	Temperatures for Heat Loss Determination: Outer Insulation °F	Inner Insulation °F	Upper Wall L-31 °F	Lower Wall L-(-1) °F	Air Inlet Temperature, °F	Total Heat Losses, Btu/hr	Heat Input by Electric Energy, Btu/hr	Vertical Distance from Bottom of Heating Element, In.	Bed Temperature, Radial Positions, Distance from Inner Bed Wall: 1-1/2 in. °F	7/8 in. °F	3/8 in. °F	1/8 in °F	Bulk Bed Temperature, °F	Temperature Outer Wall Surface, °F	Heat Added to Air, Btu/hr	Error in Heat Balance, based on Heat Added to Air, %	Over-all Integrated Mean Temperature Difference, °F	Temperature Drop Through Bed Wall, °F	Boundary Mean Temperature Difference, °F	Boundary Coefficient of Heat Transfer, Btu/hr-sq ft-°F
1	200	0.25	4.90 5	0.063	4.06	79.3	0.01	117	148	172	170	76	132	216	0 6 12 18 24 30	158 163 165 165 164 162	157 161 163 164 163 160	158 162 164 165 164 162	158 162 164 165 165 162	158 162 164 164 164 161	174 186 195 195 193 179	84	11.0	18.75	0.026	18.72	2.33
2	200	0.40	5.10	0.076	6.17	120.3	0.01	114	145	175	165	76	102	258	0 6 12 18 24 30	161 164 167 171 171 167	161 164 166 170 170 168	162 164 168 171 170 168	161 163 168 171 170 167	161 164 167 171 170 167	168 178 [illegible] 191 184 175	136	14.7	19.36	0.042	19.32	[illegible]
3	200	0.75	4.75	0.095	8.40	164.0	0.01	120	152	183	180	79	115	323	0 6 12 18 24 30	164 167 168 169 169 169	163 166 166 167 168 168	164 166 168 169 169 169	164 167 168 169 169 169	164 166 167 168 169 169	[illegible] [illegible] 195 190 195 187	183	13.6	25.73	0.057	25.67	3.54
4	200	1.25	6.10	0.109	11.16	217.5	0.01	117	152	170	164	67	113	373	0 6 12 18 24 30	153 155 155 155 155 155	153 154 154 155 155 153	153 155 154 155 155 155	153 155 155 156 156 155	153 155 155 155 155 155	166 177 178 178 176 172	220	10.6	25.97	0.073	25.90	3.78
5	200	1.25	0.00	0.105	8.97	175.0	0.01	112	142	160	135	80	190	360	0 6 12 18 24 30	112 126 138 146 148 146	110 124 [illegible] 143 149 143	111 129 143 150 157 154	110 124 139 149 155 154	110 125 [illegible] 147 154 [illegible]	111 [illegible] 119 [illegible] [illegible] 172	157	8.9	58.91	0.049	58.86	1.33
6	400	0.25	5.90	0.225	4.24	82.7	0.01	166	254	370	336	76	368	703	0 6 12 18 24 30	354 384 392 391 383 377	357 385 392 390 387 381	356 381 393 391 387 381	346 378 391 389 384 378	353 380 392 390 385 379	358 399 416 410 403 381	311	7.7	16.96	0.096	16.86	9.18
7	400	0.40	6.10	0.256	6.31	123.1	0.01	173	261	368	348	80	363	871	0 6 12 18 24 30	367 388 398 396 386 377	366 388 399 397 388 379	362 390 396 399 390 380	349 386 396 396 386 378	369 388 397 398 389 379	363 409 426 415 403 379	458	10.9	17.51	0.142	17.37	12.91
8	400	0.75	6.30	0.313	8.82	17[illegible]	0.01	178	274	360	371	77	374	1070	0 6 12 18 24 30	373 390 395 [illegible] 377 [illegible]	375 391 386 392 381 [illegible]	374 388 397 394 383 370	373 392 397 389 373 364	374 390 397 391 380 368	383 417 436 420 401 374	624	10.0	25.03	0.193	24.84	13.45
9	400	1.25	7.50	0.365	11.45	223.0	0.01	177	274	355	365	77	387	1243	0 6 12 18 24 30	[illegible] 383 386 382 365 357	371 380 385 383 371 360	366 396 387 384 372 360	367 384 387 386 369 357	368 381 386 384 370 359	376 414 437 430 398 370	785	9.1	31.52	0.243	31.28	13.50
10	400	1.25	0.00	0.232	8.97	175.0	0.01	161	241	282	218	80	462	962	0 6 12 18 24 30	172 230 271 294 298 279	166 219 257 294 300 280	167 223 266 312 317 290	169 213 258 304 311 290	167 218 260 304 309 287	241 346 401 399 400 309	450	10.9	97.50	0.139	97.36	2.30
11	600	0.25	7.00	0.338	4.42	86.2	0.01	237	394	528	536	69	602	1156	0 6 12 18 24 30	546 573 579 565 553 528	541 568 577 569 554 529	539 570 579 574 563 538	541 568 580 570 556 536	540 569 579 571 558 535	558 607 616 [illegible] 589 543	500	11.0	[illegible]	0.155	[illegible]	9.58
12	600	0.40	6.50	0.415	6.50	126.5	0.01	237	391	524	533	70	582	1413	0 6 12 18 24 30	547 572 577 565 550 533	559 570 573 568 557 [illegible]	542 566 579 572 557 536	542 568 578 569 556 531	541 568 579 569 557 533	549 600 630 610 587 543	730	13.7	27.53	0.226	27.30	13.40
13	600	0.75	7.50	0.498	9.04	176.0	0.01	255	416	526	547	82	622	1700	0 6 12 18 24 30	553 569 572 555 546 527	551 568 573 563 549 529	553 565 574 568 538 544	559 556 581 568 550 535	554 566 576 567 549 533	563 614 640 610 588 549	938	9.1	27.31	0.306	27.00	13.21
14	600	1.25	6.30	0.579	10.90	213.0	0.01	262	428	528	553	84	645	1977	0 6 12 18 24 30	552 565 563 549 536 526	550 562 564 552 539 [illegible]	547 568 565 556 545 528	552 562 570 558 543 529	550 561 567 555 542 538	569 610 637 615 593 553	1172	13.6	44.36	0.364	44.00	13.30
15	600	1.25	0.00	0.439	8.97	175.0	0.01	208	334	388	295	74	685	1496	0 6 12 18 24 30	182 296 392 439 442 392	180 286 380 438 439 396	183 294 397 464 467 411	195 282 421 456 463 414	186 287 399 453 456 407	314 473 572 580 573 435	733	10.6	142.08	0.227	141.85	2.57

Vertical Wall and Bulk Bed Temperature Gradients for Individual Tests. The vertical wall and bulk bed temperature gradient of each of the 15 tests made on the fluidized bed heat transfer apparatus are plotted in Figure 4 in such a manner that mean temperature difference between bed and wall may be determined graphically.

Effect of Mass Superficial Air Velocity on Heat Flow to an Air Fluidized Bed of Ottawa Sand. The heat flow to the fluidizing medium in Btu per hour is plotted in Figure 5 against the mass superficial air velocity for bed wall temperatures of 200, 400, and 600 °F.

Effect of Mass Superficial Air Velocity on the Bed-Wall Coefficient of Heat Transfer in an Air Fluidized Bed. The bed-wall coefficient of heat transfer of an externally heated fluidized bed is plotted in Figure 6 against the mass superficial air velocity at external bed-wall temperatures of 200, 400, and 600 °F.

