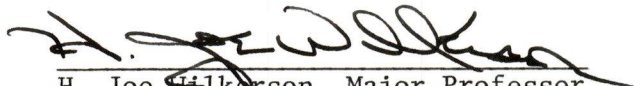
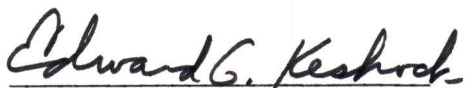


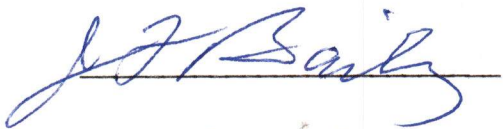
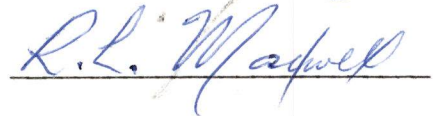
To the Graduate Council:

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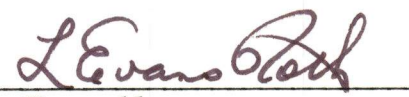

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THE ANALYSIS AND THE CONCEPTUAL DESIGN OF AN EXPERIMENTAL
APPARATUS FOR THE DETERMINATION OF PRESSURE
DROP ACROSS HEAT TRANSFER SURFACES IN
A FLUIDIZED BED

A Thesis
Presented for the
Master of Science
Degree
The University of Tennessee, Knoxville

Ganiyu B. Awelenje

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ABSTRACT

An analysis of the pressure drop across a fluidized bed with immersed heat transfer surfaces was performed. The analysis was for an isothermal bed in which no net transport of particles occurs. A momentum equation was written for a control volume of the bed and the effects of the weight and the drag forces of the fluidizing gas on the tubes, and the containing walls were evaluated. The analysis included considerations of the effective viscosity for the fluidized beds.

The conceptual design of an experimental apparatus for the measurement of the pressure drop in a fluidized bed was also completed. The bed is rectangular in shape (2 ft x 1 ft) with two opposite sides of plywood and two sides of plexiglass to permit visual observation. The heat transfer tubes are horizontal and are sealed at the end to avoid leakage. The distributor plate consists of 288 nozzle buttons made from standard size bolts.

Sample calculations are presented for the total pressure drop using 1020 microns (0.040 in.) diameter glass bead particles in the bed with $\frac{1}{2}$ in. diameter heat transfer tubes using air as the fluidizing gas. Both in-line and staggered tube arrangements are considered. For a typical geometry considered, configuration 1 (tubes on 1 in. centers, 17 rows), the total pressure drop was 18.45 ins. of water with the bed weight contributing 99.3 percent and the tube drag and wall drag contributing 0.8 percent and 0.01 percent respectively.

The effect of pressure drop on larger heat transfer tube diameters was also evaluated. With $1\frac{1}{2}$ in. diameter tubes, a typical total pressure drop of 18.43 ins. of water was obtained for configuration 7 (tubes on 2 in. centers, 9 rows). The bed weight contributed 99.5 percent and the tube drag and wall drag contributed 0.5 percent and 0.011 percent of the total pressure drop.

This analysis showed that the pressure drop due to the drag on the tubes and walls did not contribute significantly to the total pressure drop with immersed heat transfer surfaces.

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LIST OF SYMBOLS

A_B, A_{Bed}	Area of bed
A_C	Surface area of heat transfer tube
B	Breadth or width of the rectangular bed
$C_{D_{air}}, C_{D_{gas}}$	Drag coefficient of air or gas
$C_{D_{F.B}}$	Drag coefficient of the fluidized bed
C_f	Skin friction
C_{f_f}	Skin friction of fluid
$C_{f_{F.B}}$	Skin friction of fluidized bed
cv	Control volume
cs	Control surface
D_c, D_t	Diameter of the heat transfer tube
D_p, D_s	Diameter of the particle or solid
F	Fore
F_D	Drag Force
f	friction factor
f^1	Empirical friction factor
g	Local gravity constant, $32.2^{ft}/sec^2$
g_c	Conversion factor, $32.2 \text{ lbm-ft/lbf-sec}^2$
G_{mf}	Minimum fluidization mass velocity
G_{max}	Maximum mass velocity at minimum flow area
H	Height of fluidized bed grid or expanded bed
H_{mf}	Bed height at minimum superficial gas velocity

h_L	Head loss
L	Length of rectangular bed
L_t	Length of heat transfer tube
L_{sh}	Slug height
M_s	Mass of solid
M_f, m_{air}	Mass of fluid or air
N	Number of heat transfer tubes or bolts
P	Pressure
$\Delta P_{w/a}$	Pressure drop due to bed weight per unit area
$\Delta P_{wall\ shear}$	Pressure drop due to wall shear
ΔP_{tube}	Pressure drop due to heat transfer tube
ΔP_{total}	Total pressure drop
S_n	Dimension normal to flow
S_p	Dimension parallel to flow
u	Velocity
V, v	Velocity
V	Volume
V_{mf}	Minimum fluidization velocity
V_{max}	Maximum velocity at minimum flow area
V_T	Terminal velocity
V	Superficial velocity
V_{total}	Total volume
V_{solid}	Volume of solid
W_B, W_{Bed}	Weight of the bed

ρ	Density
ρ_s, ρ	Density of solid
ρ_f	Density of fluid
μ	Absolute viscosity
μ_f	Absolute viscosity of fluid
$\mu_{F.B}$	Effective viscosity of fluidized bed
ν	Kinematre viscosity
τ_w	Wall shear stress
$\tau_{w_{F.B}}$	Wall shear stress in the fluidized bed
ϵ	Void fraction

CHAPTER I

INTRODUCTION

In order to control the pollutant emissions and maintain a clean environment, the Environmental Protection Agency (EPA) issued regulations restricting the qualities of pollutant emissions from various utility plants. For 2.6 percent sulfur coal combustion, 68 percent SO_2 removal would be required to meet the EPA emission standard of 1.2 lb of SO_2 per million BTU of heating value (16). The EPA standard for NO_x emission is 0.7 lb of NO_2 per million BTU of heating value (16).

Controlling the emissions from the stacks of combustion systems can be extremely costly and may introduce additional environmental problems. These polluting problems may be difficult to prevent economically by treating the products of combustion with the scrubbers designed to remove the polluting constituents after they have been formed. To be able to meet the EPA requirements, there is a necessity for the development of new techniques that are more economical to handle or control the pollutant emissions.

A very promising technique that may achieve this goal economically is the direct combustion of coal in fluidized bed of limestone. This technique controls the SO_2 and NO_x emissions below the limits established by the EPA. The fluidized bed combustion technique permits the design of smaller, more economical steam generator for coal and lignite combustion (18).

A fluidized bed can be defined as a bed of granular particles supported by a non-sifting grid through which an upward flow of gas is passed with sufficient velocity to create fluidization. In this thesis, a rectangular bed is considered for easy construction of the tubes in the bed. The fluid used for fluidization is combustion air.

Fluidized bed combustion process involves the combustion of coal in a fluidized bed of limestone which is maintained at 1500 - 1600° F. R. C. Hoxe et al. (16) reported that within 1500 to 1700° F, the limestone material in the bed absorbs up to 90 percent of the SO₂ released during coal combustion (see Figure 1). This shows the advantage of fluidized bed of limestone because it controls the SO₂ emission below the limits established by the EPA. In the same report (16), it was shown that within normal operating excess air of 15 %, and a temperature of 1460 - 1700° F, NO_x emissions are on the order of 0.2 to 0.3 lb NO₂/10⁶ BTU, well below the EPA standard (see Figure 2).

Most of the research on fluidized bed coal combustion was carried out at the Argonne National Laboratory, Esso Research and Engineering Company, National Coal Board, U. S. Bureau of Mines, and Westinghouse Research Laboratory. These studies were carried out primarily at an atmospheric pressure. In most of these studies, the pressure drop in the boiler tubes is usually assumed to be entirely due to static head of gas-solid mixture. These assumptions are not correct because the friction loss between the solid particles and the tube walls are not considered and these affects may have a significant contribution, particularly as more heat transfer surface is added to the bed.

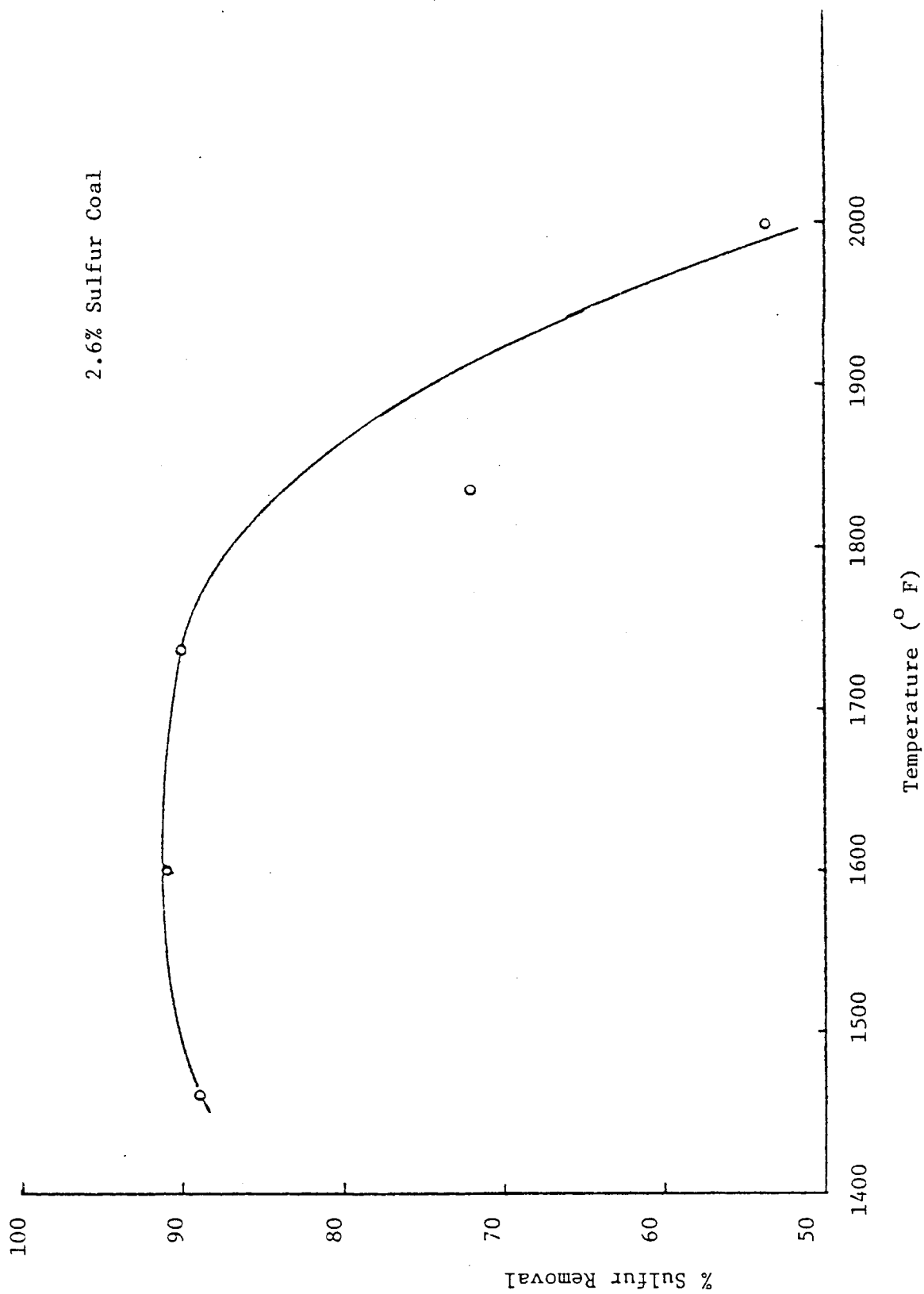


Figure 1. Effect of temperature on sulfurization in a fluidized bed (16).

2.6 % Sulfur Coal

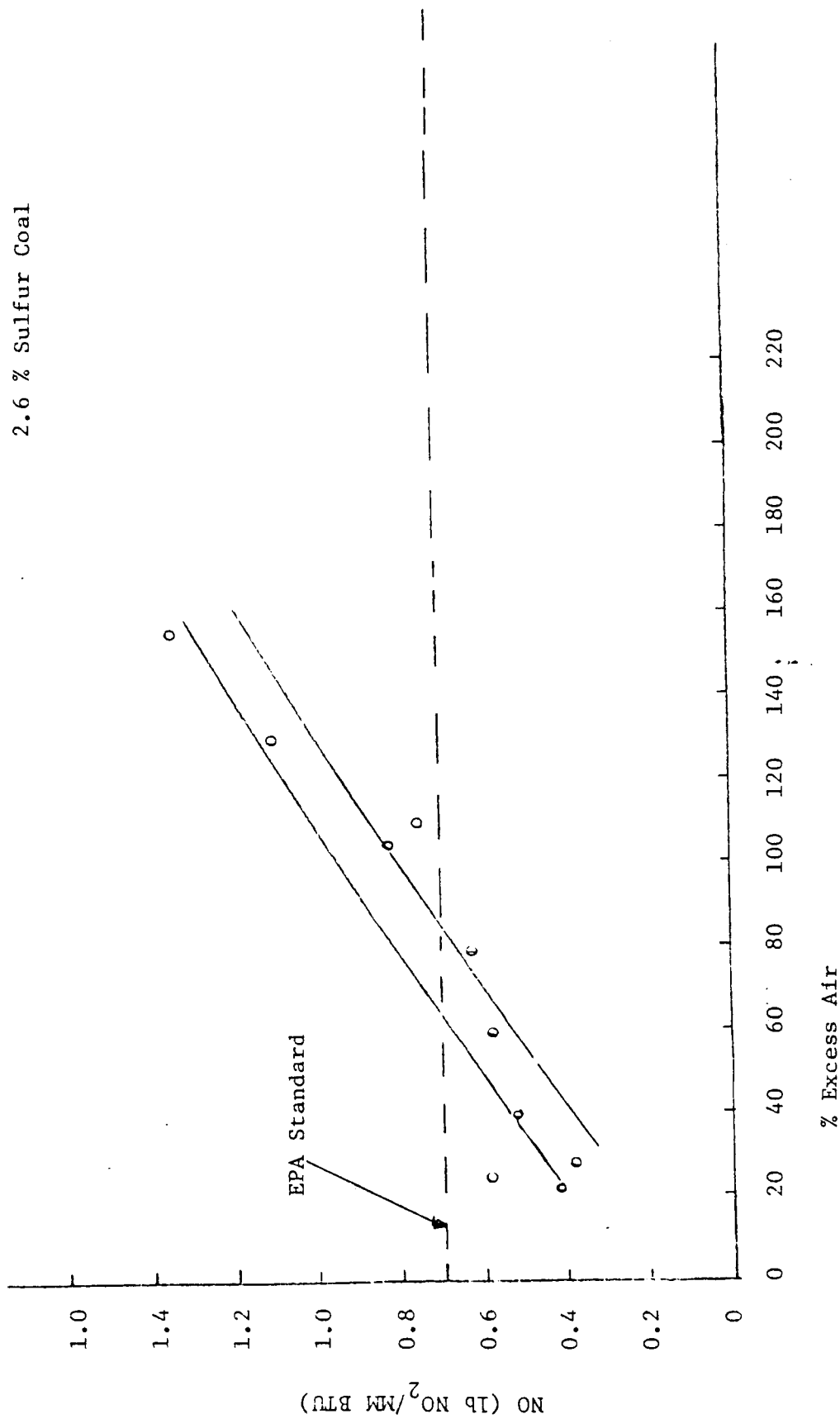


Figure 2. Effect of excess air on NO_x emissions in a fluidized bed (16).

I. REVIEW OF LITERATURE

From the literature review it was observed that most of the information on pressure drop in fluidized bed is based on the pneumatic transport of solids. Mehta et al. (9) described the pressure drop and particle velocity in air-solid flow systems experimentally for two particle sizes. Hariu and Molstad (8) presents the measurement of total pressure drop in transport of solid particles through vertical 0.267 and 0.532 in. inside diameter glass tubes by an air stream. Estimation of pressure drop for vertical pneumatic transport of solids were done experimentally by Jones et al. (4). Belden and Kassel (5) presented a paper on the pressure drop encountered in conveying particles of large diameter in vertical transfer lines. The concept of pneumatic transport of solids will be modified to estimate the pressure drop analytically for the immersed tube in a rectangular bed.

In this paper, the design of the experimental apparatus to measure the pressure drop and analytical analysis of same is given. In the analysis, the effect of the random motion of the particles, weight of the bed and the drag of the air on the tubes are considered in evaluating the pressure drop.

II. HISTORY AND APPLICATION OF FLUIDIZED BED

The first fluidized bed unit to operate on a commercial scale was the Winkler Gas Generator. This was developed in Germany in 1921. This generator was used to manufacture water and producer gas from finely divided coal.

Fluidized beds have gained many successes in the chemical industries because of the ease of solids handling and the free surface of the fluidized particles. The greatest number of fluidized applications in chemical industries involve chemical reactions and catalytic conversion of petroleum hydrocarbons. Frantz (10) outlined most of the fluidized applications in chemical industries. A list of some of the applications are given below:

<u>Process</u>	<u>Primary Solid Function</u>
Catalytic cracking	Catalyst
Thermal cracking of crude oil	Heat transfer medium
Fluid Coking	Heat transfer medium
Manufacture of sulfur trioxide	Reactant
Manufacture of cement	reactant

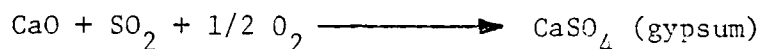
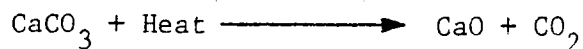
Combustion of Coal

Because fluidized bed techniques reduce the emission of SO_2 and NO_x from the burning of sulfur containing coals and have high rates of heat transfer, they are attractive for an application as fluidized bed combustion boilers. In some of the fluidized bed combustion of coal (16), (17), (24), limestone is used as the bed material.

Coal combustion fluidized beds require the use of other fuels during start up operation. Among these are propane, petroleum and natural gas. In case of propane, it is ignited by a spark plug and passed through the static bed. The bed material is preheated to a temperature of $800 - 1000^{\circ}\text{F}$ which is above the ignition temperature of coal. At this stage, the coal is injected into the system in a particulate form and the bed is maintained at an operating temperature of $1500 - 1600^{\circ}\text{F}$ ($815 - 870^{\circ}\text{C}$) (17). These conditions result in absorption by the limestone bed material of up to 90 percent of SO_2 released during coal combustion, keeping emissions within present EPA limits for utility units firing coals with up to 7 percent sulfur content. Because of low combustion temperature, NO_x formation is minimized. Introduction and combustion of more coal in the bed results in high temperature of the bed. This high temperature is controlled by the heat transfer surfaces immersed in the bed (see Figure 3).

Reactions in the Bed

With limestone as the bed material, the mechanism of SO_2 captured involves the calcination of calcium carbonate (CaCO_3) to lime (CaO) and carbon dioxide (CO_2) and the subsequent absorption of SO_2 by lime as indicated in the following reaction (17):



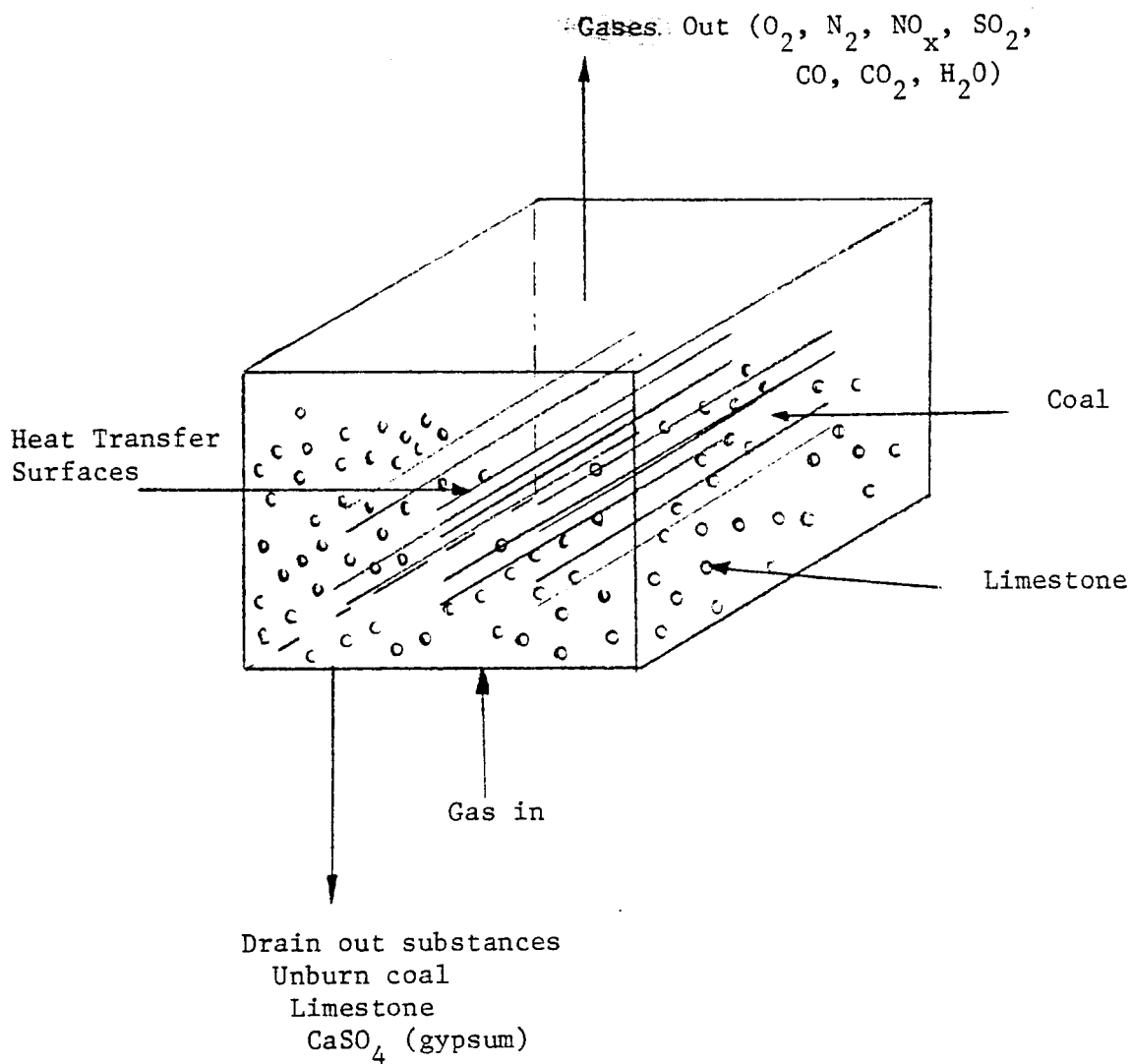


Figure 3. Combustion System.