Figure 4. Vertical Wall and Bulk Bed Temperature Gradients for Individual Tests



TEST 1
WALL TEMPERATURE 200 °F
AIR VELOCITY 79.3 LB/HR-SQ FT

TEST 2
· · · 200 °F
· · · 120.2 LB/HR-SQ FT

TEST 3
· · · 200 °F
· · · 163·8 LB/HR-SQ FT

TEST 4
· · · 200 °F
· · · 217.5 LB/HR-SQ FT

TEST 5
· · · 200 °F
· · · 175.0 LB/HR-SQ FT

TEST 6
· · · 400 °F
· · · 82·7 LB/HR-SQ F

TEST 7
· · · 400 °F
· · · 123·1 LB/HR-SQ FT

TEST 8
· · · 400 °F
· · · 171·1 LB/HR-SQ FT

TEST 9
· · · 400 °F
· · · 223.5 LB/HR-SQ FT

TEST 10
· · · 400 °F
· · · 175·0 LB/HR-SQ FT

TEST 11
· · · 600 °F
· · · 86·0 LB/HR-SQ FT

TEST 12
· · · 600 °F
· · · 126·7 LB/HR-SQ FT

TEST 13
· · · 600 °F
· · · 176·0 LB/HR-SQ FT

TEST 14
· · · 600 °F
· · · 213·0 LB/HR-SQ FT

TEST 15
· · · 600 °F
· · · 175·0 LB/HR-SQ FT

FIGURE 4. VERTICAL WALL AND BULK BED TEMPERATURE GRADIENTS FOR INDIVIDUAL TESTS.

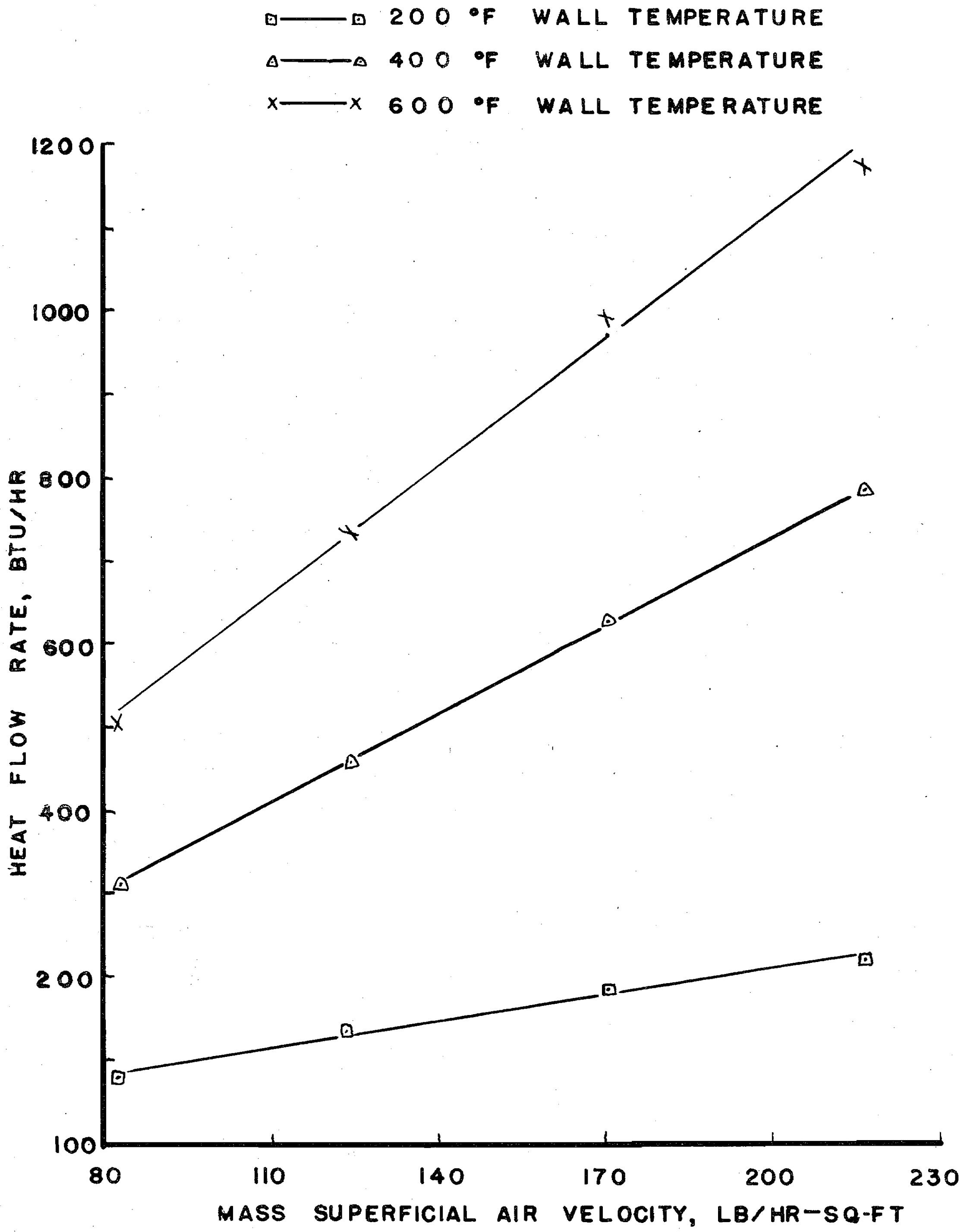


FIGURE 5. EFFECT OF MASS SUPERFICIAL AIR VELOCITY ON HEAT FLOW RATE TO THE FLUIDIZED BED.

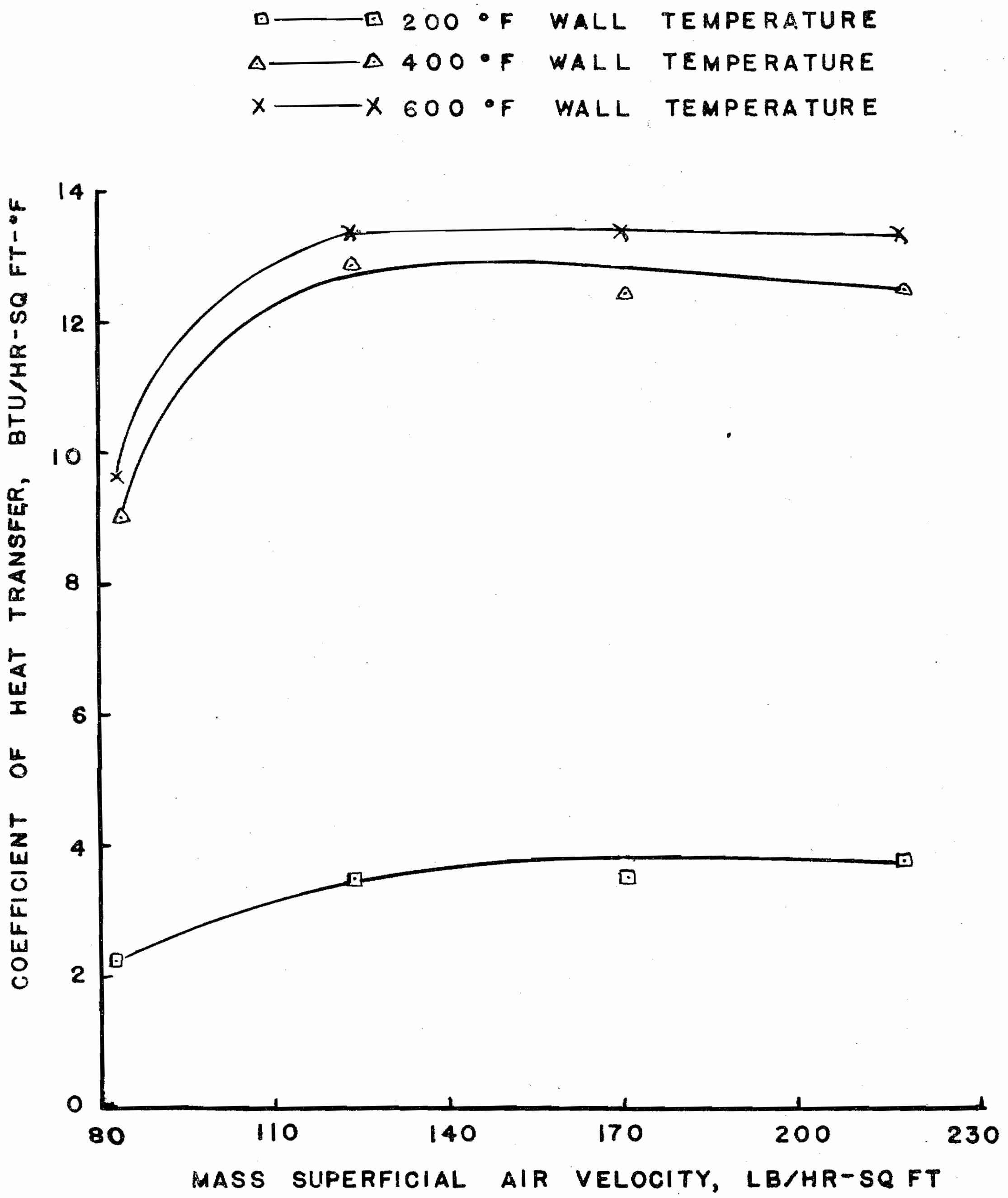


FIGURE 6. EFFECT OF MASS SUPERFICIAL AIR VELOCITY ON BED WALL COEFFICIENT OF HEAT TRANSFER.

Sample Calculations

The sample calculations include the determination of volume rate of air flow, mass superficial air velocity, rate of heat flow to fluidized bed, over-all integrated mean temperature difference, temperature drop due to pipe wall, bed wall boundary mean temperature difference, coefficient of heat transfer at the bed wall boundary, and a heat balance over the entire fluidization unit.

The calculations that were made from the data of Test 8 are shown as follows:

Calculation of Volume Rate of Air Flow at Standard Pressure. The air inlet orifice pressure drop was 0.75 inch of water, which corresponded to a volume flow rate of 105.0 cubic feet per hour as indicated on the orifice calibration curve, Figure 3, page 72. This indicated flow rate must be corrected for increase in line pressure caused by resistance of the fluidized bed and the exhaust line.

The pressure in the line was calculated as follows:

$$P_2 = P_1 + P_a \qquad (16)$$

where

P_2 = absolute line pressure, in. mercury

P_1 = indicated line pressure, in. mercury

P_a = barometric pressure, in. mercury.

$$P_2 = 6.3 + 28.1$$

$$P_2 = 34.4 \text{ in. mercury, absolute.}$$

The corrected volume flow rate for standard pressure was determined from the following relation:

$$V_c = \frac{P_2}{P_s} V \qquad (17)$$

where

V_c = corrected volume flow rate, cu ft/hr

P_s = standard atmospheric pressure, 29.9 in. mercury.

$$V_c = \frac{34.4}{29.9} (105.0)$$

$$V_c = 120.9 \text{ cu ft/hr.}$$

Calculation of Mass Superficial Air Velocity. The wet and dry bulb temperatures of the entering air were found to be 64 and 77 °F, respectively, corresponding to an absolute humidity of 0.01 pound water per pound dry air. The specific volume of the air was found to be 13.72 cubic feet per pound, from humidity tables. The mass superficial air velocity, G, based on an empty column, was determined as follows:

$$G = \frac{V_e}{\frac{1}{\rho_s} A_t} \qquad (18)$$

where

G = mass superficial air velocity, lb/hr-sq ft

$\frac{1}{\rho_s}$ = specific volume of air, cu ft/lb

A_t = cross section area of fluidization chamber, sq ft.

$$G = \frac{120.9}{\frac{(13.72)(7.393)}{144}}$$

$$G = 171.5 \text{ lb/hr-sq ft.}$$

Calculation of the Rate of Heat Flow to the Fluidized Bed. The rate of heat flow to the fluidized bed over the 30-inch heating section was based on the increase in air temperature between inlet and exit streams.

$$q = W C_p (t_o - t_i) \tag{19}$$

where

q = rate of heat flow to fluidized bed, Btu/hr

W = mass flow rate of air, lb/hr

C_p = specific heat capacity of air over the range between 77 and 400 °F, Btu/lb-°F

t_o = outlet temperature of air, °F

t_i = inlet temperature of air, °F.

$$q = (8.82)(0.244)(368 - 77)$$

$$q = 624 \text{ Btu/hr.}$$

Calculation of Over-all Integrated Mean Temperature Difference. The over-all mean temperature difference between the outer wall surface and the bed mass was obtained graphically, as indicated in Figure 4, page 75. The area between the outer wall temperature and the bulk bed temperature curves was divided by the longitudinal distance along the column to obtain the mean temperature difference.

$$\Delta t_m = \frac{A'}{L'} \qquad (20)$$

where

Δt_m = over-all integrated mean temperature difference, $^{\circ}F$

A' = area between outer wall surface and bulk bed temperature curves obtained graphically from Figure 4, $^{\circ}F$-in.

L' = height of column, in.

$$\Delta t_m = \frac{750.8}{30}$$

$$\Delta t_m = 25.03\ ^{\circ}F.$$

Calculation of Temperature Drop Across Pipe Wall.

The bed wall resistance to heat flow was calculated as follows:

$$R_w = \frac{x}{k\ A_m} \qquad (21)$$

where

R_w = thermal resistance through pipe wall, °F-hr/Btu

x = pipe wall thickness, ft

k = thermal conductivity of iron pipe, Btu/hr-sq ft-°F/ft

A_m = log mean area of 30-inch section of pipe, sq ft.

$$R_w = \frac{0.018}{(27)(2.155)}$$

$$R_w = 0.000310 \ \frac{\text{hr-°F}}{\text{Btu}}$$

The temperature drop through the pipe wall was determined from the relationship

$$\Delta t_w = q\ R_w \qquad (22)$$

where

Δt_w = temperature drop across pipe wall, °F.

$$\Delta t_w = (624)(0.000310)$$

$$\Delta t_w = 0.193\ °F.$$

Calculation of Bed Wall Boundary Mean Temperature Difference. The boundary mean temperature difference was obtained by subtracting the temperature drop caused by the bed wall resistance from the over-all integrated mean temperature difference.

$$\Delta t_m'' = \Delta t_m - \Delta t_w \qquad (23)$$

where

$\Delta t_m''$ = boundary mean temperature difference, °F

$$\Delta t_m'' = 25.03 - 0.19$$

$$\Delta t_m'' = 24.84\ ^\circ F.$$

Calculation of the Coefficient of Heat Transfer at the Bed Wall Boundary. The coefficient of heat transfer was calculated as follows:

$$h'' = \frac{q}{A''\Delta t_m''} \qquad (24)$$

where

h'' = coefficient of heat transfer at the bed wall boundary, Btu/hr-sq ft-°F

A'' = area of heat transfer surface, inner pipe wall area between levels L-0 (bottom) and L-30 (top), sq ft.

$$h'' = \frac{624}{(2.01)(24.84)}$$

$$h'' = 12.50 \text{ Btu/hr-sq ft-}^\circ\text{F}.$$

Determination of Heat Balance Over Fluidization Unit. The calculation of a heat balance across the fluidization unit involved the following: the rate of heat input by electric energy, the rate of heat content increase by air stream, the rate of heat loss up and down bed wall, and rate of heat loss through insulation surrounding the fluidized bed.