The solid CaSO_4 , unburned coal and limestone are removed from the bed (see Figure 3). The gypsum is disposed of, offered for resale (serves as a fertilizer or road aggregate) or regenerated.

As fresh limestone is added, the Ca/S ratio is maintained so as to keep SO_2 emission within EPA standard.

Mechanical Applications of Fluidization

One of the mechanical applications of fluidization is shown in Figure 4. The figure consists of a porous inclined slab that carries a layer of powder to be conveyed to the other end. Air is admitted as shown in the figure and the porous slab acts as the air distributor through which the powder is put into a minimum fluidization state. At this state, certain degrees of freedom to slide down the slab are imparted to the overlying solids. By adjusting the air flow, the velocity of the solids can be controlled.

The combinations of a fluidized bed and a proper stirrer may be an effective solids blender (3). Because a bed of solid particles exhibit a liquid property at the onset of fluidization, such a bed may be stirred very much as a liquid is. This property has been used for the development of the air-activated blender shown in Figure 5. A small air stream is introduced into the base of the conically shaped bed of solid particles. A blade stirrer, with blades arranged in a certain fashion and coordinated with the sense of rotation, agitates the low portion of the bed. By continuous stirring, the solids from the upper part of the bed are circulated and mixed.

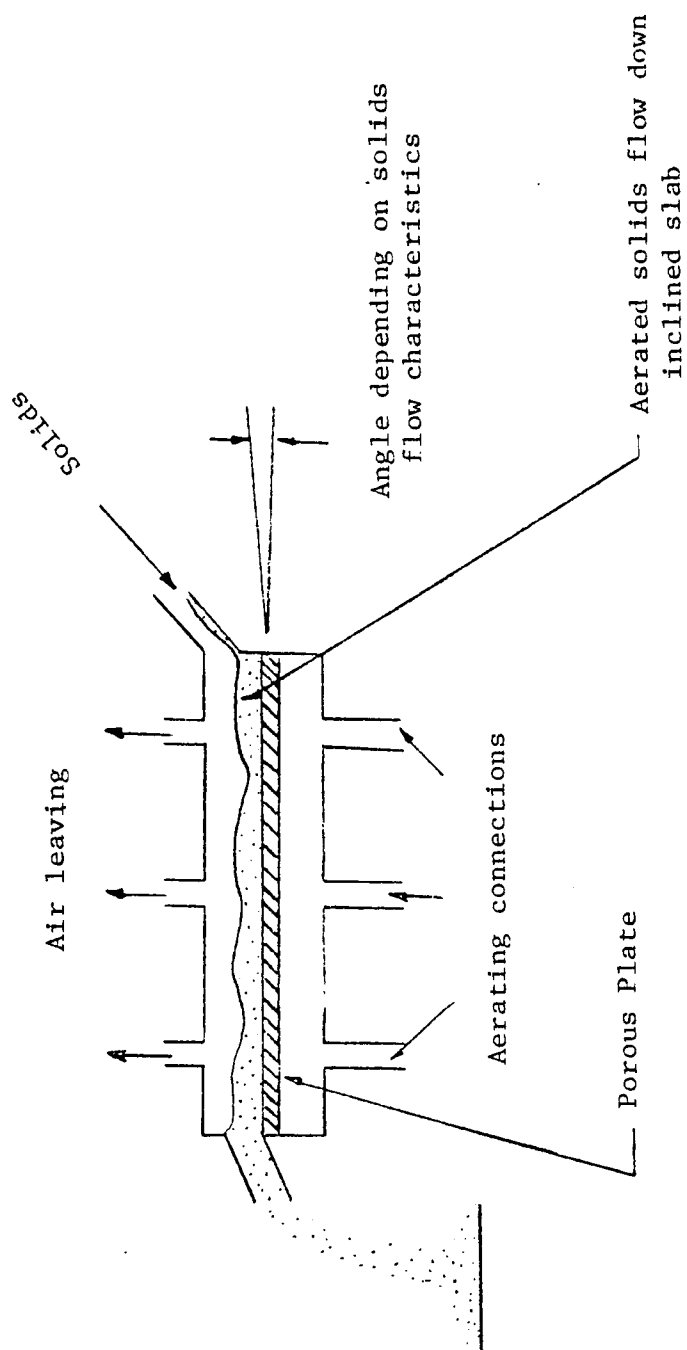


Figure 4. Airslide solids conveyor (3).

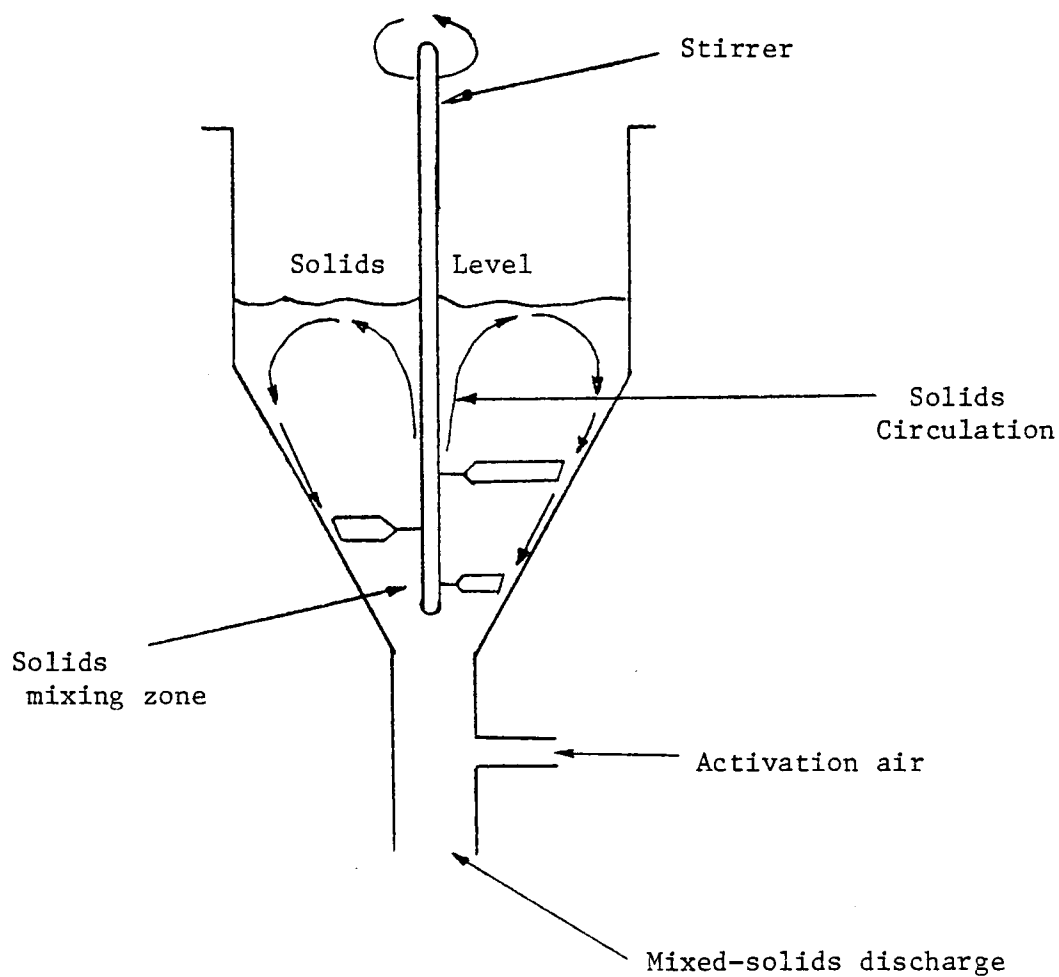


Figure 5. Air-activated solids blender (3).

CHAPTER II

ANALYTICAL DETERMINATION OF PRESSURE DROP IN A FLUIDIZED BED OF SOLID PARTICLE WITH HEAT TRANSFER SURFACES

The geometry of the bed considered is rectangular. In order to estimate the pressure drop the momentum equation is written for a control volume of a differential size of the bed, Figure 6.

Writing the momentum equation on the control volume.

$$\Sigma F = \frac{d}{dt} (mV)_{cv} + \int_{cs} V (\rho V \cdot dA) \quad (1)$$

where $(mV)_{cv}$ is the instantaneous momentum within the control volume. But the mass contained in the control volume is the density-volume product giving

$$\frac{d}{dt} (mV)_{cv} = \frac{d}{dt} \int_{cv} \rho V dV \quad (2)$$

Putting (2) in (1)

$$\Sigma F = \frac{d}{dt} \int_{cv} \rho V dV + \int_{cs} V (\rho V \cdot dA) \quad (3)$$

Since there is no flow of gas in x and z direction, the momentum equation reduce to one-dimensional. The y-direction momentum equation is:

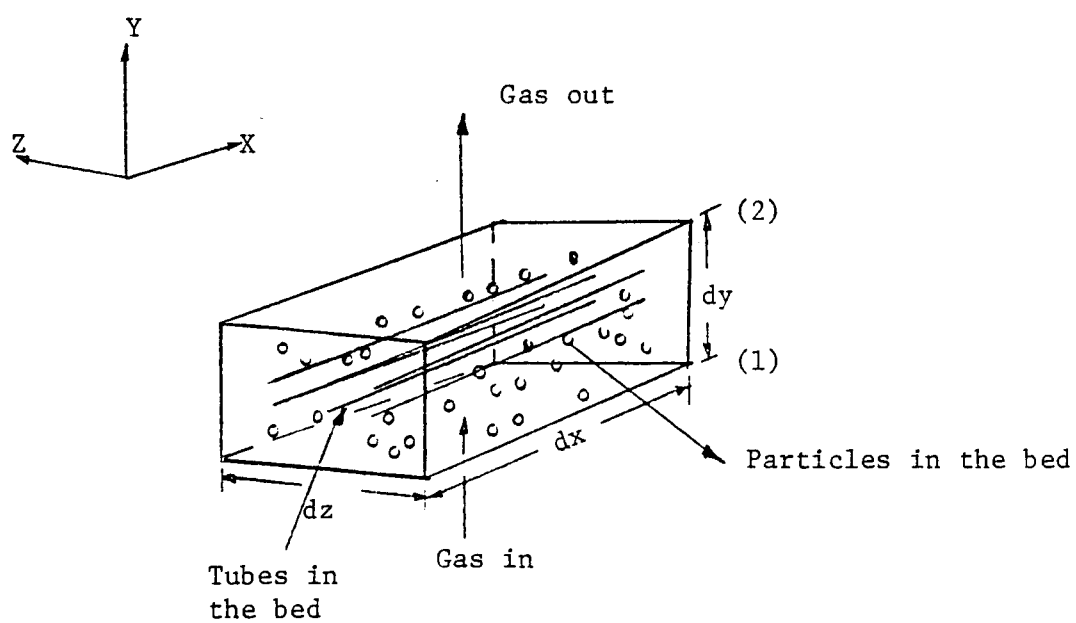


Figure 6. Control volume of a differential size of the bed.

$$\Sigma F_y = \frac{d}{df} \int_{cv} \rho V_y dV + \int_{cs} V_y (\rho V \cdot dA) \quad (4)$$

Where the change in momentum inside the control volume is zero because of steady flow. Also since the density has uniform values over area A_1 and A_2 , equation (4) can be rewritten as:

$$\begin{aligned} \Sigma F_y &= \rho_1 \int_{cs_1} V_y (V \cdot dA) + \rho_2 \int_{cs_2} V_y (V \cdot dA) \\ &= \rho_1 V_1 (-V_1 A_1) + \rho_2 V_2 (V_2 A_2) \end{aligned}$$

where V_1, V_2 are the components of V in y -direction. Uniform velocities are considered. Hence,

$$\Sigma F_y = (\rho_2 V_2 A_2) V_2 - (\rho_1 V_1 A_1) V_1 = \frac{\dot{m}_2 V_2}{g_c} - \frac{\dot{m}_1 V_1}{g_c} \quad (5)$$

$$\text{But } \Sigma F_y = P_1 A_1 - P_2 A_2 - \tau_w A_{\text{surface wall}} - W_{\text{bed}} - F_D \quad (6)$$

where P_1 and P_2 are pressures on surfaces (1) and (2) τ_w is the shear stress on the wall resulting from viscous distribution around the wall.

W_{bed} is the weight of the bed

F_D is the drag force on the multiple tubes for heat transfer immersed in the bed.

Combining equations (5) and (6):

$$P_1 A_1 - P_2 A_2 - \tau_w A_{\text{surface wall}} - W_{\text{bed}} - F_D = \frac{\dot{m}_2 V_2}{g_c} - \frac{\dot{m}_1 V_1}{g_c}$$

that is

$$P_1 A_1 - P_2 A_2 + \frac{\dot{m}_1 V_1}{g_c} - \frac{\dot{m}_2 V_2}{g_c} - \tau_w A_{\text{surface wall}} - W_{\text{bed}} - F_D = 0 \quad (7)$$

Each of the terms in equation (7) needs to be considered separately.

$A_1 = A_2$, since the geometry of the bed is rectangular.

From the continuity equation,

$\dot{m}_1 = \dot{m}_2$, for no combustion and reaction in the bed. For the low superficial velocities considered, the momentum transport terms, $\frac{\dot{m}_1 v_1}{g_c} - \frac{\dot{m}_2 v_2}{g_c}$, are approximately zero as shown in Appendix A.

For a hot bed, $\dot{m}_1 = \dot{m}_2$, due to the combustion and oxidation of coal components into NO_x , SO_2 , CO , CO_2 , and H_2O . This case is not further considered.

Therefore, equation (7) reduces to

$$P_1 A_1 - P_2 A_2 - W_{\text{Bed}} - \tau_w A_{\text{surface wall}} - F_D = 0 \quad (8)$$

where

$$\begin{aligned} A_{\text{surface wall}} &= \text{Perimeter Type Area} \\ &= (B + L + B + L) (\text{HEIGHT OF BED}) \\ &= 2 (B + L) H \end{aligned} \quad (9)$$

Under steady-state conditions Newton observed that the shear stress is directly proportional to the velocity gradient for laminar and homogenous fluid flow. Hence, the wall shear stress can be defined as follows:

$$\tau_w = \mu \frac{du}{dy} \quad (10)$$

Since the type of gas-solid flow in the bed is turbulent and the bed comprises of non-newtomian fluids; equation (10) doesn't apply. However, this equation can be modified as follows:

$$\tau_{w_{F.B}} = \mu_{F.B.} \frac{du}{dy} \quad (11)$$

Where $\tau_{w_{F.B.}}$ is the turbulent wall shear stress in the bed.

$\mu_{F.B}$ is an "effective viscosity" of the fluidized bed since the gas-solid systems possess a property, similar to that of viscosity in fluid which influence the flow characteristics. This "effective viscosity" is treated in Chapter III.

du/dy is the velocity gradient of the gas in the bed.

I. WALL SHEAR STRESS IN THE BED

To determine $\tau_{w_{F.B}}$, the following approach is taken. If gas alone is flowing through the empty grid, the wall shear stress is

$$\tau_{w_f} = \mu_f du/dy \quad (12)$$

The wall shear stress can be expressed in terms of the local skin friction as (14)

$$C_f = \frac{\tau_w}{1/2 \rho V_\infty^2} \quad (13)$$

Using equation (13).

$$\frac{C_{f_{F.B.}}}{C_{f_f}} = \frac{\left(\frac{\tau_w}{1/2 \rho V_\infty^2} \right)_{F.B.}}{\left(\frac{\tau_w}{1/2 \rho V_\infty^2} \right)_f} = \frac{(\tau_w)_{F.B.}}{(\tau_w)_f} = \frac{\mu_{F.B.} \left(\frac{du}{dy} \right)_{F.B.}}{\mu_{air} \left(\frac{du}{dy} \right)_f}$$

$(du/dy)_{F.B.} \approx (du/dy)_f$ since there is no solid transport and gas flow alone is assumed. Hence

$$\frac{\tau_{w_{F.B.}}}{\tau_{w_f}} = \frac{\mu_{F.B.}}{\mu_f}$$

Hence,

$$\tau_{w_{F.B}} = \frac{\mu_{F.B}}{\mu_f} \tau_{w_f} \quad (14)$$

The local shear stress can be expressed as (14)

$$\tau_{w(x)} = 0.332 \sqrt{\frac{V_\infty^3 \rho_f \mu_f}{x}} \quad (15)$$

where x is the distance from the leading edge of the wall. Substituting (15) in (14).

$$\begin{aligned} \tau_{w_{F.B.}}(x) &= 0.332 \sqrt{\frac{V_\infty^3 \rho_f \mu_f}{x}} \frac{\mu_{F.B.}}{\mu_f} \\ &= 0.332 \sqrt{\frac{V_\infty^3 \rho_f \mu_f}{x}} \frac{\mu_{F.B.}}{\mu_f} x^{-1/2} \end{aligned} \quad (16)$$

Applying the usual technique of averaging, $\tau_{w_{F.B.}}(x)$ can be integrated for a finite height H and divide by the integrated height to obtain;

$$\begin{aligned} \tau_{w_{F.B.}} &= 0.332 \sqrt{\frac{V_\infty^3 \rho_f \mu_f}{H}} \frac{\mu_{F.B.}}{\mu_f} \frac{\int_0^H x^{-1/2} dx}{\int_0^H dx} \\ \tau_{w_{F.B.}} &= 0.332 \sqrt{\frac{V_\infty^3 \rho_f \mu_f}{H}} \frac{\mu_{F.B.}}{\mu_f} \frac{2H^{1/2}}{H} \\ \tau_{w_{F.B.}} &= 0.664 V_\infty \frac{\mu_{F.B.}}{\mu_f} \sqrt{\frac{V_\infty \rho_f \mu_f}{H}} \end{aligned} \quad (17)$$

Since the flow is across the tube banks, the velocity must be maximum at minimum flow area. Therefore V_∞ in equation (17) should be V_{max} . Hence

$$\tau_{w_{F.B}} = 0.664 V_{\max} \frac{\mu_{F.B}}{\mu_f} \sqrt{\frac{V_{\max} \rho_f \mu_f}{H}} \quad (18)$$

II. EFFECTIVE WEIGHT OF THE BED

The effective weight of the bed, W_{Bed} is

$$W_{\text{Bed}} = \frac{m_s g}{g_c} \frac{(\rho_s - \rho_f)}{\rho_s} \quad (19)$$

III. DRAG FORCE ON THE HEAT TRANSFER TUBES

The drag force is due to the pressure distribution around the tubes, viscous effect and collision of particles. This force is represented for multiple tubes as;

$$F_D = A_c C_{D_{F.B}} \rho_f \frac{V^2}{2} N \quad (20)$$

where

A_c = projected area of the cylinder normal to the flow

$= D_c \times L_t$

ρ_f = Density of the fluid

N = Number of tubes

$V = V_{\max}$ = maximum velocity through the tubes

C_D = drag coefficient and is a function of Reynolds number.

that is,

$$C_D = f \left\{ \frac{V_{\max} D_c \rho_f}{\mu_{F.B}} \right\}$$

Therefore, the drag force F_D is,

$$F_D = \{ D_C L_t C_{D_{F.B}} \frac{V_{\max}^2}{2} \rho_f N \} \quad (21)$$

In order to evaluate the total pressure drop ΔP_{total} , the drag coefficient of the fluidized bed, $C_{D_{F.B}}$, needs to be known. Mehta et al. (9), Hariu and Molstad (8) presented data on the drag coefficients of transport particles in a fluidized bed. Because in this thesis, the net transport of particles are not considered, the drag coefficients on the tubes due to the fluidization are approached differently. Since the drag coefficients are influenced by the Reynolds number, it can be represented as follows;

$$C_D = f \left(\frac{V_{\max} D \rho}{\mu} \right) \quad (22)$$

where D is some characteristic dimension of the body.