Heat Input by Electric Energy. The total heat input was calculated using the equation:

$$q = K_w F \qquad (25)$$

where

q = heat input, Btu/hr

K_w = rate of electric energy input, Kw-hr/hr

F = conversion factor, kw-hr to Btu.

$$q = (0.315)(3413)$$

$$q = 1070 \text{ Btu/hr.}$$

Conduction Loss at Top of Column Wall. The heat loss by conduction at the top of the pipe wall was obtained from the relationship:

$$q = k_c A_c \frac{t_w30 - t_w31}{x_c} \qquad (26)$$

where

k_c = thermal conductivity of iron pipe, Btu/hr-sq ft-°F/ft

A_c = cross section area of pipe wall, sq ft

t_w30 = temperature of column wall at L-30 level, °F

t_w31 = temperature of column wall at L-31 level, °F

x_c = distance between levels L-30 and L-31, ft.

$$q = (27)(0.0155)\frac{(374 - 360)}{0.0833}$$

$$q = 70.3 \text{ Btu/hr.}$$

Similarly, the conduction loss at the bottom of the pipe wall was found to be 60.2 Btu/hr.

Heat Loss Through Insulation. The heat loss through the celite insulation was calculated from the following equation:

$$q = k_{im} A_{im} \frac{t_i1 - t_i2}{x_i} \qquad (27)$$

where

k_{im} = thermal conductivity of celite at mean temperature, Btu/hr-sq ft-°F/ft

A_{im} = logarithmic mean area of insulation, sq ft

t_i1 = temperature of insulation at inner position, °F

t_i2 = temperature of insulation at outer position, °F

x_i = thickness of insulation between the two thermocouple positions, ft.

$$q = (0.040)(5.51)\frac{(274 - 178)}{0.0833}$$

$$q = 254 \text{ Btu/hr.}$$

Corrected Heat Input to Fluidized Bed. The corrected heat input was obtained by subtracting the sum of the calculated heat losses from the electric energy input.

$$q = 1070 - (70.3 + 60.2 + 254.0)$$

$$q = 685.5 \text{ Btu/hr.}$$

<u>Error Based on Heat Flow to Air</u>. The percentage error in the calculated amount of heat flow was determined, based on the heat flow to air through the fluidized bed.

$$\% \text{ error} = \frac{(686 - 624)}{624} (100)$$

$$\text{error} = 10.0\%$$

IV. DISCUSSION

The discussion section of the investigation of heat transfer to an externally heated, fluidized bed of ottawa sand contains a presentation of the data and results obtained, recommendations for further work on the subject, and the limitations involved in the experimental procedure.

Discussion of Results

This section includes a discussion of the various factors of consequence in obtaining the results of the investigation. Topics covered include: the use of humid air as fluidizing medium, fluidization characteristics of bed material, conditions of operation, determination of equilibrium conditions, horizontal bed temperature gradients, temperature drop through bed wall, vertical bed temperature gradients, boundary mean temperature difference effect of mass superficial air velocity on the bed wall coefficient of heat transfer, and comparison of coefficient of heat transfer between empty and fluidized beds.

Use of Humid Air as Fluidizing Medium. The fluidization unit, as designed and constructed by Breckon(1), included two air drying chambers (Figure 2) filled with sova bead desiccant. These chambers were originally used for drying the air entering the fluidized bed. For any given test, one chamber was used for drying while the other was being regenerated by passing hot exhaust gases from the fluidization unit through it. The effectiveness of this procedure is somewhat doubtful because there was no way of checking on the drying efficiency of the unit during a test, and there was no assurance that the moisture content of the air entering the unit was constant, even though it may have been very low. In one of the preliminary tests of this investigation, the mass air flow rates employed were so low, 4.24 to 9.00 pounds per hour, that by the time the air had passed through the uninsulated piping between the fluidized bed and the chamber being regenerated, its temperature had decreased to that of the surrounding atmosphere. Thus it was impossible to remove moisture from the desiccant at such a low temperature. Several tests were made to determine the moisture content of the air supply without drying, and the absolute humidity over a period of two weeks was found to be constant at 0.01 pounds water vapor per pound dry air. Furthermore, these tests showed no signs of entrained water droplets; thus it was assumed that the air used in this investigation had a constant specific heat for all tests over a given temperature range.

Fluidization Characteristics of Bed Material. Observations were made of the characteristics of a fluidized bed of Ottawa sand prior to making the experimental tests. The apparatus used for these observations was a transparent lucite plastic fluidization test unit constructed by Seidel(77). Since this unit was 2.75 inches in diameter as compared to the 3-inch standard diameter pipe used in the heat transfer investigation, the characteristics of a fluidized bed would be similar in both units under the same conditions.

Matheson(39) has indicated that in gas fluidized beds, particles in the range of 0.015-inch diameter show the best fluidization characteristics. A type of turbulent flow in which small gas bubbles rise through the fluidized bed was obtained with particles of this size. A similar type of fluidization was obtained in this investigation using particles of 0.026-inch average diameter. The range of mass air velocity for optimum fluidization was found to be 91.0 to 175.0 pounds per hour-square foot. At 66.0 pounds per hour-square foot, it was noted that only the upper surface of the bed showed any sign of motion. No movement was visible in the solids phase at the bed wall. The bed was not fluidized. At 90.0 pounds per hour-square foot, a slight agitation occurred all along the bed wall and about one small bubble per second could be seen bursting at the surface. At 109.0 pounds per hour-square foot mass velocity, the bed showed increased signs of agitation, and about three bubbles per second burst at the surface. At 123.0 pounds per hour-square foot, bubbles increased in size until they reached the diameter of the column in the upper inch of the bed. Solid particles were thrown as high as 1.5 inches above the top of the bed by bursting slugs. At 162.0 pounds per hour-square foot, slugging occurred in the upper half of the bed, throwing sand as high as three inches above the surface level. At 210.0 pounds per hour-square foot slugging occurred in the entire bed.

A mass superficial air velocity of 82.5 pounds per hour-square foot was selected because it corresponded to a bed condition occurring just prior to fluidization. Velocities of 123.2 and 170.3 pounds per hour-square foot were chosen because they occurred at intermediate points of fluidization. The highest velocity used, 217.5 pounds per hour-square foot, was selected because it caused a slugging condition; i. e., it was higher than the ideal fluidization range. The four mass velocities given above therefore should indicate the characteristics of the bed over the entire range of nonfluidized, fluidized, and slug flow conditions, and data could be obtained that would indicate the effect of each state on the over-all coefficient of heat transfer.

Conditions of Operation. Ottawa sand was chosen as the solid to be fluidized for an investigation of heat transfer for two reasons: first, because it had previously been used in similar studies under different conditions of temperature, mass flow rate, and particle size, and second, because its fluidization properties, such as air flow rate required for fluidization, and lack of solids entrainment in gas stream, were ideal for use with the apparatus(1) and facilities available.

The size of the fluidized bed was fixed by the unit that had been constructed; bed diameter was that of a three-inch nominal diameter pipe, and the heating section was 30 inches high. Air was chosen as the fluidizing medium because of its availability in the quantities required, and because the physical properties of the common gases have practically no effect on the fluidization characteristics, and no effect on the coefficient of heat transfer between the bed solids and bed wall. Air flow rates were selected as described in the preceding paragraph.

Bed wall temperatures of 200, 400, and 600 °F were used in the tests to evaluate the effects of temperature driving force on the coefficient of heat transfer.

Determination of Equilibrium Conditions. It was essential that steady-state conditions exist in the heat transfer test unit over the period of time when data for a test were being recorded. Preliminary temperature readings at significant points in the fluidization unit were taken at 20-minute intervals until steady-state conditions were indicated. These significant points included the bed wall at six-inch intervals from L-0 (lower level of heating element), to L-30 (corresponding to temperature of exit air), and the outer and inner temperatures of the insulation (as indicated by thermocouples 13 and 14, respectively). Constant readings by thermocouple number 13 located in the outer portion of the column insulation, constituted the most accurate indication of steady-state conditions, because in all tests, this thermocouple took longer to become constant than any of the other above-mentioned thermocouples. In each test, the mass air velocity and central, external bed wall temperature, as indicated by the "capacitrol" temperature controller, were held constant during the entire period of attaining equilibrium and recording of data.

Horizontal Bed Temperature Gradient. The horizontal bed temperature gradients at each bed level for each of the 15 tests may be determined from Table V. Several points of interest may be seen, as discussed below.

Effect of Bed Wall Temperature. An increase in horizontal bed temperature gradient occurred with an increase in bed wall temperature. At a constant mass air velocity of 82.5 pounds per hour-square foot, the average horizontal bed temperature gradient increased from 1 to 4 to 5 °F while a corresponding increase from 200 to 400 to 600 °F in bed wall temperature occurred. At 123.2 pounds per hour-square foot, the increase was from 0 to 4 to 5 °F. At 170.3 pounds per hour-square foot, the hotizontal temperature gradient increased from 0 to 1 to 6 °F, and the at the highest mass air velocity employed, 217.5 pounds per hour-square foor, the horizontal temperature gradient increased from 0 to 3 to 5 °F for wall temperatures of 200, 400, and 600 °F. It is of interest to note that even at the highest mass air velocity used, the highest horizontal temperature gradient obtained in this study was approximately equal to the lowest value obtained by Breckon(1) (4 °F). This fact serves to indicate the high quality of fluidization that existed in the tests of this investigation.

Effect of Mass Air Flow. No general statement may be made about the effect of mass air flow rate on the horizontal bed temperature gradient. The data obtained showed that the gradient fluctuated by about 2 °F over the range of mass flow rates at each of the controlled wall temperatures of 200, 400, and 600 °F. The gradient varied from 0 to 1 °F at 200 °F, 1 to 4 °F at 400 °F, and 5 to 6 °F at 600 °F. These facts indicated that the state of fluidization had very little effect on the horizontal temperature gradient.

Comparison of Empty to Fluidized Bed. Tests number 5, 10, and 15 were conducted under conditions similar to those of Tests 3, 8, and 13 (at 200, 400, and 600 °F and 175.0 pounds per hour-square foot), except that there were no bed solids present in the former. The average horizontal temperature gradients obtained with empty beds were 4, 9, and 19 °F, as compared with 0, 1, and 6 °F for fluidized beds at wall temperatures of 200, 400, and 600 °F. Thus it is shown that the intermixing of solids in the fluidized state reduces considerably the temperature gradient across the bed.

Temperature Drop Through Bed Wall. With reference to Table V, the calculated temperature drop through the bed wall varied from a minimum of 0.026 °F in Test 1 at 200 °F and 82.5 pounds per hour-square foot to a maximum of 0.364 °F in Test 14 at 600 °F and 217.5 pounds per hour-square foot. The temperature drop increased approximately in proportion with with mass air velocity at a given bed wall temperature and increased, although not linearly, with bed temperature at a constant mass air velocity. The bed wall temperature drop was calculated primarily to determine if the bed wall offered any appreciable resistance to heat flow from the heating element. It was shown conclusively that the resistance offered was negligible; in every case the temperature drop across the bed wall was less than one per cent of the total drop between the outer wall and fluidized bed.

Vertical Bed Temperature Gradient. The vertical bed and bed wall temperature profiles for each of the 15 tests are shown in Figure 4, page 75. The effects of mass air flow rate and bed wall temperature on the profiles can be noted by reading the curves from left to right and top to bottom, respectively. The curves for Tests 5, 10, and 15 correspond to those of Tests 3, 8, and 13, where conditions were similar, except that no bed solids were present in the former tests.

Effect of Bed Wall Temperature. Referring to Figure 4, page 75, and Table V, page 73, it may be seen that the curvature of the bed temperature profile increases with increasing bed wall temperatures. At 200 °F wall temperature, the average bed temperature increased from 159 °F at the L-0 level to 165 °F at the L-12 level and decreased to 163 °F at the L-30 level. For tests at 400 °F and 600 °F, the ranges were from 364 to 393 to 371 °F, and 546 to 575 to 532 °F, respectively. Thus the maximum vertical bed gradients were 6, 29, and 43 °F for temperatures of 200, 400, and 600 °F. It is immediately apparent that in the tests at 600 °F, the average temperature at the top of the bed, L-30, was lower than the temperature at the bottom, L-0. This would seem to indicate that the fluidizing medium, air, was being cooled at the top of the bed. It must be noted, however, that the temperatures reported are those of the bed solid, and not those of the gas passing through it. Leva(20) reported that a difference in temperature may exist between the solid particles and the gas phase, but that bare thermocouples, such as the ones used in this investigation, indicate only the temperature of the solids. This fact becomes immediately evident when the bed temperatures at the L-0 level are examined for the several tests conducted. Immediately below the L-0 level, air is being distributed to the bed at its entering temperature of approximately 80 °F; however, the bed temperatures indicated at the L-0 level are within 30 °F of the maximum temperature of the bed. It is a physical impossibility for the gas to increase in temperature by several hundred degrees at the instant it enters the bed. Thus it becomes logical that the temperature of the bed solids is affected more by some variable other than the mass velocity and temperature of the fluidizing medium than it is by these two factors. It is suggested, in view of the inverted U-shaped curves obtained for vertical temperature profiles (Figure 4) in this investigation, that the temperature of the solid particles is a function of the distance from the center level (<u>L-15</u>) of the heating element. This statement would be true only when the fluidization apparatus used was similar to the one employed in this investigation, in that the bed wall was insulated only over the 30-inch section containing the heating coil, and consisted of several feet of exposed metal surface above and below the heating unit. The bare metal provided a means for heat to escape to the atmosphere by radiation and conduction, thus creating a temperature gradient in the bed wall that extended from the top of the column to the <u>L-15</u> level and from the bottom of the column to the <u>L-15</u> level. This gradient was reflected in the bulk bed temperatures over the same interval in a similar manner.