In a fluidized bed of large and heavy particles with no net transport, the velocity of the particle may be considered small compare to that of the fluid. Hence, by continuity, the mass of the fluid flowing into the tube banks must be equal to that going out of it. That is,

$$\dot{m}_{\text{Flow}} = \rho_f V_{\max} A \quad (23)$$

$$\frac{\dot{m}_{\text{Flow}}}{A} = \rho_f V_{\max} \quad (24)$$

The Reynolds number can be represented as

$$R_e = \frac{\dot{m}_{\text{Flow}}}{A} \frac{D}{\mu} \quad (25)$$

Figure 7 shows the drag coefficients, C_D as a function of the Reynold numbers for cylinders. This figure can be used to estimate the drag coefficients of the tubes during fluidization if the effective viscosity of the fluidized bed is used instead of that of the viscosity of the fluid. The effective viscosity data are given on page 32. Equation (25) can be modified with effective viscosity as,

$$C_{D_{F.B}} = f \left(\rho \frac{V_{\max} D}{\mu_{F.B}} \right) \quad (26)$$

Because a fluidized bed resembles a fluid of high viscosity (3), it has a lower Reynold's number and a high drag coefficient compare to a single phase fluid (see Appendix B).

Maximum Velocity in the Tube Bank .

In order to evaluate ΔP_{total} , the maximum velocity, $V_{\max}$ occurring in the tube bank needs to be known. This is the velocity through the minimum flow area. In order to know this velocity, the minimum flow area needs to be determined. From Figure 8, the minimum flow area is $(S_n - D)$. The ratio of minimum flow area to the total frontal area is,

$$(S_n - D) / S_n$$

Thus, the maximum velocity is

$$V_{\max} = V_{\infty} \frac{S_n}{S_n - D} \quad (27)$$

where V_{∞} is the superficial velocity of the gas. Putting equations 9, 18, 19 and 21 in 8,

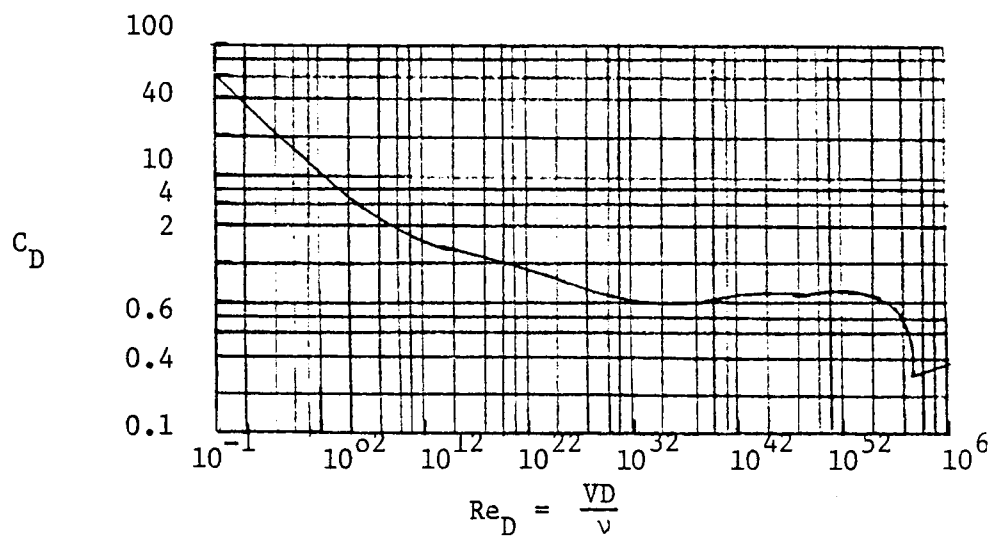


Figure 7. Drag coefficients for cylinders (14).

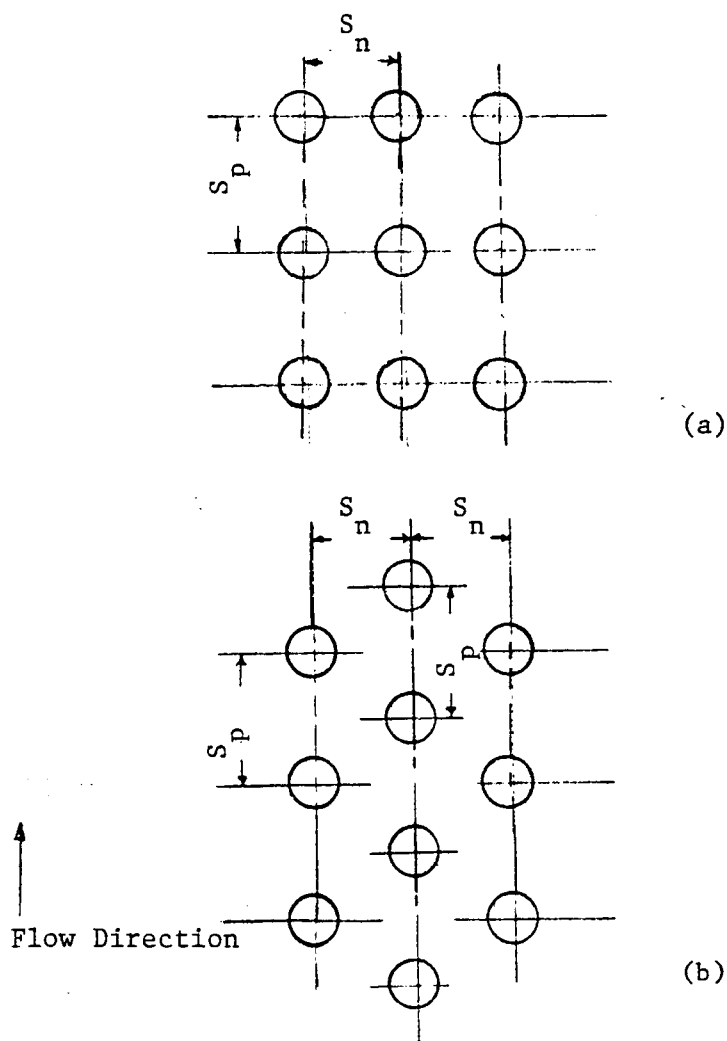


Figure 8. (a) In-line tube rows; (b) Staggered tube rows.

$$P_1 A_1 - P_2 A_2 - \frac{m_s g}{g_c} \frac{(\rho_f - \rho_s)}{\rho_s} - 0.664 V_{\max} \frac{\mu_{F.B}}{\mu_f} \sqrt{\frac{V_{\max} \rho_f \mu_f}{H}} \\ (2(B+L)H) - D_{CTD} \frac{V_{\max}^2}{2} \rho_f N = 0 \quad (28)$$

But $A_1 = A_2 = A_B$, therefore

$$\Delta P = \frac{1}{A_B} \left(\frac{m_s g}{g_c} \frac{(\rho_s - \rho_f)}{\rho_s} + 0.664 V_{\max} \frac{\mu_{F.B}}{\mu_f} \sqrt{\frac{V_{\max} \rho_f \mu_f}{H}} \right. \\ \left. + D_{CTD} \frac{V_{\max}^2}{2} \rho_f N \right) \quad (29)$$

where $\Delta P = (p_1 - p_2)$

But from equation (27), $V_{\max} = V_{\infty} \frac{S_n}{S_n - D}$, substitute in equation (29),

$$\Delta P = \frac{1}{A_B} \left(\frac{M_s g}{g_c} \frac{(\rho_s - \rho_f)}{\rho_s} + 0.664 \frac{V_{\infty} S_n}{S_n - D} \frac{\mu_{F.B}}{\mu_f} \sqrt{\frac{V_{\infty} S_n}{(S_n - D)} \frac{\rho_f \mu_f}{H}} (2(B+L)H) \right. \\ \left. + D_{CTD} \frac{(V_{\infty} S_n)^2}{(S_n - D)^2} \rho_f N \right) \quad (30)$$

It can be seen from equation (30), that the total pressure drop may be considered to consist of three parts, namely,

$$\Delta P_{\text{total}} = \Delta P_{w/a} + \Delta P_{\text{wall shear}} + \Delta P_{\text{tube drag}} \quad (30a)$$

where,

$\Delta P_{w/a}$ - the pressure drop due to the weight of the bed per unit area

$\Delta P_{\text{wall shear}}$ - the pressure drop due to the wall shear

ΔP_{tube} - the pressure drop due to the heat transfer tubes

Alternate Method of Determining Drag Force on the Heat Transfer Tube

The method described in this section can also be used to determine the drag force on the heat transfer tubes (or the pressure drop due to the heat transfer tubes). The procedure is as shown in Figure 9.

Figure 9 shows a control volume with the gas alone flowing across the tubes. Writing the momentum equation on the control volume,

$$\Sigma F = \frac{d}{dt} \int_{cv} \rho V dV + \int_{cs} V(\rho V \cdot dA) \quad (31)$$

Since there is no flow of gas in x and z direction, the momentum equation reduces to one-dimensional. The y-direction momentum equation is

$$\Sigma F_y = \frac{d}{dt} \int_{cv} \rho V_y dV + \int_{cs} V_y(\rho V \cdot dA) \quad (32)$$

Since the density has a uniform value over area A_1 and A_2 , equation (32) can be rewritten as,

$$\Sigma F_y = \rho_1 \int_{cs_1} V_y(V \cdot dA) + \rho_2 \int_{cs_2} V_y(V \cdot dA) = \dot{m}_2 V_2 - \dot{m}_1 V_1 \quad (33)$$

$$\text{But } \Sigma F_y = P_1 A_1 - P_2 A_2 - F_D \quad (34)$$

Substituting (33) in (34), the following equation is obtained,

$$P_1 A_1 - P_2 A_2 - F_D + \dot{m}_1 V_1 - \dot{m}_2 V_2 = 0 \quad (35)$$

$A_1 = A_2$, geometry of the figure is rectangular.

$$\frac{\dot{m}_1 V_1}{g_c} - \frac{\dot{m}_2 V_2}{g_c} \approx 0 \quad (\text{See Appendix A})$$

$$m_1 = m_2 \quad (\text{continuity})$$

$$V_1 = V_2$$

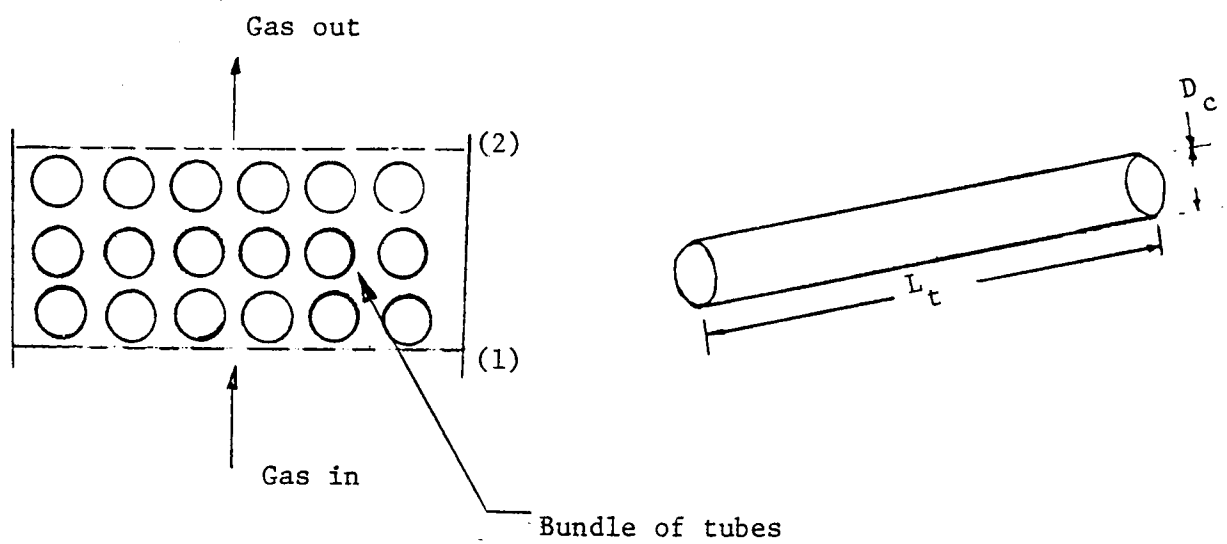


Figure 9. Control volume with gas alone flowing across the tubes.

Hence, equation (35) reduce to

26

$$(P_1 - P_2) A - F_D = 0 \quad (36)$$

that is,

$$(P_1 - P_2) A = F_D \quad (37)$$

where

$$F_D = \frac{F_D}{N} \times N = F_D^1 \times N$$

Putting the above in (37).

$$F_D^1 = \frac{(P_1 - P_2) A}{N} = C_D \rho_f \frac{V_{\max}^2}{2} A$$

where $A = D_C L_T$

$$\text{Hence } F_D^1 = \left(\frac{(P_1 - P_2)}{N} \right)_{\text{gas alone}} A = C_D \rho_f \frac{V_{\max}^2}{2} D_C L_T \quad (38)$$

$$\text{that is, } ((P_1 - P_2) A)_{\text{gas alone}} = C_D \rho_f \frac{V_{\max}^2}{2} D_C L_T N \quad (39)$$

Assuming that the pressure drop in the fluidized bed due to heat transfer tubes with no solid transport may be represented by:

$$((P_1 - P_2) A)_{F.B(tubes)} = C_{D_{F.B}} \rho_f \frac{V_{\max}^2}{2} D_C L_T \quad (40)$$

Divide (39) into (40);

$$\frac{((P_1 - P_2) A)_{F.B.(tube)}}{((P_1 - P_2) A)_{\text{gas alone}}} = \frac{C_{D_{F.B}}}{C_{D_{\text{gas}}}}$$

that is

$$((P_1 - P_2))_{F.B(tubes)} = \frac{C_{D_{F.B}}}{C_{D_{\text{gas}}}} ((P_1 - P_2))_{\text{gas}} \quad (41)$$

Equation (41) gives the pressure drop due to the heat transfer tubes in the fluidized bed. This is also the drag per unit area on the tubes. $(P_1 - P_2)_{\text{gas}}$ can be evaluated from equation given by Holman (15). This equation is used for sample calculation in the Appendix B and C. $C_{D_{F.B}}$ can be estimated from Figure 7 by using the effective viscosity to calculate the Reynolds number.

In equation (29), the pressure drop due to the heat transfer tubes is expressed as:

$$\Delta P_{\text{tubes}} = (D_{C,T} L_{C,T} C_{D_{F.B}} \frac{v_{\text{max}}^2}{2} \rho_f N) \frac{1}{A_B} \quad (42)$$

Also equation (41) can be rewritten as

$$\Delta P_{F.B} (\text{tubes}) = \frac{C_{D_{F.B}}}{C_{D_{\text{gas}}}} \Delta P_{\text{gas}} \quad (43)$$

CHAPTER III

EFFECTIVE VISCOSITY OF THE FLUIDIZED BED

It has been found possible to relate some flow properties of fluidized solids to the effective viscosity of the fluidized bed. Matheson et al. (13) showed that a well fluidized bed was analogous to the upward flow of gas through a low viscosity fluid, and that aerated viscous fluids show the same type of slugging observed with coarse particles.

The "effective viscosity" of the fluidized bed had been investigated by number of investigators (11), (12), (13), (23) and (25). Some of these investigators (11), (12), (13), (25) used a modified Stormer-type viscometer provided with a paddle that turns in the bed. This method of measurement does not produce a true viscosity measurement because of the acceleration of masses and added interparticle frictional effects which occur coupled with normal shear properties; however, it has been shown to be a good measurement of flow characteristics of a bed (12) (13).

A hypothetical variation of viscosity with fluid velocity (11) is shown in Figure 10. The figure is divided into three regions. In region I, the viscosity decreases with small increase in gas velocity. The rapid decrease was caused by the exposure of the bed to the gas whereby the particles experience a little motion which resulted

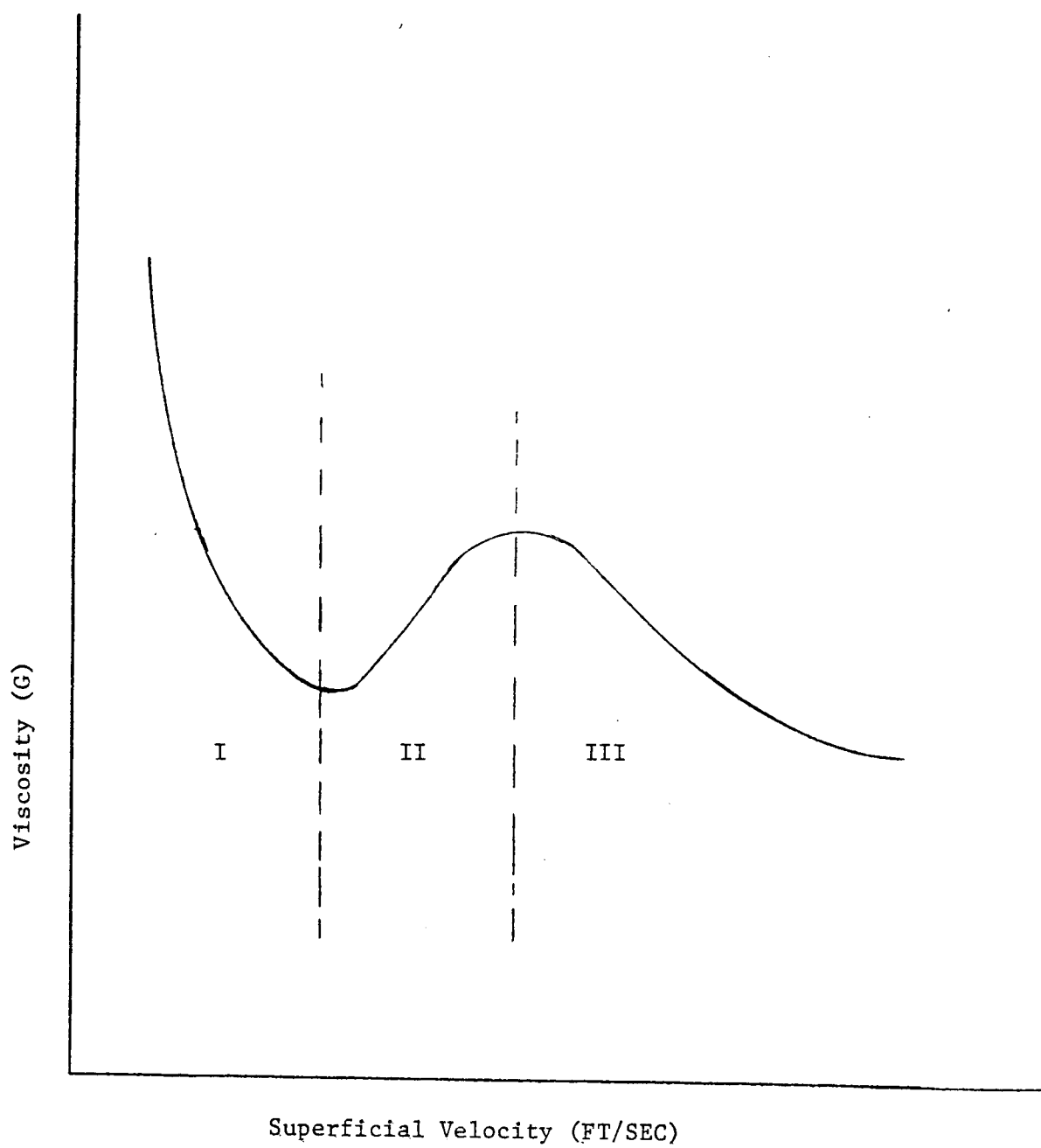


Figure 10. Theoretical variation of viscosity with flow rate (11).

in decrease in bed density. In region II, there is an increase in superficial velocity and a certain degree of freedom is imparted to the particles. At this stage, the particles try to redistribute themselves and in doing so, they caused an increased number of collisions with the rotating paddle which resulted in a net increase in viscosity. In region III, with increasing gas velocity, the particle motions increased and the bed density decreased. This decrease in bed density affects viscosity and the latter decreases again. At very high velocity, increase in velocity has little effect on viscosity and the latter approaches a constant value.

The above discussions showed that the effective viscosity of fluidized bed is a function of bed density (number of particles per unit volume) and the fluid velocity.