The preceding paragraph is not to be construed in such a manner that it is thought that the fluidizing medium never reached the average bed temperature in its progress up through the bed. The gas reached bed temperature very rapidly, but the main point in the above discussion was that the gas does not assume the bed temperature on first contact with solid particles. In calculating the sensible heat added to the air stream, the increase in temperature of the air was taken as the difference between the entering air temperature of approximately 80 °F (instead of the L-0 bed level temperature) and the average bulk bed temperature at the L-30 level. Thus the air was assumed to reach the temperature of the bed at some point during its passage through the bed.

Heat Flow to the Fluidized Bed. The heat input to the fluidized bed was calculated by two methods. One, an indirect method, consisted of determining the total heat input by electric energy, and subtracting from this value the sum of the calculated heat losses by conduction along the bed wall and through the celite insulation. The other method, a direct means, was based on the increase in temperature of the air stream passing through the fluidized bed. The temperature increase of the air was multiplied by the product of the weight of air passing through the column and the average specific heat of air over the range of temperatures involved. The latter method was considered to be the more accurate; first, it was a direct determination, and second, the values of heat input by electric energy were at best good approximations, because of the extreme difficulty encountered in reading quantities as low as three- or four-tenths of a kilowatt-hour on the kilowatt-hour meter. The discrepancy between the two methods of calculation varied from 7.7 to 14.7 per cent of the heat flow calculated by the direct method, with the indirect method showing more heat flow in every test.

The effect of mass superficial air velocity on the rate of heat flow to the fluidized bed was linear for a given temperature, as shown in Figure 5.

Boundary Mean Temperature Difference Between Bed and Bed Wall. The boundary mean temperature differences, listed in Table V, were obtained by the method of graphical integration suggested by Leva(25). The areas between the vertical bed and wall temperature profile curves, plotted in Figure 4, were used to obtain the over-all temperature difference from the outer wall surface to the bed mass.

Table V shows that at a constant bed wall temperature, the boundary mean temperature difference increased with increased mass air velocity. At 200 °F, the temperature difference increased from 18.72 to 19.32 to 25.73 to 25.90 °F for increased in mass air rate from 82.5 to 123.2 to 170.3 to 217.5 pounds per hour-square foot. At 400 °F the increase in temperature difference was more pronounced, changing from 16.86 to 17.67 to 24.84 to 31.28 °F over the same range of mass velocities. At 600 °F, the temperature difference increased from 26.10 to 27.30 to 37.00 to 44.00 °F over same range of mass velocities as above. The major point worthy of note here is that at all three wall temperatures, the temperature difference increased between 82.5 and 123.2 pounds per hour-square foot mass velocities was very small (less than 2 °F) as compared with an average increase of 8 °F between mass velocities of 123.2 and 170.3 pounds per hour-square foot. The only variable in conditions of operation that can be related to the low increase in temperature differential between 82.5 and 123.2 pounds per hour-square foot is the state of fluidization of the bed. At 82.5 pounds per hour-square foot, the bed was not fluidized, while it was fluidized at 123.2 pounds per hour-square foot. From the definition of the term fluidization, it becomes evident that the conditions which would tend toward reduction of mean temperature difference exist in the fluidized state, but do not appear in the nonfluidized state. It should be noted, however, that there is no sudden change in conditions, or in the effect of these conditions, as shown by the fact that the rate of change of mean temperature difference is only slightly modified as the bed becomes fluidized.

Bed Wall Coefficient of Heat Transfer. The values of the bed wall coefficient of heat transfer obtained in this investigation are listed in Table V, and are plotted in Figure 6 as a function of mass superficial air velocity and of bed wall temperature.

Effect of Mass Superficial Air Velocity. The curves of Figure 6 indicate that the effect of mass superficial air velocity on the bed wall coefficient of heat transfer is the same at each of the bed wall temperatures used (200, 400, and 600 °F) in this investigation. At each bed wall temperature employed, the lowest coefficient of heat transfer occurred at the lowest mass air rate, 82.5 pounds per hour-square foot. The coefficient increased for the next higher mass velocity, 123.2 pounds per hour-square foot, and then remained essentially constant for the other two flow rates, 170.3 and 217.5 pounds per hour-square foot. It was to be expected that the coefficient of heat transfer would be lowest at 82.5 pounds per hour-square foot, because it had previously been shown (Table III) that the minimum mass air velocity for fluidization of the bed material was 91.0 pounds per hour-square foot. At velocities lower than that required for fluidization, a layer of stationary solid particles existed at the bed wall (as observed in Seidel(77) apparatus) as opposed to a state of continuous agitation at the bed wall in the fluidized condition. Thus the heat being transferred through the column wall to the bed would have to pass through this layer of stationary particles before being transmitted to the turbulent mass of particles that constituted the bed mass. Therefore the layer of particles would offer a resistance to heat flow not encountered in the fluidized state, where heat would be transmitted to the bed by movement of heated particles, as well as by conduction through the particles. Thus the coefficient of heat transfer would be lower at a mass air velocity below that required for fluidization than at a velocity in the fluidization range.

The fact that the coefficient of heat transfer is affected very little by mass velocities above the minimum required for fluidization has been shown by Leva(23). In this investigation it was shown

(Figure 6) that once the bed was fluidized, the coefficient of heat transfer varied by less than four per cent over the range of mass velocities employed at a constant bed wall temperature.

Effect of Integrated Mean Temperature Difference. The curves of Figure 6 indicate that the coefficient of heat transfer was approximately 0.5 Btu per hour-square foot-°F greater at a bed wall temperature of 600 °F than at 400 °F, as compared with an average increase of about 8 Btu per hour-square foot-°F between bed wall temperatures of 200 and 400 °F. In Table V, the data are listed from which the coefficient was calculated, and it may be seen that the mean temperature difference used in the calculation was the factor responsible for the smaller increase in coefficients at 600 °F. The relationship used in the calculation was as follows:

$$h = \frac{q}{A \Delta t_m}$$

where

h = coefficient of heat transfer, Btu/hr-sq ft-°F

A = inner bed wall area, sq ft

q = rate of heat flux, Btu/hr

Δt_m = integrated mean temperature difference between bed wall and bulk bed, °F.

The factor that determined the coefficient is therefore the relation, $\frac{q}{\Delta t_m}$, as $\underline{A}$ is a constant and does not affect the relative values obtained from the equation. The value of $\underline{\Delta t_m}$ increases with $\underline{q}$ between bed wall temperatures of 200 °F and 400 °F so slightly that the value of $\frac{q}{\Delta t_m}$ is affected primarily by the relative change in $\underline{q}$ alone, which showed an increase. Between bed wall temperatures of 400 and 600 °F, however, $\underline{\Delta t_m}$ showed a marked increase; in fact its relative increase over the 400-600 °F range was almost as great as that exhibited by $\underline{q}$ over the same range. Thus the value of $\frac{q}{\Delta t_m}$, and consequently, the coefficient of heat transfer, increased only slightly at a bed wall temperature above 400 °F.

Comparison of Coefficient of Heat Transfer Between Empty and Fluidized Beds. For purposes of comparison, tests number 5, 10, and 15 were made under the same conditions as tests number 3, 8, and 13 (200, 400, and 600 °F at 170 pounds per hour-square foot), except that no bed solids were present in the former tests. Thus the value of the presence of fluidized solids can be effectively compared with an empty reactor in terms of heat transfer coefficients. Test 5, at 200 °F, resulted in a heat transfer coefficient of 1.33 Btu per hour-square foot-°F, as compared with 3.54 Btu per hour-square foot-°F for test 3 with solids present. The solid particles showed even greater effectiveness at higher temperatures. Values of 2.30 and 12.45 Btu per hour-square foot-°F were obtained at 400 °F for empty and fluidized beds, respectively. At 600 °F, the coefficient for an empty bed was 2.57 Btu per hour-square foot-°F as compared with 13.31 Btu per hour-square foot-°F for a bed with solids present.

Summary of Discussion. The results of this investigation on the coefficient of heat transfer to a fluidized bed were four-fold: (1) the coefficient of heat transfer is about 25 per cent greater in a fluidized bed than in a nonfluidized bed of the same solid particles;

(2) the coefficient of heat transfer is increased an average of four times by the presence of fluidized solids as compared with an empty reactor; (3) the coefficient of heat transfer increases with increasing bed wall temperature, but the rate of increase decreases as higher bed wall temperatures are used; (4) when mass air velocity has reached the minimum rate for fluidization, the coefficient of heat transfer is not affected by further increases in mass air flow rate; i. e., slugging has no effect on the coefficient of heat transfer.

Recommendations

The recommendations evolving from the observations made during this investigation of heat transfer to a fluidized bed of Ottawa sand include variation of operating conditions and equipment modifications.

Variation of Air Velocities. It was believed that the maximum mass air velocity for good fluidization, (175.0 pounds per hour-square foot) was employed in this investigation. However, good fluidization of Ottawa sand was found to include a wide range of mass air velocities; from 91.0 to 175.0 pounds per hour-square foot. Only two of these (123.2 and 170.3) pounds per hour-square foot) are herein reported. It is recommended that future work be done using velocities of 100 and 150 pounds per hour-square foot. In addition, special emphasis should be put on mass flow rates within 10 pounds per hour-square foot of 90 pounds per hour-square foot, the minimum velocity necessary for fluidization.

Variation of Wall Temperature. It is strongly recommended that wall temperatures of 800 and 1000 °F be employed in future investigations. These are the temperatures most often employed in catalytic operations, and heat transfer coefficients in the 1000 °F range would be of more value than those obtained at lower temperatures. It would be necessary, however, to enlarge the present heating capacity by removing the old unit and replacing it with a unit consisting of more coils or with a wire offering greater electrical resistance, because the present upper temperature limit of the heating unit is at about 700 °F.

Use of Ottawa Sand as Bed Solid. It is recommended that the use of Ottawa sand be continued in future investigation. The use of this material makes possible the employment of the simplest form of apparatus and eliminates several hard-to-control variables from consideration. Some of these are: separation of entrained particles from exit air stream, method of introducing fresh or recycled solids to the unit, heat correction for cold solids added, and height of column in recycle standpipe necessary to feed solids to unit.

Installation of Additional Thermocouples. Thermocouples should be installed at two-inch intervals for six inches above and below the heating unit in the column wall in order to determine more accurately the heat losses by conduction in a direction parallel to the column axis. A thermocouple should also be installed in the entering air line just below the air distribution plate, so that the inlet air temperature could be better evaluated, and used to give a more accurate calculation of heat flux to the air stream. Additional thermocouples should be placed in the insulation at six-inch vertical intervals starting from the bottom of the heating unit, in order that a better evaluation of heat loss by conduction in a direction perpendicular to the column axis could be made.

Installation of Additional Insulation. The present insulation, which now covers only the 30-inch heating section should be extended at least one foot above and below its present position. This change would enable a more accurate determination of a heat balance over the unit, and would tend to smooth out the vertical temperature profiles obtained in this investigation.

Comparison of Data Between Different Fluidization Units. A transparent lucite plastic fluidization test unit was used in this investigation to make possible the observation of fluidization characteristics of the bed material. The phenomena thus observed were assumed to be identical with those occurring in the metal column of the fluidized bed heat transfer apparatus. It is quite possible, however, that the data obtained in the lucite column could not be applied directly to a unit constructed of another material. The walls of the lucite unit were smooth and even, while the metal column walls displayed a slight surface roughness. Therefore, it seems likely that the critical points in the fluidization range (onset of fluidization and beginning of slugging conditions) would occur at somewhat lower velocities in the plastic unit than in the unit with metal walls. Also, in the heat transfer apparatus used, a $\frac{1}{4}$-inch pipe was used as a thermocouple traversing rod in the center axis of the fluidized bed. The presence of this rod could conceivably have reduced the formation of slugs within the bed, and thus increased the velocity required for slugging throughout the bed.

It is recommended that future investigations include a study of the various effects described above, together with an evaluation of the reliability of transferring data obtained in the operation of one unit to the operation of another unit.

Data Necessary for Correlation of Variables. Insufficient data were available from this investigation to correlate the coefficient of heat transfer with the variables encountered in operation of a fluidization unit. The most complete correlations yet developed have been those by Leva(20), in which the concept of efficiency of fluidization, E , was an important factor. Fluidization efficiency is determined by the following factors: composite diameter of bed and its composition, density of bed material, shape of particles, kinematic viscosity, and rate of flow of fluidizing medium through the bed. The additional information that would be necessary for calculation of efficiency of fluidization in the tests of this investigation would be the kinematic viscosity of the fluidized bed over the range of mass velocities employed.

It is recommended that data be taken on the kinematic viscosity of the bed material used in future investigations. The necessary measurements could be made with a stormer-type viscosimeter inserted in the top of a fluidization test unit, as described by Matheson(41).

Determination of Effect of Temperature on Heat Transfer Coefficient. From the tests made in this investigation, it may be seen that bed-wall temperature has a pronounced effect on the coefficient of heat transfer. The coefficient increases with bed-wall temperature, but the eate of increase decreases with increasing temperature. It may be noted that the Dittus-Boelter equation for film coefficient of heat transfer by conduction contains no term that is directly affected by temperature. Yet, as stated above, temperature does have an effect on the heat transfer coefficient in fluidized beds.

It has been shown in this investigation that the bed solids are heated by the bed wall, and that the fluidizing medium, air, is heated by the solids. It is thought that the mechanism of the heat transfer is as follows: The fluidized solids are heated by radiation from the bed wall; this would account for the variation of heat transfer coefficient with temperature, as radiation is a function of the fourth power of temperature. The air stream receives its heat by conduction from the solids, and thus is affected very little by temperature. It should also be noted that at equilibrium conditions, heat flow to the air from the solids must equal heat flow to the solids from the bed wall. At the lower temperatures, in the range from 200 to 400 °F, the heat transfer by radiation is controlling, and the bed-wall coefficient of heat transfer increases with increasing temperature. At the higher temperatures (400 to 600 °F), however, heat transfer by conduction is the limiting factor, and the coefficient of heat transfer remains relatively constant for further increases in temperature.

It is recommended that bed solids of varying degrees of blackness be used in future investigations, in order to determine if radiation is a significant factor in heat transfer to a bed of fluidized solids.

Limitations

This investigation was conducted with the following limitations:

Fluidization Chamber. The fluidization chamber used was constructed from a five-foot length of three-inch nominal diameter black iron pipe of 3.068-inch inside diameter. The central 30-inch section constituted the heating section.