Haas (12) measured the viscosity of fluidized bed of glass beads using modified Stormer viscometer shown on Figure 11. The viscosity is measured by determining the weight required to rotate the paddle at 200 revolutions per minute in the bed. The weight required to rotate the paddle at the same speed with the same air velocity in an empty column was subtracted from this value to give a net value of the effective viscosity. Some of the data obtained for beads of sizes 280, 484, 1020 microns are shown on Table I. Furnkawa and Ohmae (20) performed similar experiment with spherical poly (vinyl acetate) beads but used 113 revolutions per minute. Schügerl et al. (23) also present data on viscosity of bed of glass beads and quartz particles using couette viscometer. The plot of Haas data is shown on Figures 12-14. It

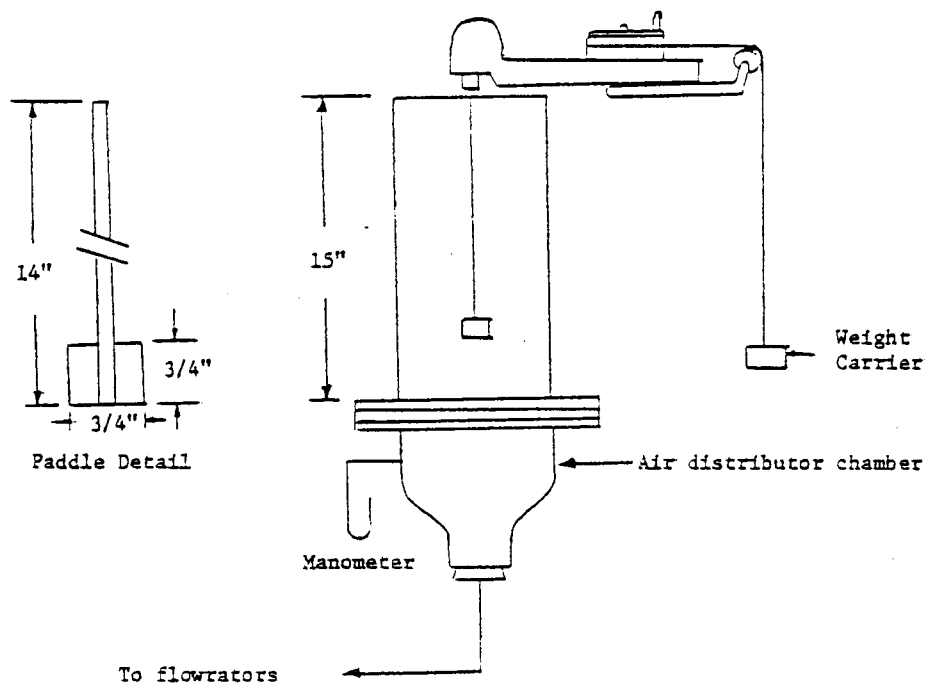


Figura 11. Stormer viscometer (12).

Table 1

Effective Viscosity of the Fluidized Bed of Glass Beads (12)

$D_p = 1020$ Microns			
Velocity (ft/sec)	Weight to turn the paddle in the bed at 200 rpm (gm)	$\mu_{F.B}$ (gm)	μ_{air} (gm)
1.46	43.5	35.5	8.0
1.62	39.5	31.5	8.0
1.37	54.5	46.5	8.0
1.73	48.5	40.5	8.0
1.85	52.0	44.5	8.0
1.96	51.5	43.5	8.0
2.02	47.0	39.5	8.0
2.08	45.5	37.5	8.0
2.18	44.5	36.5	8.0

$$(\mu_{air}/\mu_{F.B}) = \frac{8.0}{36.5} \approx 0.22$$

fully developed
fluidization
 $D_p = 484$ Microns

0.74	22.2	18.9	3.3
0.78	24.2	20.9	3.3
0.85	25.1	21.8	3.3
0.96	26.5	23.2	3.3
1.07	27.7	24.4	3.3
1.16	27.2	23.9	3.3
1.24	28.2	24.9	3.3
1.38	27.0	23.7	3.3

Table 1 (cont'd)

Velocity (ft/sec)	$D_p = 484$ Microns		
	Weight to turn the paddle in the bed at 200 rpm (gm)	$\mu_{F.B}$ (gm)	μ_{air} (gm)
1.50	26.0	22.7	3.3
0.68	23.7	20.4	3.3
0.63	25.2	21.9	3.3

$$\left(\frac{\mu_{air}}{\mu_{F.B}} \right)_{\text{fully developed}} \approx 0.127$$

F.B fluidization

$D_p = 280$ Microns			
1.268	14.7	13.6	1.1
0.237	19.2	18.1	1.1
0.216	31.7	30.6	1.1
0.319	17.4	16.3	1.1
0.350	18.2	17.1	1.1
0.383	19.7	18.6	1.1
0.41	20.2	19.1	1.1
0.49	21.2	20.1	1.1
0.70	23.7	22.6	1.1
0.91	21.7	20.6	1.1
0.98	21.0	20.0	1.1
1.05	20.2	19.1	1.1

$$\left(\frac{\mu_{air}}{\mu_{F.B}} \right)_{\text{fully developed}} \approx 0.058$$

F.B fluidization

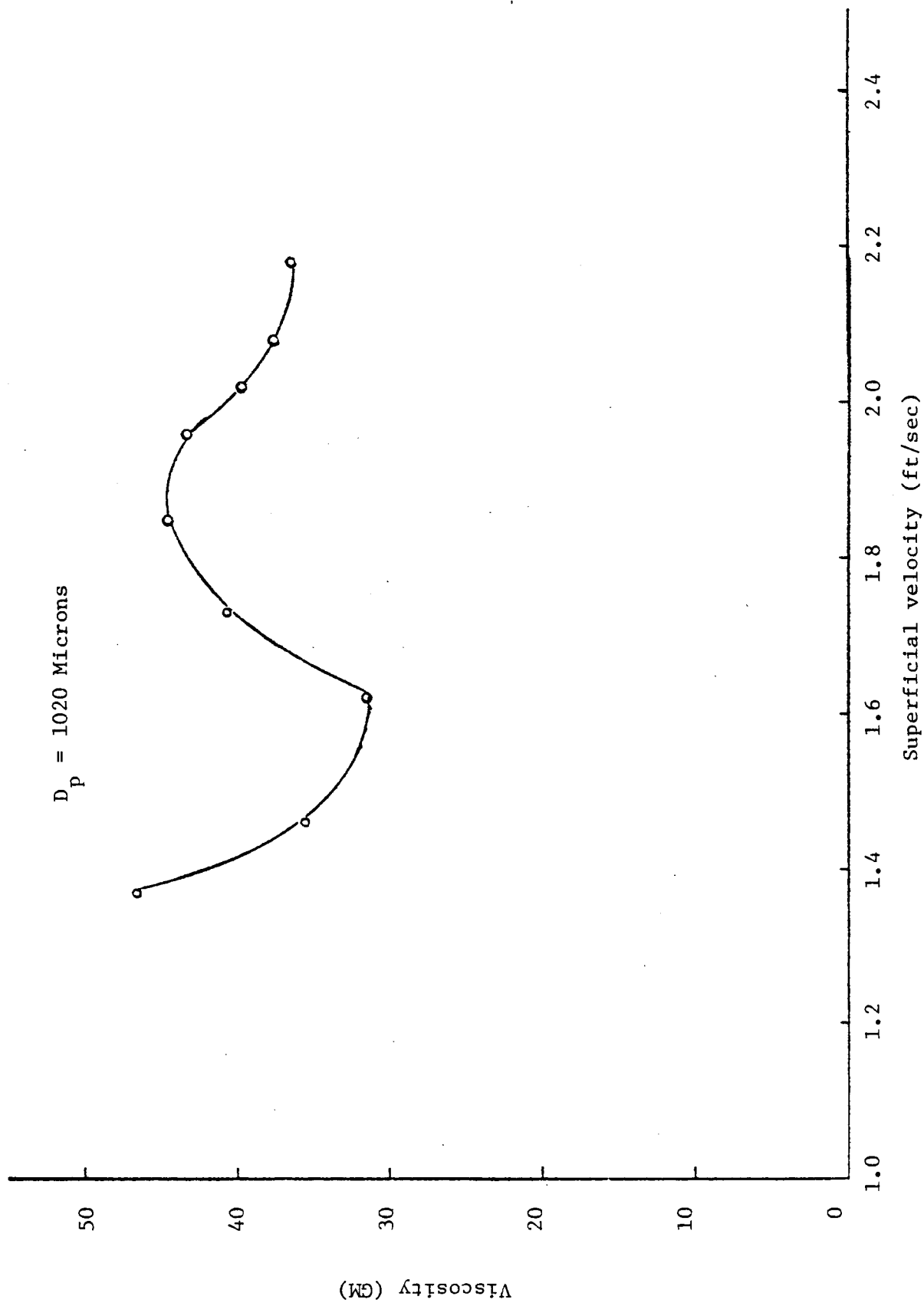


Figure 12. Effective viscosity of fluidized bed of glass beads particle (12).

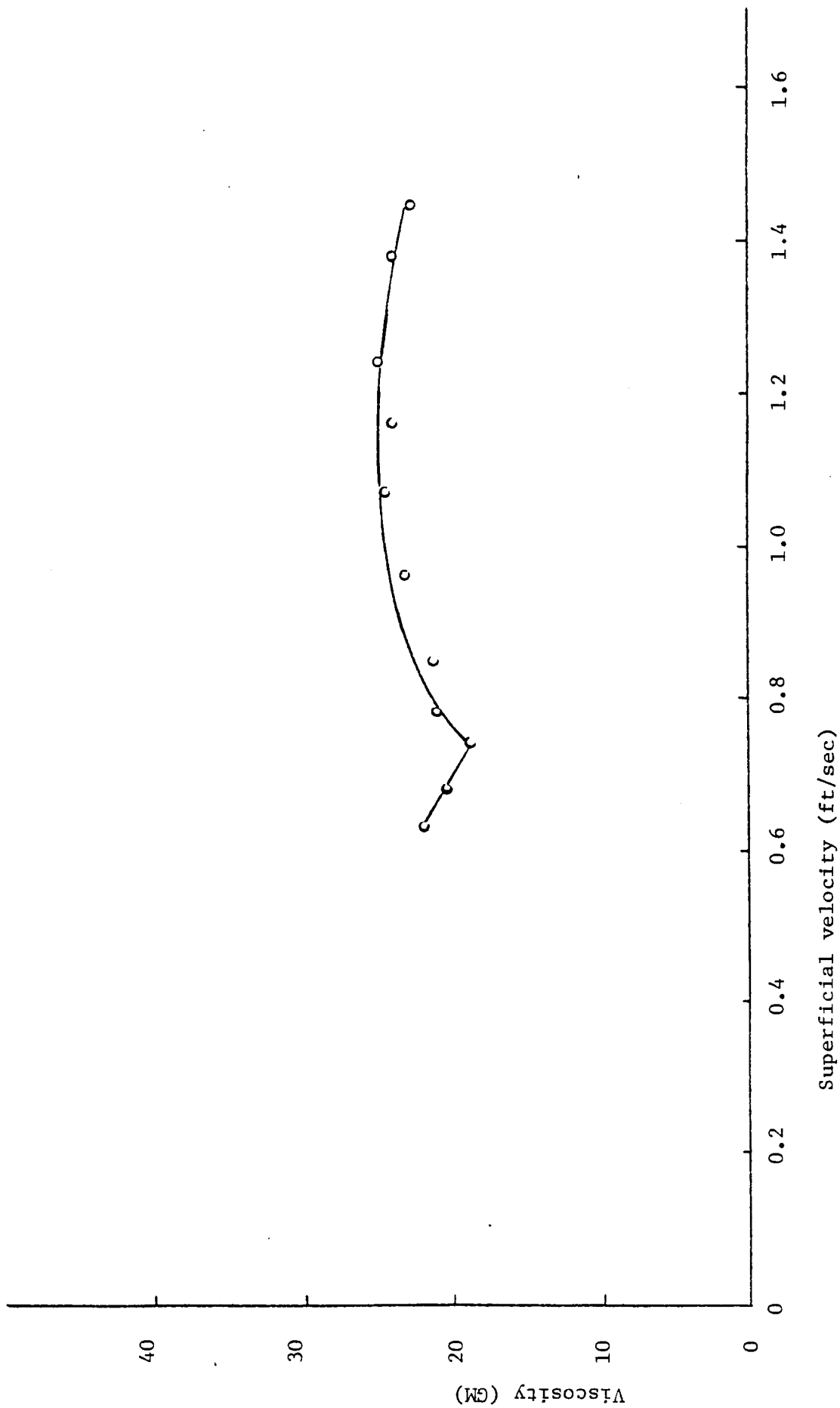


Figure 13. Effective viscosity of fluidized bed of glass beads particle ($D_p = 484$ Microns) (12).

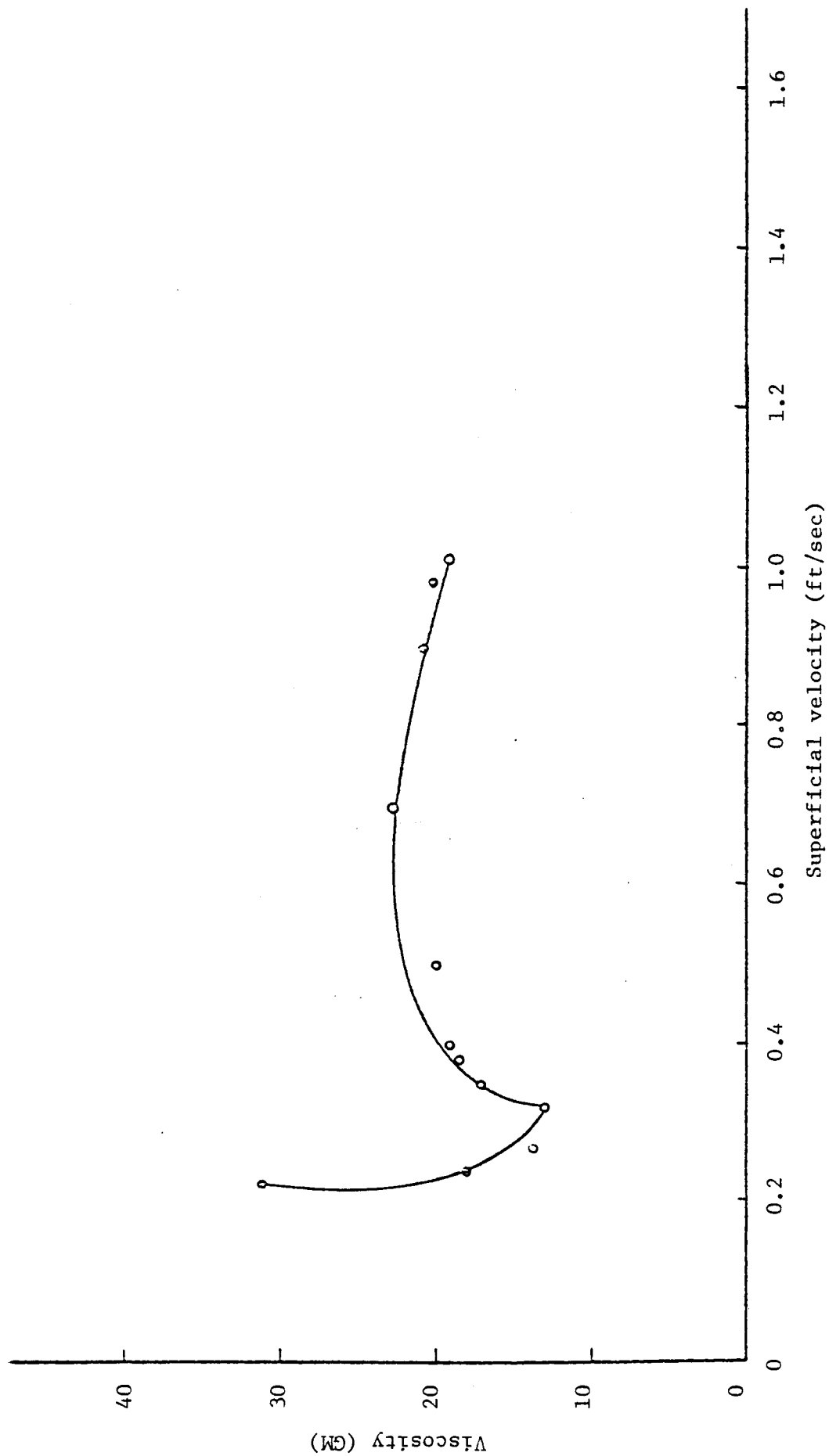


Figure 14. Effective viscosity of fluidized bed of glass beads particle ($D_p = 280$ Microns) (12).

can be recognized from these figures that at low velocity, the viscosity is very high and decreases very rapidly with increases gas velocity to a point where it approaches a constant value. It is assumed in this thesis that at the superficial velocity of interest, the effective viscosity approaches a constant value because at this stage, the whole column of the bed will be saturated with the particles and the bed density will be independent of degree of aeration.

Because Haas quantified his measurement with that of air, his data are used to express the effective viscosity of the fluidized bed (in the constant value region as a function of that of air (see Table 1).

CHAPTER IV

EXPERIMENTAL APPARATUS FOR PRESSURE DROP MEASUREMENT IN FLUIDIZED BED

I. GENERAL

This chapter presents the conceptual design of an experimental apparatus for the pressure drop measurement in a fluidized bed grid. This apparatus consists of plenum, air distribution plate, bed particle, inner and out plexiglass, manometer, and heat transfer surfaces (see page 42). Each part of the apparatus is described under the "Experimental Apparatus Discussion."

II. INTRODUCTION

The design of a fluidized bed grid requires the knowledge of particle size, bed density, and gas fluidizing velocity. In addition to these parameters, pressure drop knowledge is also significant. The earlier investigations on pressure drop were done on pneumatic transport of solid particles in a fluidized medium. Experimental and theoretical studies of pressure drop in vertical and horizontal tubes were carried out by Mehta et al. (9), Jones et al. (4), Hariu and Molstad (8) and Belden and Kasel (5).

Max Leva (3), Zenz and Othmer (2), and Davidson and Harrison (19) showed that pressure drop is a direct function of the gas fluidizing velocity for fixed bed and become constant after incipient

fluidization. The hypothetical variation of pressure drop with gas flowing in a fluidized bed without heat transfer surfaces is shown in Figure 15. In the fixed bed at low velocity, the passage of gas up through the bed causes a relatively low pressure drop. The gas percolates through the bed without agitating the individual particles and the pressure drop is less than that equivalent to the weight of the bed. As the velocity is increased, expansion of the bed takes place and a certain degree of freedom to move is imparted to the individual particle. This is in the region A-B in the figure where pressure drop is a direct function of gas velocity. Very small particles of diameter less than 10 microns do not follow this type of generalization because of their cohesive flow property.

At point B, the particles are fully suspended on the gas stream and gain some freedom to move. In the region B-C, a small rearrangement of particles occurs at a slightly higher gas velocity as the particles are in random motion in all directions. In the region D-E, the bed is completely supported by the gas stream and from this point on the pressure drop over the bed is equivalent to the particles weight in the bed per unit area with no heat transfer surfaces in the bed.

The design reported here gives the experimental apparatus for the measurement of the pressure drop in a bed of glass beads with heat transfer surfaces running across the bed horizontally. Air is the fluidizing gas.

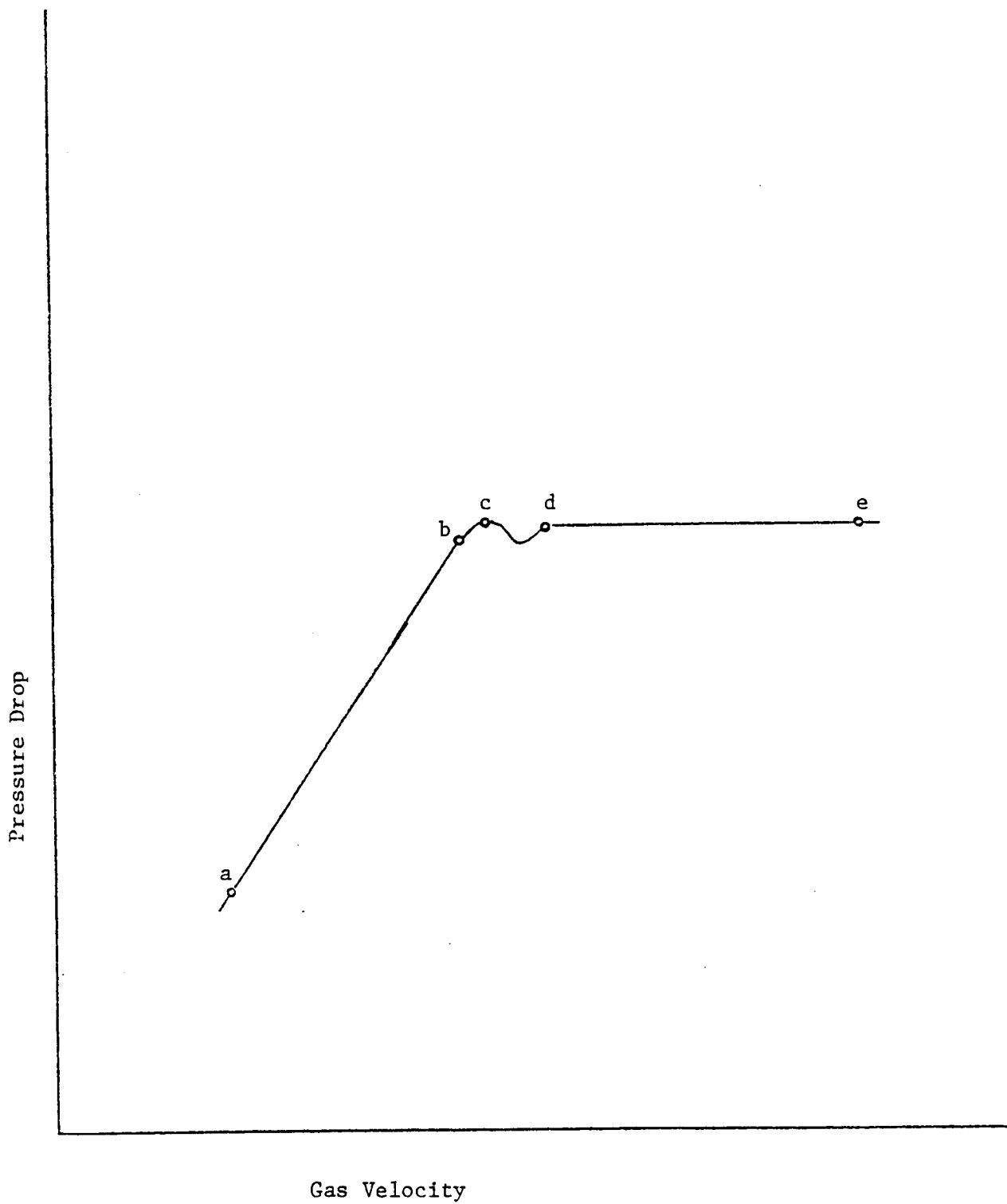


Figure 15. Hypothetical variation of pressure drop with gas velocity.

III. DISCUSSIONS

The basic experimental apparatus for measuring the pressure drop in a fluidized bed grid is shown in Figures 16 and 17. It is designed in such a way that the particles behavior as the gas velocity is increased can be visually observed. The grid is rectangular in shape, 2 ft by 1 ft. This will ease the horizontal heat transfer surface construction and also give a uniform flow of particles over the tubes. Sides A and B of the grid are covered with double wall plexiglass each. The inner plexiglass are tube holders and they are screwed to the 1" angle iron or aluminum frame. Each end of the tube is sealed to avoid gas leakage. The outer plexiglass is connected to the frame as shown with full rod. These sides were designed like this for easy removal whenever the tubes configuration needs to be changed. In the sample calculations made (see Appendix B) both in-line and staggered tubes configuration are considered. Manometer are attached to the inner glass of side B as shown to measure the pressure drop across the grid and across the heat transfer tubes.

Sides C and D are covered with plywood. Other parts of the grid are described below.

IV. PLENUM

This is a space provided below the air distributor grid. The air supply from the blower or fan passes through this space and up to the air distributor. A fan size of 480 cfm will be required to supply the air at a superficial velocity of 4ft/sec.

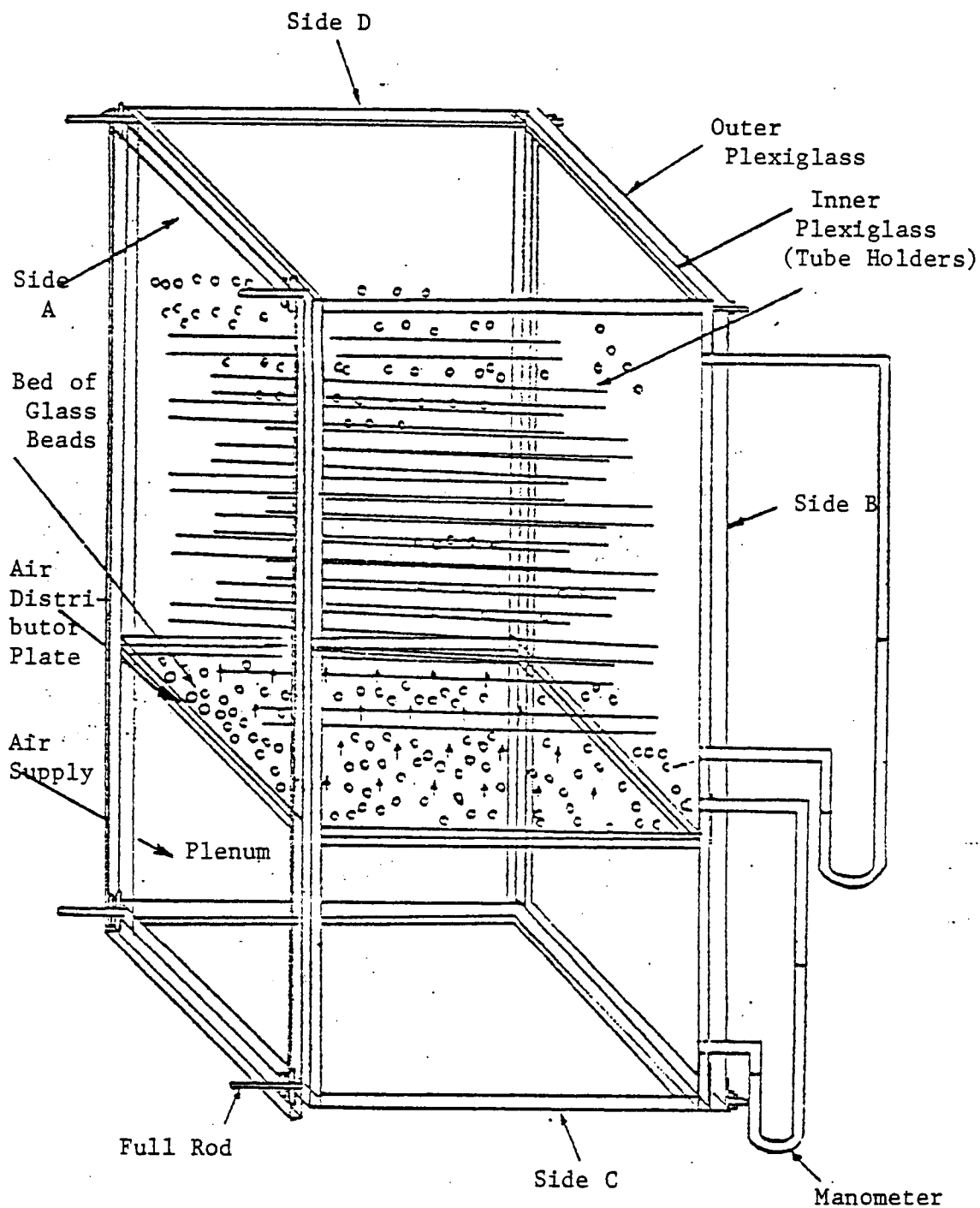


Figure 16. Fluidized bed grid.

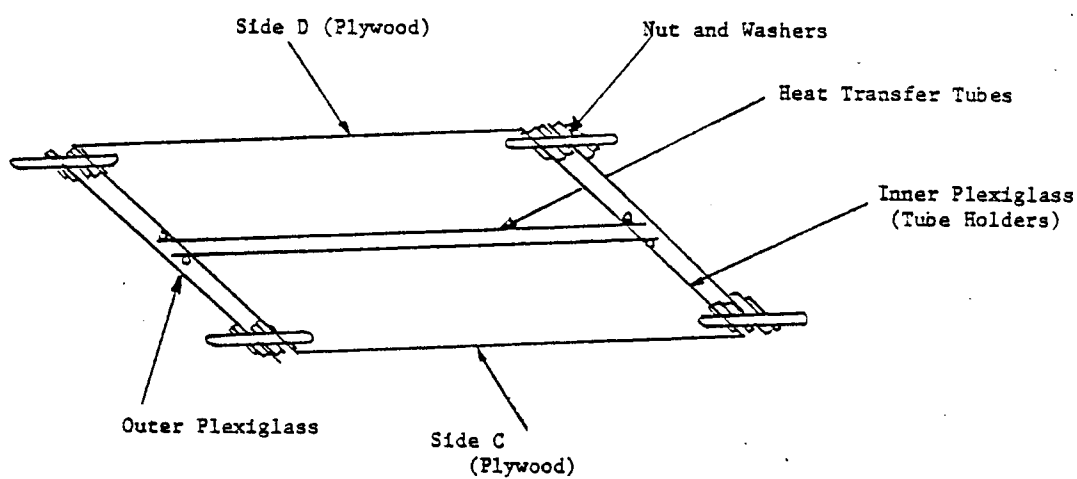


Figure 17. Top view of fluidized bed grid.

V. HEAT TRANSFER TUBE DIAMETER

A heat transfer tube diameter of 0.5 in. was chosen for this design. Sample calculation of pressure drop due to this size of tubes are made in the Appendix B. In order to examine the effect of pressure drop on larger size of tubes, additional calculations were made with 1.5 in. diameter tube in the Appendix C.

VI. BED SHAPE AND HEIGHT

For ease of tube construction, a rectangular bed is chosen. Also, rectangular bed with heat transfer surfaces extending across tends to decrease wall effects.

In determining the height of the bed, many factors are considered. These factors include calculating the slug height from the equation given by Yaqi and Muchi (21) and the expansion factors from the equation given by Nguyen and Bergougnou (22). (see Appendix A1). From these calculations, a slugging height of 19.0 in. was obtained, therefore an expansion height of 20 in. was chosen. An expansion factor of 2.02 was calculated which results in a minimum height of 9.9 in.

VII. HEAT TRANSFER TUBE ARRANGEMENT

In this design, a tube diameter of 0.5 in. was chosen. The distribution of the heat transfer tubes will be 3 in. from the air distributor's grid up to a height of 20 in.

In order to know the effect of the pressure drop on different heat transfer surface configurations, both in-line and staggered tube arrangements need to be considered. Sample calculations for the pressure drop for different configurations for both in-line and staggered tube arrangements are given in the Appendices B and C.

VIII. AIR DISTRIBUTOR GRID

Bangulo et al. (26) tested different types of air distributor plates for good fluidization. Among these distribution plates are perforated plate, plates with drilled vertical orifices and nozzle buttons. Perforated plate air distributor grid gave non acceptable fluidization because part of the bed remained static and part formed intermediate channelling. Plates with drilled vertical orifices resulted in small static area between holes, overheating and clinkering (in case of hot bed), and plugging of the orifice hole. Nozzle buttons produced good fluidization and resisted plugging. The design of the nozzle made from automatic screws is shown in Figure 18 (26). The plane into which the orifices were drilled was sloped at 15° and sufficient thickness allowed for the hole to impede the infiltration of the bed material into the plenum.

Since the design shown in Figure 18 may be difficult to produce and can not be bought ready made, it was suggested by Belk et al. (27) that a standard size bolt be adopted for the design (see Figure 19). This design permits cheap production and easy removal. The nozzle button is designed to the dimension shown in Figure 19, and produced on standard size bolt. Figure 20 shows the arrangement of

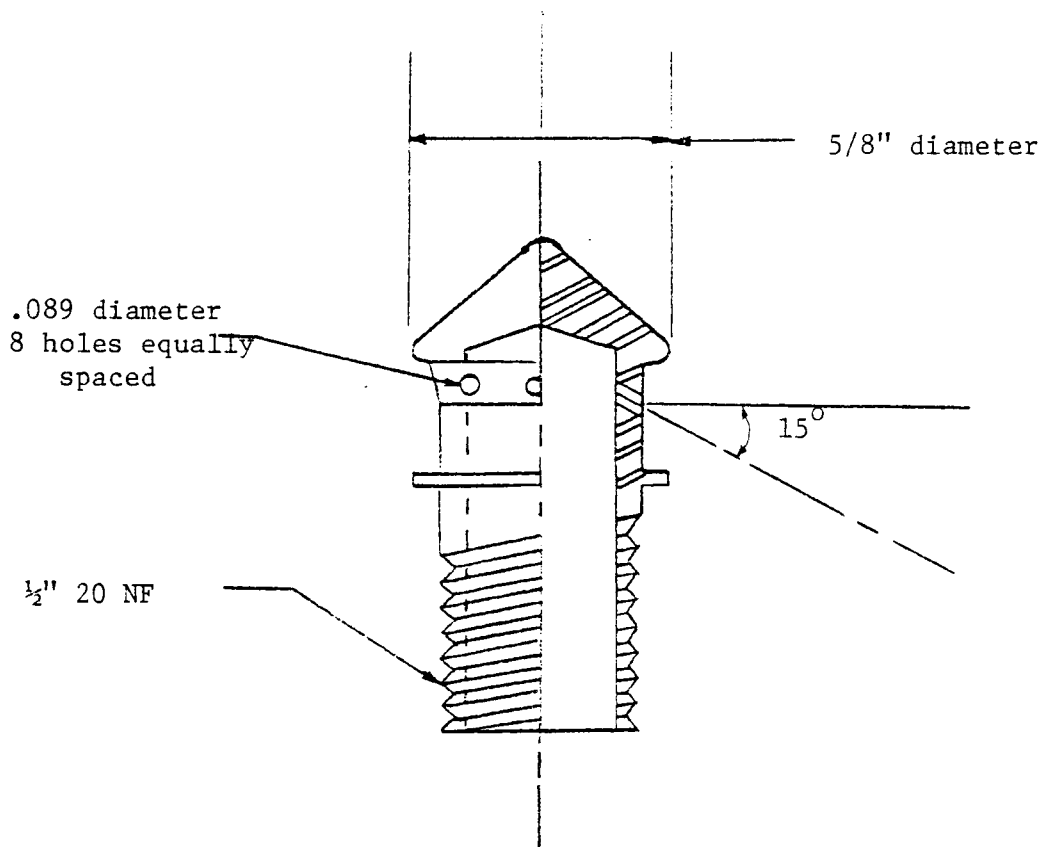


Figure 18. Nozzle button.

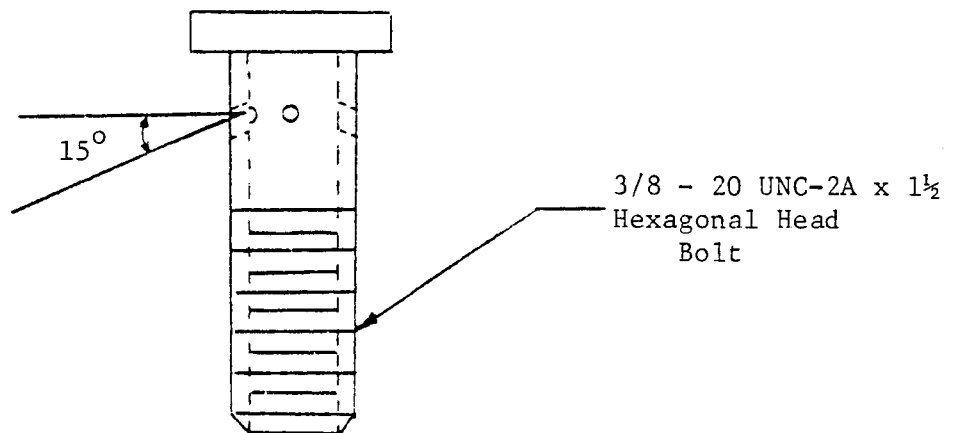


Figure 19. Standard size bolt for nozzle button.

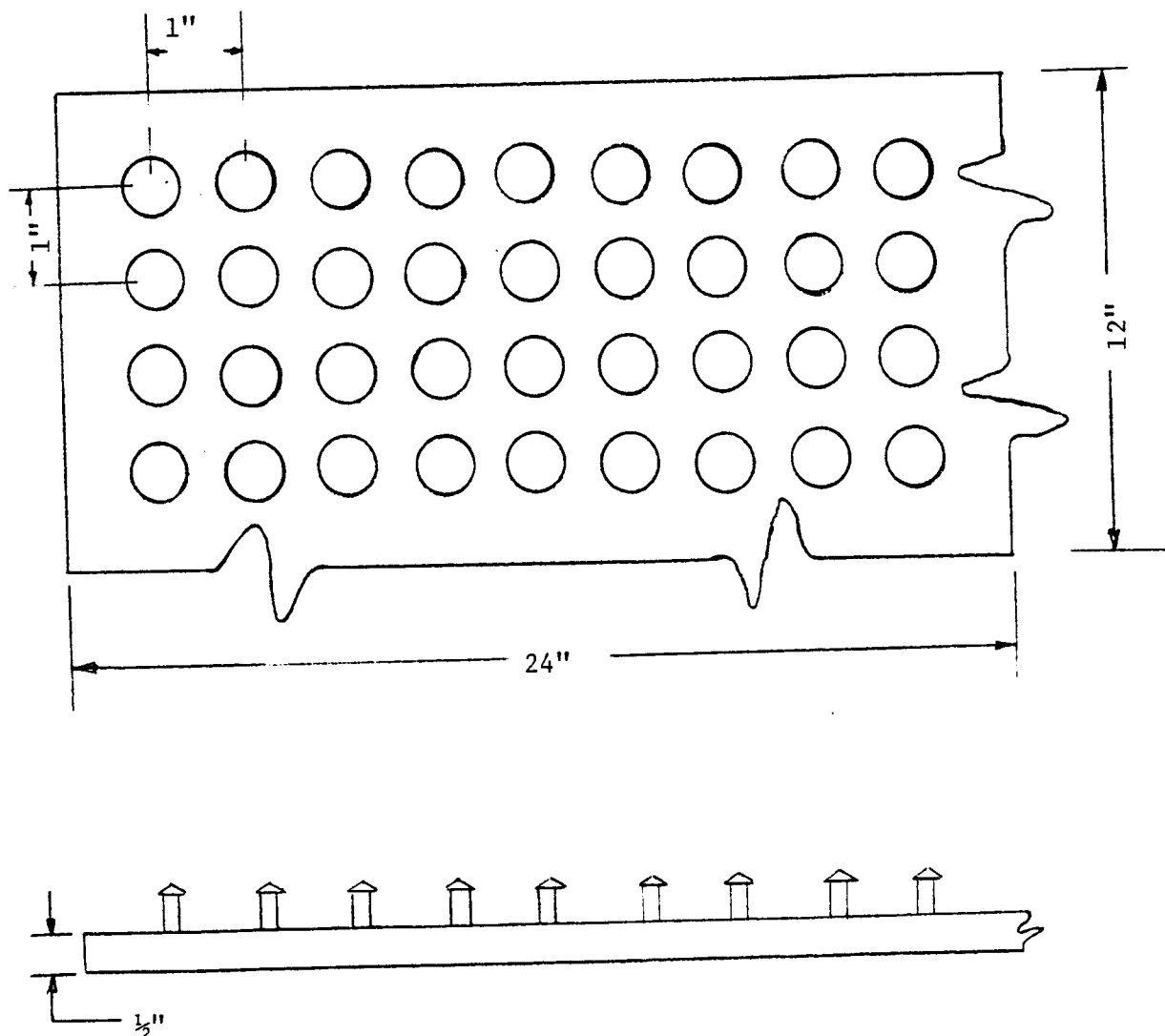


Figure 20. Air distribution grid arrangement.

the nozzles on air distributor grid. The nozzle spacing is 1.0 in. center on rows and columns. This arrangement gives twelve rows and twenty four columns of buttons with a total of 288. The pressure drop obtained due to the air distributor is 5.6 in. of water (see Appendix A-2) and it is approximately one-third of the total pressure drop in the fluidized bed.

Because of the nozzle distribution over the plate, a detailed analysis of the stresses presents a difficult problem. The bed load is only 191 lbm and as the thickness of the plate is not critical to the overall design, the chosen thickness of 1/2 in. would be quite adequate.

The detail of nozzle bolt and distribution grid is shown on Figure 21. The nozzle bolt is 3/8 in. 304 stainless steel type with 288 nozzles on the plate. Each bolt contain eight holes (orifices) and equally spaced. The holes are to be drilled at a 15° slope with sufficient thickness allowed so that the bed material may not fall through into the plenum chamber during operation.