Fluidized Solid. The material employed as the fluidized solid was rounded Ottawa sand (20/30 U. S. Standard sieve) with an average particle diameter of 0.026-inch.

Fluidizing Medium. Air at approximately 80 °F and containing 0.01 pound water vapor per pound dry air was used as the fluidizing medium.

Air Flow Rates. Mass superficial air velocities of 82.5, 120.3, 170.3, and 217.5 pounds per hour-square foot were employed.

Wall Temperatures. Bed wall temperatures at the control point of 200, 400, and 600 °F were employed.

V. CONCLUSIONS

An evaluation of the results of the study of the heat transfer characteristics of a fluidized bed of 0.026-inch average diameter Ottawa sand and air in a 3.068-inch inside diameter fluidization column for a 30-inch externally heated vertical section, where wall temperatures of 200, 400, and 600 °F and average mass superficial air velocities of 82.5, 123.2, 170.3, and 217.5 pounds per hour-square foot were the variables, led to the following conclusions:

1. The range of mass superficial air velocities during which the bed material, Ottawa sand, was fluidized, but did not slug throughout its entirety, was 91.0 to 210.0 pounds per hour-square foot (as determined in a transparent plastic fluidization test unit); thus fluidization was not attained in the three tests conducted at 86.3, 82.7, and 79.4 pounds per hour-square foot.

2. The horizontal bed temperature gradient increased with increasing bed wall temperature. For bed wall temperatures of 200, 400, and 600 °F, the gradient increased from 1 to 4 to 5 °F, from 0 to 4 to 5 °F, from 0 to 1 to 6 °F, and from 0 to 3 to 5 °F for mass superficial air

velocities of 82.5, 123.2, 170.3, and 217.5 pounds per hour-square foot, respectively.

3. The temperature drop across the bed wall increased with increasing bed wall temperature, from an average of 0.049 $^{\circ}$F, at 200 $^{\circ}$F wall temperature to 0.169 $^{\circ}$F and 0.263 $^{\circ}$F at wall temperatures of 400 and 600 $^{\circ}$F, respectively.

4. The temperature drop across the bed wall increased with increasing mass air velocity, from an average of 0.092 $^{\circ}$F at 82.5 pounds per hour-square foot to 0.170, 0.185, and 0.227 $^{\circ}$F for mass velocities of 123.2, 170.3, and 217.5 pounds per hour-square foot.

5. The heat flow to the fluidized bed at constant bed wall temperature increased with increased mass air velocity, increasing at 200 $^{\circ}$F from 84 to 220 Btu per hour, from 311 to 785 Btu per hour at 400 $^{\circ}$F, and from 500 to 1172 Btu per hour at 600 $^{\circ}$F, at mass air flow rates from 82.5 to 217.5 pounds per hour-square foot, respectively.

6. The boundary mean temperature difference at constant bed wall temperature increased with increasing mass air velocity, reaching maximum of 44.00 $^{\circ}$F at a bed wall temperature of 600 $^{\circ}$F and mass air velocity of 217.5 pounds per hour square foot. The minimum difference was

16.86 °F, occurring at 400 °F and 82.5 pounds per hour-square foot.

7. The boundary coefficient of heat transfer increased at constant mass air velocity from a minimum at 200 °F to a maximum at 600 °F. At a constant bed wall temperature of 200 °F, the coefficient of heat transfer reached a maximum of 3.78 Btu per hour-square foot-°F at a mass velocity of 217.5 pounds per hour-square foot. At 400 °F a maximum of 12.91 Btu per hour-square foot-°F was reached at a superficial mass air velocity of 120.3 pounds per hour-square foot. At 600 °F, a maximum coefficient of 13.40 Btu per hour-square foot-°F was obtained at 123.2 pounds per hour-square foot. At each bed wall temperature, the minimum coefficient of heat transfer was found at 82.5 pounds per hour-square foot.

8. The boundary coefficient of heat transfer was increased 2.66-fold at 200 °F, 5.42-fold at 400 °F, and 5.18-fold at 600 °F by the presence of fluidized solids as compared with an empty column, at a mass superficial air velocity of 170.3 pounds per hour-square foot.

VI. SUMMARY

Fluidization, a relatively new unit operation by which a solid and gas or liquid may be contacted, is being widely developed in the field of catalytic cracking of petroleum because of its characteristic reduction of temperature gradients within the reaction mass. Basic research of the heat transfer properties of fluidized systems has lagged far behind industrial applications. It was the purpose of this investigation to evaluate the effect of temperature driving force and mass superficial air velocity on the coefficient of heat transfer at the bed wall of an externally heated, fluidized bed of ottawa sand at wall temperatures of 200, 400, and 600 °F, and average mass superficial air velocities of 82.5, 123.2, 170.3, and 217.5 pounds per hour-square foot.

The tests were carried out under steady state conditions of air flow and bed wall temperature. A complete heat and material balance, including evaluation of heat losses, was made for each test. The boundary coefficients based on the internal area and temperature of the pipe wall were calculated. The effects of mass superficial air velocity and wall temperature on the boun-

dary coefficient of heat transfer and on bed and bed wall temperature gradients were studied.

From observations made it was noted that the fluidization range of the Ottawa sand bed began at a mass superficial air velocity of 91.0 pounds per hour-square foot and ended at 210.0 pounds per hour-square foot, the velocity at which slugging occurred throughout the bed.

The horizontal temperature gradient across the bed increased with increasing bed wall temperature, increasing from a minimum of 0 °F at 200 °F wall temperature to 6 °F at 600 °F. The rate of heat flux to the air stream passing through the fluidized bed increased with mass air flow rate at constant bed wall temperature. The minimum heat flux was 84 Btu per hour and occurred at 200 °F and 82.5 pounds per hour-square foot, while the maximum was 1172 Btu per hour and occurred at 600 °F and 217.5 pounds per hour-square foot. The coefficient of heat transfer increased with bed wall temperature, reaching maximum values of 9.55, 13.40, 13.31, and 13.30 Btu per hour-square foot-°F at 600 °F and at mass superficial air velocities of 82.5, 123.2, 170.3, and 217.5 pounds per hour-square foot, respectively.

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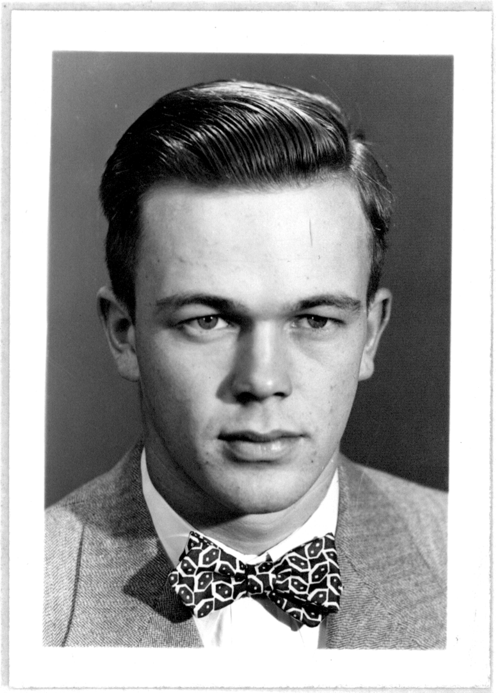
IX. VITA



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