IX. FLUIDIZATION

Particle Size

This is one of the variables that require a careful examination in the design of a fluidized bed. Particles of diameter less than 10 microns usually yield cohesive flow which results in balling of the solids into spheres of several millimeters in diameter at gas velocity of up to 1 foot per second. The individual particles appear to be largely locked together by the attractive forces which all small

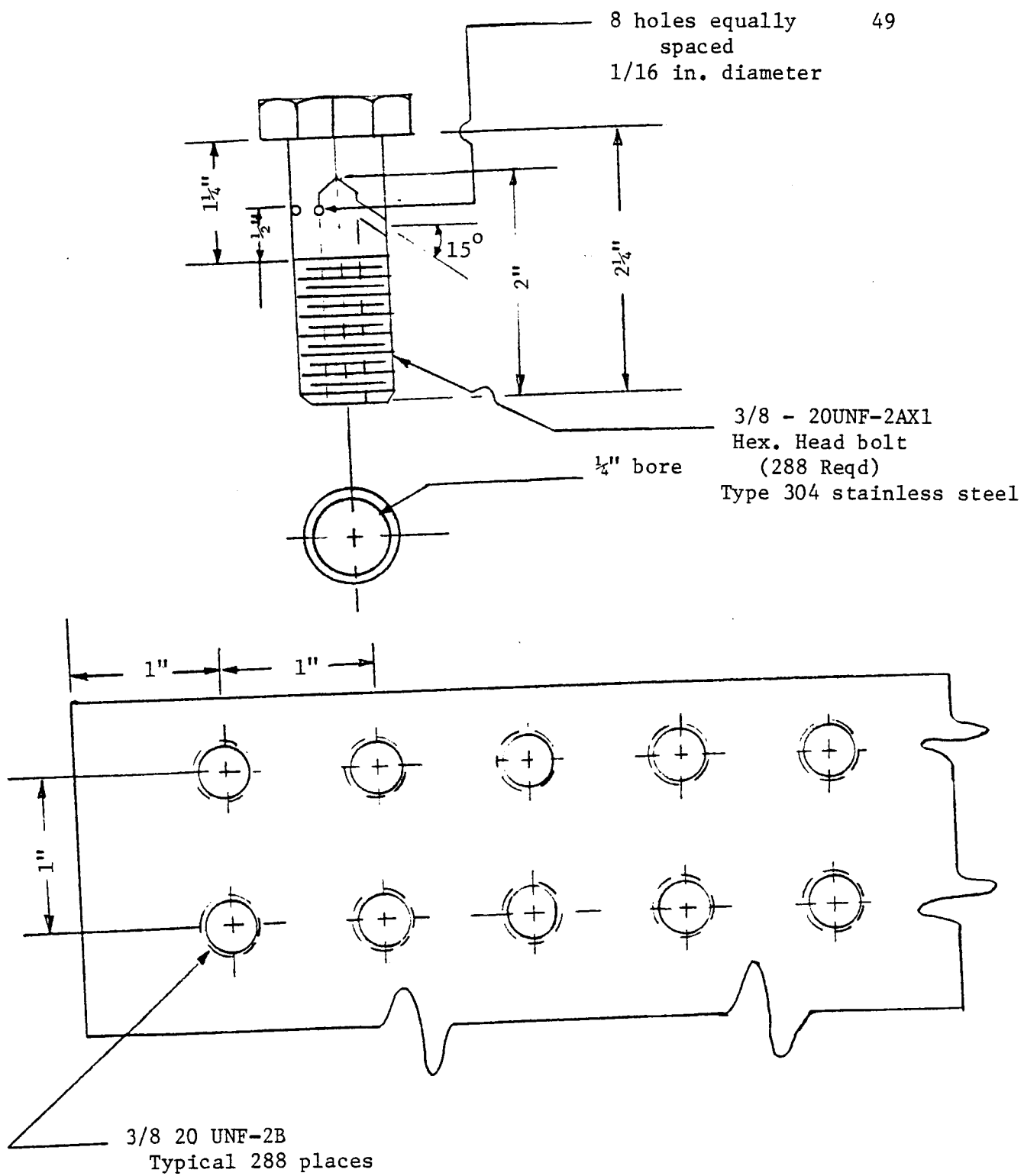


Figure 21. Detail of nozzle bolt and distribution grid.

particles seem to have. This agglomeration of particle distrupts the flow and may cause formation of large lumps of particles. On the other hand, large particles of diameter greater than 100 microns result in slug flow (10), (13). Because of the elutriation problem coupled with cohesion of smaller particles, a particle size of 1020 microns (0.040 in.) was chosen for this design.

Minimum Gas Fluidization Velocity

This is the velocity at which the bed of particles begin to fluidize. Frantz (10) gave a modified equation for predicting the minimum fluidization mass velocity. This equation is shown in the Appendix A1. With this equation, a minimum fluidization velocity of 1.6 ft/sec was calculated for the chosen particle and fluid.

Superficial Gas Velocity

For a bed of solids to expand upward the velocity must exceed the minimum fluidization velocity. This velocity is known as the superficial velocity. In this design, a superficial velocity which is only slightly above the minimum fluidization velocity will be used. The reason for this are as follows:

1. Bed is composed of medium sized particles.
2. Required bed height is small.
3. No net particles transportation is required.

Apart from the above listed conditions operating at the superficial velocity which is slightly in excess of minimum fluidization velocity minimizes elutriation losses. These conditions led to the choosing of a superficial velocity of 4 ft/sec.

Terminal Velocity

Minimum velocity predicts the incipience of fluidization, the onset of bed expansion, characterised by minimum void fraction. As the velocity increases, an extreme will be reached whereby the void fraction will approach a unit value, which corresponds to a single particle being suspended in the bed due to bed expansion. At this stage, the maximum velocity is the same as the velocity at which the particle will fall from a large body of stagnant fluid. This maximum velocity is termed to be the terminal velocity, V_t . Frantz (10) gave an equation for determining this velocity (for the general case, including turbulent flow). This equation is shown in the Appendix A1. With this equation, a terminal velocity of 14.6 ft/sec was obtained.

CHAPTER V

CONCLUSIONS AND RECOMMENDATIONS

The analysis of the pressure drop in a fluidized bed of solid particles with immersed heat transfer tubes was conducted. This analysis indicates that the total pressure drop consists of three parts. These are the pressure drop due to the bed weight, drag force on the heat transfer tubes and wall shear stress. The calculations made from the derived equations give the typical pressure drop due to the bed weight, drag force on the tubes and wall shear stress for a typical tube configuration I (see Appendix B) as 18.33, 0.147 and 0.002 in. of water. The pressure drops for other configurations are shown in Table III, Appendix (p. 104).

The pressure drop with $1\frac{1}{2}$ ins. diameter heat transfer tube due to the bed weight, drag force on the tubes and wall shear stress for a typical configuration 7 (see Appendix C) are 18.33, 0.096 and 0.002 ins. of water. The pressure drops for other configurations are shown in Table V, Appendix (p. 107).

It can be concluded from this analysis that the pressure drop due to the drag on tubes and wall shear did not contribute significantly to the total pressure drop in the fluidized bed with immersed heat transfer tubes and may be considered negligible.

The experimental apparatus designed is rectangular in shape with a dimension of 2 ft x 1 ft x 2 ft. The heat transfer tubes are horizontal with a diameter of 0.5 in. Two opposite walls of the grid are plywood and the other two are plexiglasses. The air distributor

consists of 288 nozzle buttons made from standard size bolts. The pressure drop in the nozzle is 5.6 ins. of water.

For the fluidization, a glass bead particles of 0.040 in. diameter will give a minimum fluidization velocity of 1.60 ft/sec with air as the fluidizing gas. Because the bed is composed of medium sized particles and no net transportation is required, a superficial velocity of 4 ft/sec is chosen. This will give a fluidized bed height of 20 ins. A 480 cfm size fan is required with a superficial velocity of 4 ft/sec. This will give a mass flow rate of 36 lbm/min of air.

This analysis was done for glass beads particle because informations on effective viscosity of the fluidized bed of limestone is not available. It is recommended that in the future if this information is available, the analysis can be carried out using limestone. Also, the pressure drop due to the tubes and wall shear stress are 0.8 percent and 0.011 percent of total pressure drop. In order to measure these pressures, a sensitive manometer will be required however, they may be considered negligible in the design of the fluidized bed.

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APPENDIXES

APPENDIX A

CALCULATIONS OF MINIMUM FLUIDIZATION VELOCITY, SUPERFICIAL
VELOCITY, TERMINAL VELOCITY, SLUG HEIGHT, BED EXPANSION, MASS FLOW
RATE OF AIR, AND MASS OF SOLID PARTICLES

$$D_p = 0.040 \text{ in.} = 0.0033 \text{ ft}$$

$$\rho_s = 170 \text{ lbm/ft}^3$$

$$\mu_{\text{air}} = 1.2321 \times 10^{-5} \text{ lbm/ft}\cdot\text{sec} = 4.45 \times 10^{-2} \text{ lbm/ft}\cdot\text{hr.}$$

$$\rho_{\text{air}} = 0.0754 \text{ lbm/ft}^3$$

$$A_B = \text{Cross sectional area of bed} = (2 \times 1) \text{ ft}^2 = 2 \text{ ft}^2$$

$$P = \text{Perimeter of the bed}$$

$$D_h = \text{Hydraulic diameter of the bed} = \frac{4A}{P} = \frac{4(2)}{2(3)} = 1.33$$

$$\mu_{\text{F.B.}} = 4.55 \mu_{\text{air}} = 5.61 \times 10^{-5} \text{ lbm/ft}\cdot\text{s} = 2.03 \times 10^{-1} \text{ lbm/ft}\cdot\text{hr.}$$

$$L_T = 2 \text{ ft.}$$

$$D_T = 0.5 \text{ in.} = 0.0417 \text{ ft.}$$

Minimum Fluidization Velocity

Fratz(10) gave an expression for minimum fluidization mass velocity as,

$$G_{m_f} = (1.4 \times 10^5) D_p^{1.82} \rho_f (\rho_s - \rho_f)^{0.94} \mu^{0.8} \quad (\text{A-1})$$

$$\text{for } Re \leq 5$$

where

$$D_p = \text{particle diameter in ft.}$$

ρ_f = fluid density in lbm/ft^3

ρ_s = particle density in lbm/ft^3

μ = fluid viscosity in lb/ft.hr

G_{mf} = minimum fluidization mass velocity lb/hr.ft^2

Substituting the fluid and particle values in the above equation.

$$G_{mf} = \frac{(1.4 \times 10^5)(0.0033)^{1.82}(0.0754)(170-0.0754)^{0.94}}{(4.45 \times 10^{-2})^{0.88}}$$

$$= 621 \text{ lb/hr.ft}^2$$

But $G_{mf} = \rho_f V_{mf}$ (A-2)

Where V_{mf} is the minimum fluidization velocity.

Check for Reynolds Number

$$R_e = \frac{D_p G_{mf}}{\mu}$$

$$= \frac{621 (0.0033)}{0.0445}$$

$$= 46$$

Since the Reynolds number is greater than 5, the calculated G_{mf} needs to be corrected. The correction factor is shown in Figure 22.

With Reynolds number of 46 the correction factor is 0.7. Hence,

the minimum fluidization mass velocity is;

$$G_{mf} = 0.7 (621) \text{ lb/hr.ft}^2$$

$$= 435 \text{ lb/hr.ft}^2$$

Using equation (A-2),

$$V_{mf} = \frac{G_{mf}}{\rho_f} = \frac{435 \text{ lb/hr.ft}^2}{0.0754}$$

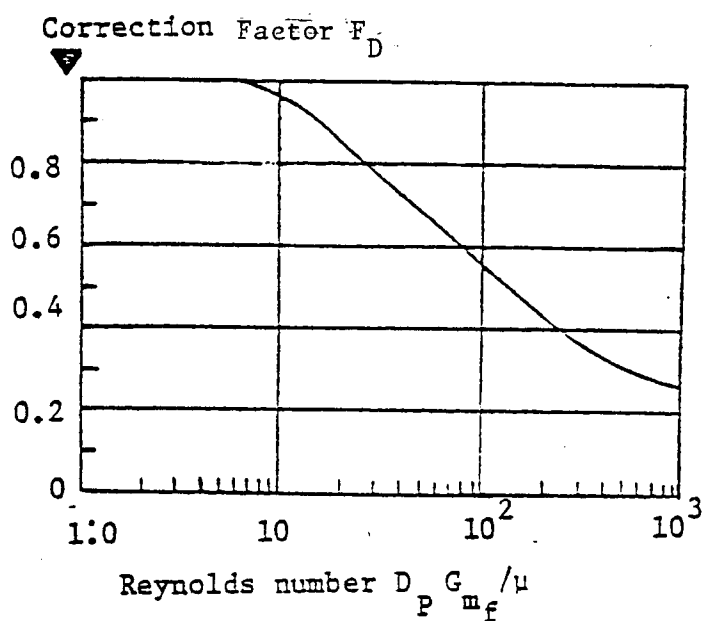


Figure 22. Correction factor for minimum fluidization velocity.

that is,

$$\begin{aligned} V_{mf} &= \frac{435}{0.0754 \text{ hr}} \left(\frac{1 \text{ hr}}{3600 \text{ sec.}} \right) \\ &= 1.60 \text{ ft/sec} \end{aligned}$$

Estimation of Slug Height in the Bed

Frantz (10) reported an expression developed by Yagi and Muchi (21) for estimating slugging-bed height. The expression is,

$$L_{sh} = 1.18 D_t (D_p \rho_s)^{-0.3} \quad (\text{A-3})$$

where D_t is the diameter of the column of the bed. Since in this thesis a rectangular bed is used, the equation can be modified as,

$$L_{sh} = 1.18 D_h (D_p \rho_s)^{-0.3} \quad (\text{A-4})$$

where D_h is the hydraulic diameter of the bed.

Substituting D_L , D_p and ρ_s values in (A-4)

$$\begin{aligned} L_{sh} &= 1.18 (1.33) (0.0033(170))^{-0.3} \\ &= 19.0 \text{ ins.} \end{aligned}$$

Terminal Velocity

Sissoms and Pitts (14) gave an equation for calculating the terminal velocity for a single particle in a viscous flow as fluidized bed. The equation is:

$$V_T = \frac{D_p^2}{18} \left(\frac{\rho_s - \rho_f}{\mu} \right) g_c \quad R_e < 1 \quad (\text{A-5})$$

Leva (3) reported that equation (A-5) can be used for $Re \leq 2$.

For the general case, including turbulent flow, Frantz (10) gave an expression for calculating the terminal velocity as

$$V_t = \frac{4 g_c D_p}{3 \rho_f C_D} (\rho_s - \rho_f)^{1/2} \quad (A-6)$$

where C_D is the drag coefficient and can be correlated as a function of Reynolds number from Figure 23 (1).

Since the calculated Reynolds number is 46, equation, (A-6) is applicable. Hence,

$$V_t = \frac{4 g D_p (\rho_s - \rho_f)^{1/2}}{3 \rho_f C_D}$$

where

g = acceleration due to gravity in ft per (hr)(hr)

D_p = particle diameter in ft

ρ_s = particle (solid) density, lb. per cu. ft.

ρ_f = fluid density, lb per cu. ft.

that is

$$V_t = \left(\frac{4 (32.2) (0.0033) (170 - 0.0754)}{3 (0.0754) C_D} \right)^{1/2}$$

with $Re = 46$, using Figure 23, C_D is approximately equal to 1.5

Therefore,

$$\begin{aligned} V_t &= \left(\frac{4 (32.2) (0.0033) (169.92) (3600)^2}{3 (0.0754) (1.5)} \right)^{1/2} \\ &= 14.6 \text{ ft/sec} \end{aligned}$$

Drag Coefficient C_D

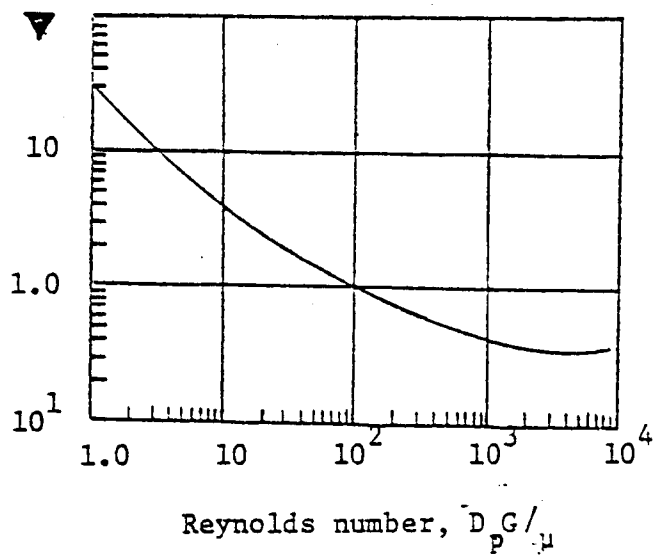


Figure 23. Drag coefficient for single sphere.

that is,

$$V_t = 14.6 \text{ ft/sec}$$

Equation (A-1) gives the minimum velocity for incipient fluidization, the onset of bed expansion, characterised by minimum void fraction (Void Volume per total Volume). As the bed start to expand, the void fraction increases until it approaches a unit value. This corresponds to a stage when single particle is being suspended in the bed. For this case the limiting (maximum) velocity is the same as the velocity which a particle dropped with a large stagnation fluid will acquire. This limiting velocity is called the terminal velocity of the particle in the fluid (see equation (A-6)). The bed is expanded for all velocities between V_{mf} and V_t .

Superficial Velocity

A superficial velocity, V_∞ , of 4 ft/sec will be used in this design calculation. In choosing this value, many factors are considered and they are discussed under the discussion section of the Design of Experimental Apparatus.

Mass Flow Rate of the Fluid

The fluid used in this design calculation is air.

$$\text{Mass Flow Rate, } \dot{m}_{\text{air}} = \rho_f A_B V_\infty \quad (\text{A-7})$$

$$= (0.0754 \text{ gnm/ft}^3) (2 \text{ ft}^2) (4 \text{ ft/sec})$$

$$= 0.60 \text{ lbm/sec}$$

that is,

$$\dot{m}_{\text{air}} = 0.603 \text{ lbm/sec} \left(\frac{3600 \text{ sec}}{1 \text{ hr.}} \right) = 2171.52 \text{ lbm/hr}$$

$$\begin{aligned} \text{Volume Flow Rate} &= A_B V_{\infty} = (2 \text{ ft}^2) (4 \text{ ft/sec}) \\ &= 8 \text{ ft}^3/\text{sec} (60 \text{ sec/min}) \\ &= 480 \text{ ft}^3/\text{min} = 480 \text{ cfm} \end{aligned}$$

Bed Expansion

Using equation given by Nguyen and Bergonguon (22).

$$\frac{H}{H_{m_f}} = 1 + \frac{V_{\infty} - V_{m_f}}{0.36 \sqrt{g D_h}} \quad (\text{A-8})$$

where,

H - fluidized bed height (ft)

H_{m_f} = bed height at minimum superficial gas velocity (ft)

g = acceleration due to gravity, ft per (sec) (sec)

As calculated from equation (A-3), the maximum slugging bed height is 19 in. Let fluidized bed height be 20 in. Using equation (A-8)

$$\frac{H}{H_{m_f}} = 1 + \frac{(4 - 1.6)}{0.36 \sqrt{(32.2) (1.33)}} = 2.02$$

But $H = 20$ ins. hence;

$$H_{m_f} = 9.9 \text{ in.}$$

Minimum Volume of Bed

The minimum volume of the bed can be calculated from the product of the cross sectional area and the minimum height of the bed. That is,

$$\begin{aligned} \text{Volume} &= H_{m_f} \times A_B & (A-9) \\ &= 9.9 \text{ ins } \left(\frac{1 \text{ ft}}{12 \text{ in}} \right) (2 \text{ ft}^2) \\ &= 1.65 \text{ ft}^3 \end{aligned}$$

Volume of Solid

To know the volume of solid, the knowledge of the void fraction is required.

$$\epsilon = \frac{V_{\text{total}} - V_{\text{solid}}}{V_{\text{total}}} \quad (A-10)$$

Void fraction of the particle (assume spherical) is 0.32 (14).

Hence, the volume of the solid, V_{solid} is,

$$\begin{aligned} V_{\text{solid}} &= 1.65 \text{ ft}^3 - 0.32 (1.65 \text{ ft}^3) \\ &= 1.12 \text{ ft}^3 \end{aligned}$$

Mass of Solids

Knowing the volume of the solids in the bed, the mass is calculated as follows;

$$\begin{aligned} m_s &= \rho_s V_{\text{solid}} & (A-11) \\ &= 170 \frac{\text{lbs}}{\text{ft}^3} (1.122 \text{ ft}^3) \\ &= 191 \text{ lbm} \end{aligned}$$

Expanded Bed Density

The bed density is related to the expanded bed height by

$$\rho = W / H_{AB} \quad (A-12)$$

where W is the total mass of the particles in the bed. Hence,

$$\rho = \frac{191 \text{ lbm}}{(23/12 \text{ ft})(2 \text{ ft}^2)} = 50 \text{ lbm/ft}^3$$

PRESSURE DROP IN AIR DISTRIBUTOR PLATE

$$\dot{m}_f = 0.60 \text{ lbm/sec}$$

$$\begin{aligned} \text{Velocity of inlet air} &= \frac{\dot{m}_f}{A \rho_f N} & (A-13) \\ &= \frac{0.60 \text{ lbm/sec}}{\pi/4 \left(\frac{1}{12}\right)^2 (0.0754) (288)} \text{ lbm/ft} \\ &= 81.5 \text{ ft/sec} \end{aligned}$$

where N = number of nozzles required = 288 (see Figure 24)

$$\begin{aligned} \text{Velocity of outlet air} &= \frac{0.60 \text{ lbm/sec}}{(0.0754 \text{ lbm/ft}^3) \left(\frac{\pi}{4} \left(\frac{1}{12}\right)^2 \text{ ft}^2/\text{hole}\right) \times} \\ &\quad \frac{0.60 \text{ lbm/sec}}{(8 \text{ holes}) (288 \text{ bolts})} = 163 \text{ ft/sec} \end{aligned}$$

The governing equation is the energy equation,

$$\frac{V_1^2}{2g_c} + \frac{P_1}{\rho_f} + \frac{g}{g_c} Y_1 = \frac{V_2^2}{2g_c} + \frac{P_2}{\rho_f} + \frac{g}{g_c} Y_2 + h_L \quad (A-14)$$

where h_L is the head loss due to the pipe friction and contraction.

$$\therefore P_1 - P_2 = \rho_f \frac{(V_2^2 - V_1^2)}{2g_c} + \rho_f (Y_2 - Y_1) \frac{g}{g_c} + \rho_f h_L$$

Because ρ_f is very small, and the bolt is short (see Figure 24),

$\rho_f (Y_2 - Y_1) \frac{g}{g_c}$ can be considered negligible. Hence,

$$P_1 - P_2 = \rho_f \frac{(V_2^2 - V_1^2)}{2g_c} + \rho_f h_L \quad (A-15)$$

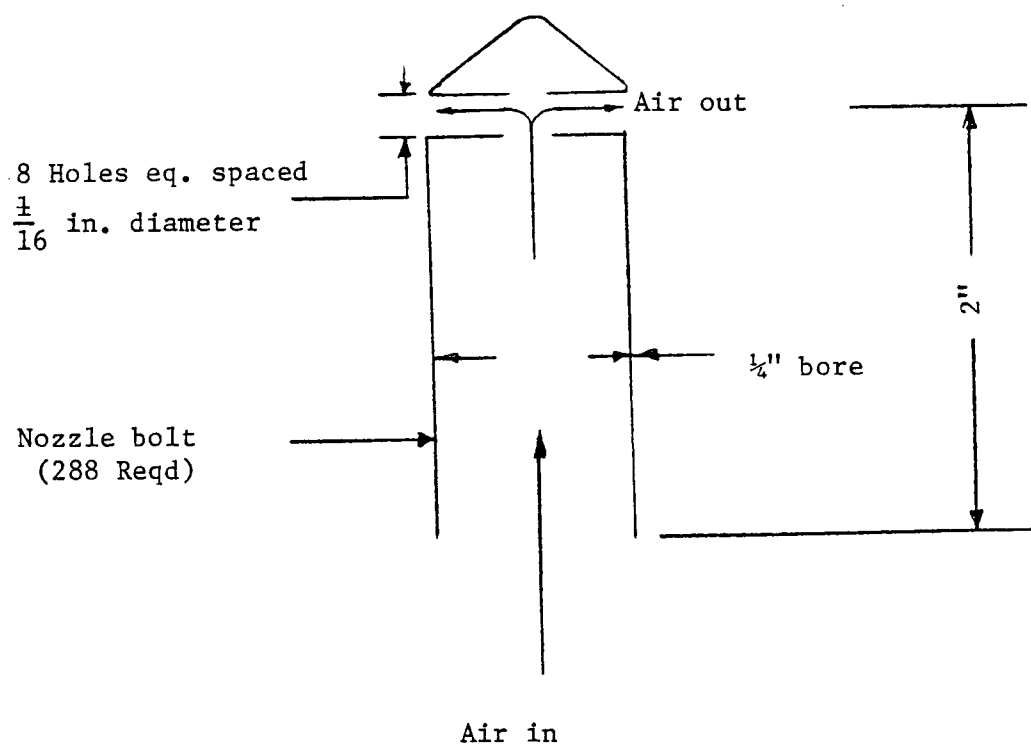


Figure 24. Schematic diagram for calculating pressure drop in the nozzle button and air distribution grid.

$$\begin{aligned} \frac{(v_2^2 - v_1^2)}{2g_c} \rho_f &= \frac{((162.9)^2 - (81.46)^2) \text{ ft}^2/\text{sec}^2}{64.4 \frac{\text{lbm} \cdot \text{ft}}{\text{lbf} \cdot \text{s}^2}} \cdot 0.0754 \frac{\text{lbf}}{\text{ft}^3} \\ &= 23.3 \frac{\text{lbf}}{\text{ft}^2} = 0.16 \text{ psi} \end{aligned}$$

The flow Reynolds number is:

$$R_e = \frac{VD\rho_f}{\mu_f} = \frac{81.49 (\frac{1}{4}/12) (0.0754)}{1.2321 \times 10^{-5}}$$

Using this Reynolds number and assuming the bore is smooth, the friction factor, f is 0.03 (14). Therefore, the pressure loss due to friction is,

$$\begin{aligned} \Delta P_{\text{friction}} &= f \frac{L}{D} \rho_f \frac{v^2}{2g_c} \quad (\text{A-16}) \\ &= 0.03 \left(\frac{2}{1/4}\right) (0.0754) \frac{(81.49)^2}{64.4} \\ &= 1.86 \frac{\text{lbf}}{\text{ft}^2} = 0.01 \text{ psi} \end{aligned}$$

The loss due to the contraction of the bore from 1/4 in. to 1/16 in. is,

$$\Delta P_{\text{contraction}} = \rho_f (h_L)_{\text{contraction}} = \rho_f K_L \frac{v^2}{2g_c} \quad (\text{A-17})$$

where K_L is the contraction loss and is given by (14) as 0.5

$$\begin{aligned} \therefore \Delta P_{\text{contraction}} &= (0.0754)(0.5) \frac{(81.46)^2}{64.4} = 3.89 \frac{\text{lbf}}{\text{ft}^2} \\ &= 0.03 \text{ psi} \end{aligned}$$

Hence total pressure drop in the bolts (air distribution plate) is,

$$\Delta P = (0.16 + 0.01 + 0.03) \text{ psi} = 0.20 \text{ psi}$$

$$= 0.20 \frac{\text{lbf}}{\text{in}^2} \frac{(27.673 \text{ in. of water})}{1 \frac{\text{lbf}}{\text{in}^2}}$$

$$\Delta P = 5.6 \text{ in. of water}$$

MAGNITUDE COMPARISON OF EQUATION (7)

Examining the magnitude of the pressure terms in equation (7)

$$P_1 A_1 - P_2 A_2 + \frac{\dot{m}_1 V_1}{g_c} - \frac{\dot{m}_2 V_2}{g_c} - \tau_w A_{\text{surface wall}} - W_{\text{bed}} - F_D = 0$$

$$\text{By continuity, } \dot{m}_1 = \dot{m}_2 = \rho_1 A_1 V_1 = \rho_2 A_2 V_2 = \dot{m} \quad (\text{A-18})$$

$$A_1 = A_2 \quad \text{Physical Boundary Condition}$$

$$\therefore \rho_1 V_1 = \rho_2 V_2 \quad (\text{A-19})$$

$$\text{Define } P_1 - P_2 = \Delta P \quad (\text{A-20})$$

The third and fourth terms in equation (7) can be written as,

$$\frac{\dot{m}_1 V_1}{g_c} - \frac{\dot{m}_2 V_2}{g_c} = \frac{\dot{m}}{g_c} (V_1) \left(1 - \frac{V_2}{V_1}\right)$$

$$\text{From equation (A-19), } \frac{V_2}{V_1} = \frac{\rho_1}{\rho_2} = \frac{P_1 / RT_1}{P_2 / RT_2} = \frac{T_2}{T_1} \left(\frac{P_1}{P_2}\right)$$

$$\text{But } T_2 = T_1 \quad (\text{isothermal bed})$$

$$\therefore \frac{V_2}{V_1} = \frac{P_1}{P_1 - \Delta P}$$

that is,

$$\frac{\dot{m}}{g_c} (V_1) \left(1 - \frac{V_2}{V_1}\right) = \frac{\dot{m}}{g_c} (V_1) \left(1 - \frac{P_1}{(P_1 - \Delta P)}\right) \quad (\text{A-21})$$

Combining with terms 1 and 2,

$$P_1 A_1 (1 - P_2/P_1) + \frac{\dot{m}}{g_c} (V_1) \left(1 - \frac{P_1}{(P_1 - \Delta P)}\right)$$

$$P_1 A_1 \left(1 - \left(\frac{P_1 - \Delta P}{P_1}\right)\right) + \frac{\dot{m} V_1}{g_c} \left(1 - \frac{P_1}{(P_1 - \Delta P)}\right)$$

$$P_1 A_1 \left(\frac{\Delta P}{P_1}\right) + \frac{\dot{m} V_1}{g_c} \left(1 - \frac{\Delta P}{(P_1 - \Delta P)}\right)$$

$$A_1 \Delta P + \frac{\dot{m} V_1}{g_c} \left(\frac{-\Delta P}{(P_1 - \Delta P)}\right)$$

$$\text{But } \dot{m} = \dot{m}_1 = A_1 \rho_1 V_1$$

$$\rho = \frac{P}{RT}$$

$$A_1 \Delta P + \rho_1 \frac{A_1 V_1^2}{g_c} \left(\frac{-\Delta P}{(P_1 - \Delta P)}\right)$$

$$A_1 \Delta P + P_1/RT_1 \frac{A_1 V_1^2}{g_c} \left(\frac{-\Delta P}{(P_1 - \Delta P)}\right)$$

$$A_1 \Delta P + \frac{A_1 V_1^2}{g_c} \frac{1}{RT_1} \left(\frac{-\Delta P}{1 - \frac{\Delta P}{P_1}}\right)$$

$$A_1 \Delta P \left(1 - \frac{V_1^2}{RT_1 g_c} \left(\frac{1}{1 - \frac{\Delta P}{P_1}}\right)\right)$$

For this thesis, the fluidizing velocity is 4^{ft}/sec and the fluidizing gas is air,

$$V_1 = 4^{\text{ft/sec}}$$

$$R = \text{Gas Constant} = 53.34 \frac{\text{ft} - \text{lb f}}{\text{lbm } ^\circ\text{R}}$$

$$T_1 = \text{Ambient Temperature} = 70^\circ \text{F} = 530^\circ\text{R}$$

$$g_c = 32.17 \frac{\text{lbm} - \text{ft}}{\text{lb f} - \text{sec}^2}$$

Substituting these values in the above,

$$A_1 \Delta P \left(1 - \frac{16 \frac{\text{ft}^2}{\text{sec}^2}}{53.34 \frac{\text{ft-lbf}}{\text{lbm}^{\circ}\text{R}} \times 530^{\circ}\text{R} \times 32.17 \frac{\text{lbm-ft}}{\text{lbf-sec}^2}} \left(\frac{1}{1 - \frac{\Delta P}{P_1}} \right) \right)$$

$$A_1 \Delta P \left(1 - 0.0000176 \left(\frac{1}{1 - \frac{\Delta P}{P_1}} \right) \right)$$

Since $\Delta P < P_1$, hence $1 - \frac{\Delta P}{P_1}$ is in the order of 1.

This shows that in $\frac{\dot{m}_1 V_1}{g_c} - \frac{\dot{m}_2 V_2}{g_c} \ll 1$ for the range of the

fluidizing velocity used in this thesis. Therefore these terms are negligible. Therefore, the analysis is continued from equation (8).

APPENDIX B

APPENDIX B

CALCULATION OF PRESSURE DROPS

Pressure Loss Due to the Bed Weight

Pressure drop across the bed area A_B , is equal to the weight per unit area of the particle. This is because, at the on-set fluidization, the drag exerted on the particle by the gas flow supports the weight of the bed.

Hence,

$$\begin{aligned}\Delta P &= \frac{m_s (\rho_s - \rho_f) g}{\rho_s A_B g_c} & (B-1) \\ &= \frac{191. (170-0.0754) \frac{\text{lbm}}{\text{ft}^3} (32.2 \frac{\text{X}}{\text{s}^2})}{170 \frac{\text{lbm}}{\text{ft}^3} (2 \text{ ft}^2)}\end{aligned}$$

that is,

$$\begin{aligned}\Delta P &= 3069.5 \frac{\text{lbm}}{\text{ft} \cdot \text{sec}^2} \cdot \frac{\text{lbf} \cdot \text{sec}^2}{32.2 \text{ lbm} \cdot \text{ft}} \cdot \frac{1 \text{ ft}^2}{144 \text{ in}^2} \\ \Delta P &= 0.66 \frac{\text{lbf}}{\text{in}^2} \\ &= 0.66 \frac{\text{lbf}}{\text{in}^2} \left(\frac{27.673 \text{ in of H}_2\text{O}}{\text{lbf/in}^2} \right) \\ \Delta P &= 18.3 \text{ ins of H}_2\text{O}\end{aligned}$$

Pressure Loss Due to Tubes Alone and Air Flowing Across

The pressure loss due to air flowing through tube banks (in-line) of 24 rows and 12 columns. The tubes are 1-in. centers. (see Figure 25).

Holman (15) gives an equation for calculating pressure drop for flow over a tube banks as,

$$\Delta P = \frac{f_{G_{\max}}^2 N}{\rho (2.09 \times 10^8)} \left(\frac{\mu_w}{\mu_b} \right)^{0.14} \quad (B-2)$$

Configuration 1

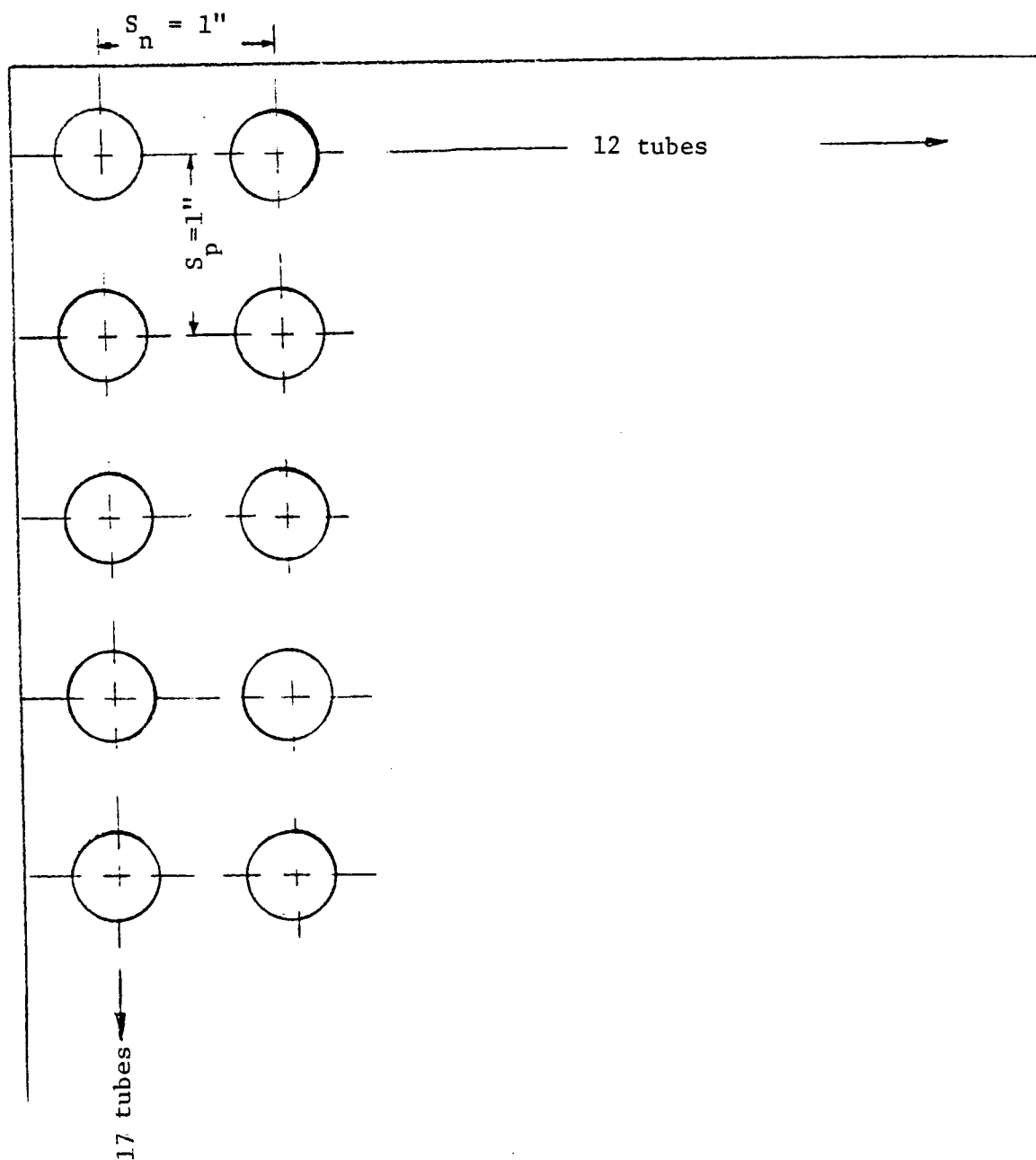


Figure 25. In-line distribution - 1 in. centers both sides.

where,

$G_{\max}$ = mass velocity at minimum flow area, lbm/hr.ft^2

ρ = density evaluated at freestream conditions, lbm/ft^3

N = number of transverse rows

The empirical friction factor f^1 is given as

$$f^1 = \left(0.044 + \frac{0.08 S_p/d}{((S_n-d)/d)^{0.43+1.13d/S_p}} \right) R_{e_{\max}}^{-0.15} \quad (\text{B-3})$$

where

S_p = frontal area parallel to the flow, in.

S_n = frontal area normal to the flow, in.

d = diameter of the tube in in.

To calculate the maximum velocity of the air flowing through the tube banks, we need to determine the minimum flow area. Using the in-line tubes of Figure 25.

The minimum flow area = $S_n - d$

Frontal area = S_n

Ratio of minimum flow area to the frontal area = $\frac{S_n - d}{S_n}$ (B-4)

Hence, maximum velocity = $V_{\max} = \frac{V S_n}{(S_n - d)}$ (B-4)

$$= \frac{4 \text{ ft/s (1/12 ft)}}{(1/12 - 0.0417) \text{ ft}}$$

that is, $V_{\max} = 8.01 \text{ ft/s}$

$$\begin{aligned} R_{e_{\max}} &= \frac{V_{\max} d}{V_f} = \frac{(8.01 \text{ ft/s})(0.0417 \text{ ft})}{0.165 \times 10^{-3} \text{ ft}^2/\text{s}} \\ &= 2.02 \times 10^3 \end{aligned}$$

$$\begin{aligned}
 f^1 &= (0.044 + \frac{0.08 S_p/d}{((S_n-d)/d)^{0.43+1.13d/S_p}}) R_{e_{\max}}^{-0.15} \\
 &= (0.044 + \frac{0.08 (1/0.5)}{((1-0.5/0.5)^{0.43+1.13(0.5)/1})}) (2.02 \times 10^3)^{-0.15} \\
 &= 0.065
 \end{aligned}$$

$$\begin{aligned}
 G_{\max} &= \rho V_{\max} = 0.0754 \text{ lbm/ft}^3 (8.01 \text{ ft/s}) (\frac{3600 \text{ sec}}{1 \text{ hr}}) \\
 &= 2174.23 \text{ lbm/hr.ft}^2
 \end{aligned}$$

$$\Delta P = \frac{f^1 G_{\max}^2 N}{\rho_f (2.09 \times 10^8) \mu_b} \left(\frac{\mu_w}{\mu_b} \right)^{0.14}$$

In this design calculation, an atmospheric temperature of 70° F is assumed. The tubes are also assumed to be at this temperature. Therefore, μ_w , and μ_b are equal. Hence,

$$\Delta P = \frac{f^1 G_{\max}^2 N}{\rho_f (2.09 \times 10^8)}$$

Substituting values from above and knowing $N = 24$ from figure 25,

$$\begin{aligned}
 \Delta P &= \frac{0.065 (2174.23)^2 (17)}{(0.0754)(2.09 \times 10^8)} \\
 &= 0.332 \text{ lbf/ft}^2
 \end{aligned}$$

$$\begin{aligned}
 \Delta P_{\text{air}} &= 0.0023 \text{ lbf/in}^2 \\
 &= 0.06 \text{ in. of water}
 \end{aligned}$$

$$\Delta P = 0.06 \text{ in. of water}$$

Knowing ΔP_{air} , equation (43) will be used to calculate the pressure loss due heat transfer tubes

$$\Delta P_{F.B(tubes)} = \frac{C_{D_{F.B}}}{C_{D_{gas}}} \Delta P_{gas} \quad (43)$$

$$R_{e_{F.B.}} = \frac{V_{max} D_t \rho_f}{\mu_{F.B.}} = \frac{(8.01)(0.0417)(0.0754)}{5.61 \times 10^{-5}} = 448.57$$

Therefore,, $C_{D_{F.B}} \approx 1.2$

$$R_{e_{air}} = \frac{V_{max} D_t \rho_f}{\mu_{air}} = \frac{8.01 (0.0417)(0.0754)}{12321 \times 10^{-5}} = 2044.1$$

Therefore, $C_{D_{air}} \approx 1.0$

Hence,

$$\begin{aligned} \Delta P_{F.B(tubes)} &= \left(\frac{1.2}{1.0} (0.06) \right) \text{ ins. of water} \\ &= 0.072 \text{ ins. of water} \end{aligned}$$

Using equation (42),

$$\begin{aligned} \Delta P_{tubes} &= (D_{C L T} C_{D_{F.B}} \frac{V_{max}^2}{2} \rho_f N) \frac{1}{A_B g_c} \text{ where } N = 12 \times 17 = 204 \\ &= \left(\frac{0.5}{12} (2) (1.2) \frac{(8.01)^2}{2} (0.0754) (204) \right) \frac{1}{2} \left(\frac{1}{32.2} \right) \left(\frac{1}{144} \right) \text{ psi} \\ \Delta P_{tubes} &= 0.147 \text{ in. of water} \end{aligned}$$

Pressure Loss Due to Staggered Tubes Distribution Using 1 in. Centers on Both Sides

The tubes distribution are as shown on Figure 26.

From equation (B-4), $V_{max} = 8.01 \text{ ft/sec.}$

$$\therefore R_{e_{max}} = 2.02 \times 10^3$$

$$G_{max} = 2174 \text{ lbm/hr.ft}^2$$

The pressure loss equation is given by equation (B-2). Holman (15) expressed the empirical friction factor f^1 for staggered tubes as,

Configuration 2

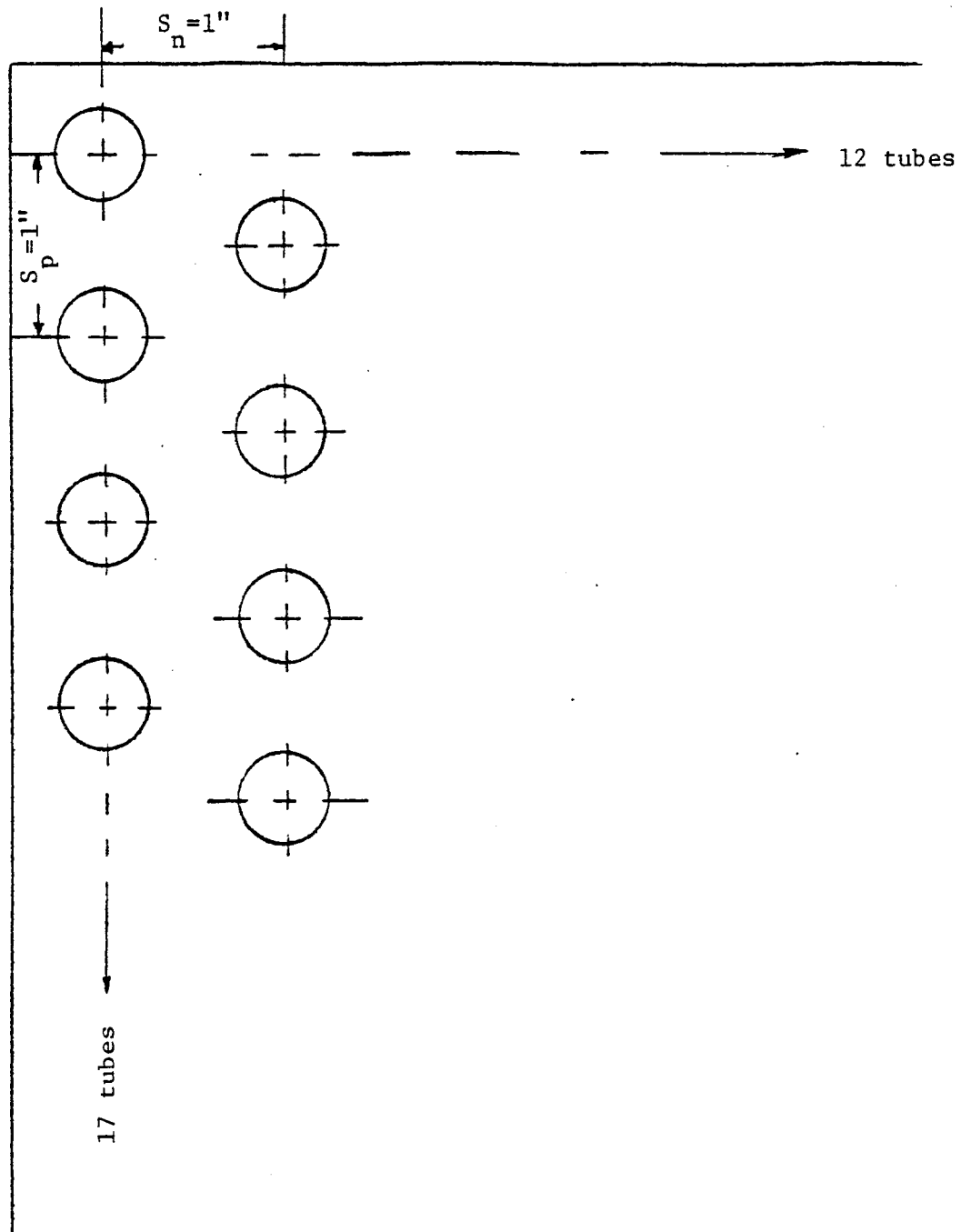


Figure 26. Staggered distribution - 1 in. centers on both sides.

$$f^1 = (0.25 + \frac{0.118}{((S_n - d)/d)^{1.08}}) R_{e_{\max}}^{-0.16} \quad (B-5)$$

Putting values of S_n , d , and $R_{e_{\max}}$ in (B-5),

$$\begin{aligned} f^1 &= (0.25 + \frac{0.118}{((1-0.5)/0.5)^{1.08}}) (2.02 \times 10^3)^{-0.16} \\ &= 0.109 \end{aligned}$$

Putting f^1 in (B-2) and solved for pressure loss,

$$\Delta P = \frac{0.109 (2174.23)^2 (17)}{(0.0754)(2.09 \times 10^8)}$$

that is,

$$\begin{aligned} \Delta P &= 0.556 \text{ lbf/ft}^2 \\ &= 0.004 \text{ lbf/in}^2 \end{aligned}$$

$$\Delta P = 0.11 \text{ in. of water}$$

$$\text{i.e } \Delta P_{\text{air}} = 0.11 \text{ in. of water}$$

$$\text{But } \Delta P_{\text{F.B(tubes)}} = \frac{C_{D_{\text{F.B}}}}{C_{D_{\text{air}}}} \Delta P_{\text{air}} \quad (43)$$

Using Figure 7, page 21, to estimate $C_{D_{\text{F.B}}}$ and $C_{D_{\text{Air}}}$,

$$\begin{aligned} \Delta P_{\text{F.B(tubes)}} &= (1.2(0.11)) \text{ in. of water} \\ &= 0.132 \text{ in. of water} \end{aligned}$$

Using equation (42)

$$\Delta P_{\text{tubes}} = (D_C L_T C_{D_{\text{F.B}}}) \frac{V_{\max}^2}{2} \rho_f N \frac{1}{A_B g_c} \left(\frac{27.673}{144} \right), \quad N = 204$$

$$\text{i.e } \Delta P = 0.147 \text{ in. of water}$$

Pressure Loss Due to In-Line Tube Distribution Using 2 in. Centers
on Rows and 1 in. Centers on Columns

The tube distributions are as shown in Figure 27. From equation (B-4),

$$V_{\max} = \frac{V_{\infty} S_n}{S_n - d} = 4 \text{ ft/s} \frac{1/12}{(1/12 - 0.5/12)} = 8 \text{ ft/s}$$

$$G_{\max} = V_{\max} = 0.0754 \text{ lbm/ft}^3 (8 \text{ ft/s}) \left(\frac{3600 \text{ s}}{1 \text{ hr}} \right) \\ = 2171.52 \text{ lbm/hr.ft}^2$$

$$R_{e_{\max}} = 2.02 \times 10^3$$

From equation (B-3), the friction factor for in-line tube arrangements is,

$$f^1 = (0.044 + \frac{0.08 (2/0.5)}{((1-0.5)/0.5)^{0.43+1.13(0.5)/2}}) (2.02 \times 10^3)^{-0.15} \\ = 0.116$$

From equation (B-2) with $N = 12$,

$$\Delta P = \frac{0.116 (2171.52)^2 (9.)}{(0.0754)(2.09 \times 10^8)} \\ = 0.313 \text{ lbf/ft}^2 \\ = 0.002 \text{ lbf/in}^2$$

$$\Delta P = 0.06 \text{ ins. of water}$$

that is, $\Delta P_{\text{air}} = 0.06 \text{ ins. of water}$

Using equation (43) and Figure 7, page 21, to estimate $C_{D_{F.B}}$ and $C_{D_{\text{Air}}}$,

$$\Delta P_{F.B(\text{tubes})} = (1.2(0.06)) \text{ in. of water} \\ = 0.072 \text{ in. of water}$$

Using equation (42),

Configuration 3

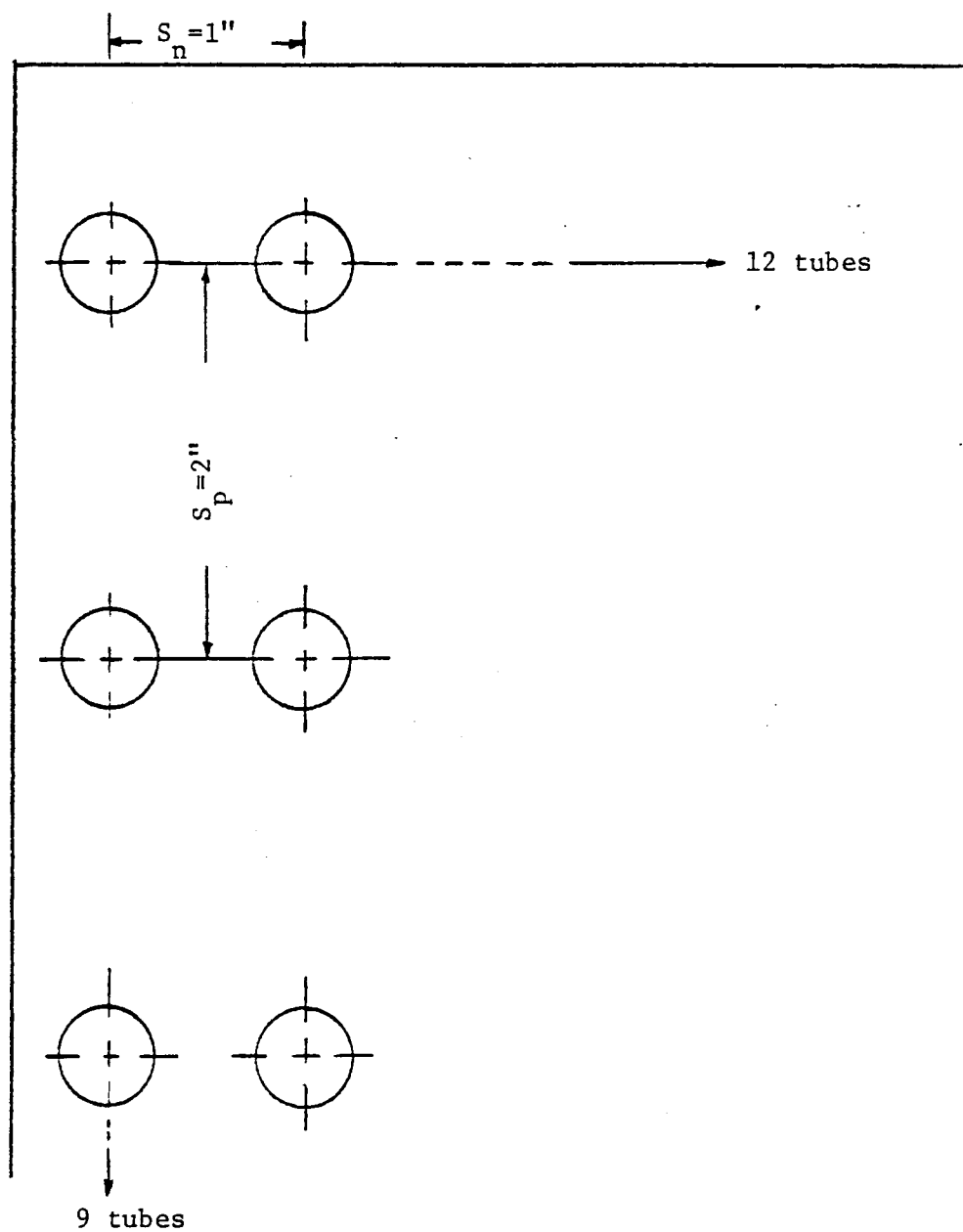


Figure 27. In-line distribution - 2 in. centers on rows and 1 in. centers on columns.

$$\begin{aligned}\Delta P_{\text{tubes}} &= (D_C L_T C_{D_{F.B}} \frac{V_{\text{max}}^2}{2} \rho_f N) \frac{1}{A_B g_c} \left(\frac{27.673}{144} \right), N = 108 \\ &= \left(\frac{0.5}{12} (2) (1.2) \left(\frac{8.01}{2} \right)^2 (0.0754) (108) \right) \frac{1}{2(32.2)} \left(\frac{1}{144} \right) \text{psi}\end{aligned}$$

i.e $\Delta P = 0.078$ in. of water

Pressure Loss Due to Staggered Tube Distribution Using 2 in. Centers
On Rows and 1 in. Centers on Columns

The tube arrangements are as shown in Figure 28. Using equation (B-4)

$$V_{\text{max}} = 8 \text{ ft/s}$$

$$\text{Therefore, } G_{\text{max}} = 2171.52 \frac{\text{lbm}}{\text{hr.ft}^2}$$

$$R_{e_{\text{max}}} = 2.02 \times 10^3$$

Using equation (B-5), the friction factor for staggered tube arrangements is

$$\begin{aligned}f^1 &= \left(0.25 + \frac{0.118}{((1-0.5)/0.5)^{1.08}} \right) (2.02 \times 10^3)^{-0.16} \\ &= 0.109\end{aligned}$$

From equation (B-2), with $N = 12$, the pressure loss is,

$$\begin{aligned}\Delta P &= \frac{0.109 (2171.52)^2 (9)}{0.0754 (2.09 \times 10^8)} \\ &= 0.294 \frac{\text{lb f}}{\text{ft}^2} \\ &= 0.002 \frac{\text{lb f}}{\text{in}^2} \\ &= 0.055 \text{ in. of water}\end{aligned}$$

that is, $\Delta P_{\text{air}} = 0.055$ in. of water

Using equation (43) and figure 7 for $C_{D_{F.B}}$ and $C_{D_{\text{air}}}$, Figure 7, page 21,

$$\Delta P_{F.B(\text{tubes})} = 1.2(0.055) = 0.066 \text{ in. of water}$$

Using equation (42)

$$\Delta P_{F.B(\text{tubes})} = 0.0028 \frac{\text{lb f}}{\text{in}^2} = 0.078 \text{ in. of water, } N = 108$$

Configuration 4

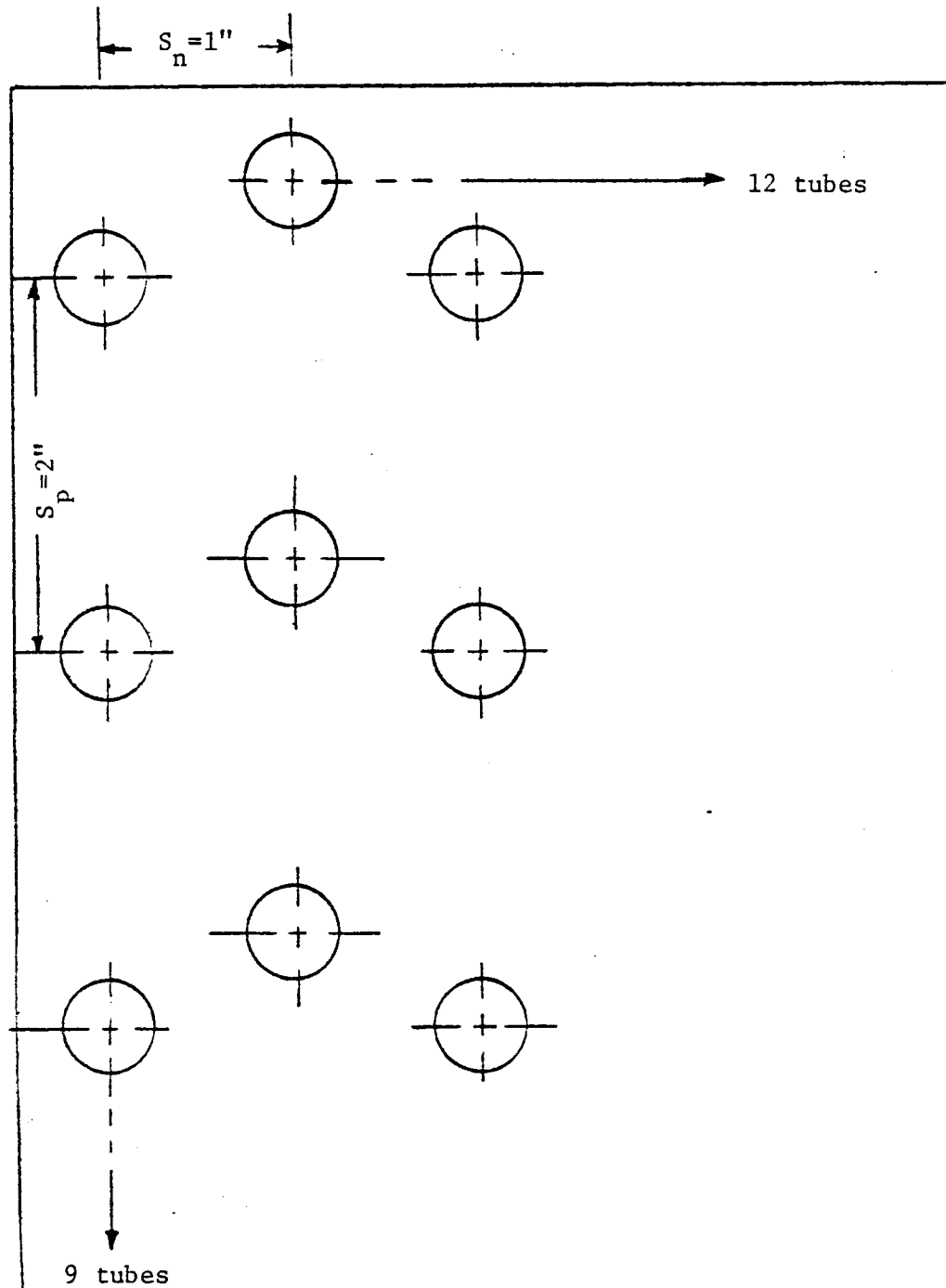


Figure 28. Staggered distribution - 2 in. centers on rows and 1 in. centers on columns.

Pressure Loss Due to In-Line Distribution Using 1 in. Centers on Rows
and 2 in. Centers on Columns

The tube arrangements are as shown in Figure 29 from equation (B-4)

$$V_{\max} = V_{\infty} \frac{S_n}{S_n - d} = 4 \text{ ft/s} \frac{2}{(2-0.5)} = 5.33 \text{ ft/s}$$

Therefore,

$$\begin{aligned} G_{\max} &= \rho_f V_{\max} = (0.0743 \text{ lbm/ft}^2)(5.33 \text{ ft/s}) \left(\frac{3600 \text{ s}}{1 \text{ hr}} \right) \\ &= 1446.8 \text{ lbm/ft}^2 \text{ hr} \\ R_{e_{\max}} &= \frac{V_{\max} d}{\nu_f} = \frac{(5.33 \text{ ft/s})(0.0417 \text{ ft})}{0.165 \times 10^{-3} \text{ ft}^2/\text{s}} \\ &= 1.35 \times 10^3 \end{aligned}$$

From equation (B-3), the empirical friction factor is,

$$\begin{aligned} f^1 &= (0.044 + \frac{0.08 (1/0.5)}{((2-0.5)/0.5)^{0.43+1.13(0.5)/1}}) (1.35 \times 10^3)^{-0.15} \\ &= 0.033 \end{aligned}$$

From equation (B-2), the pressure loss is,

$$\begin{aligned} \Delta P &= \frac{0.033 (1446.8)^2 (17)}{(0.0754)(2.09 \times 10^8)} \\ &= 0.075 \text{ lbf/ft}^2 \\ &= 5.18 \times 10^{-4} \text{ lbf/in}^2 \end{aligned}$$

i.e. $\Delta P_{\text{air}} = 0.014 \text{ in. of water}$

Using equation 43,

$$\begin{aligned} \Delta P_{\text{F.B(tubes)}} &= (0.014)(1.5) \text{ in. of water} \\ &= 0.021 \text{ in. of water} \end{aligned}$$

Configuration 5

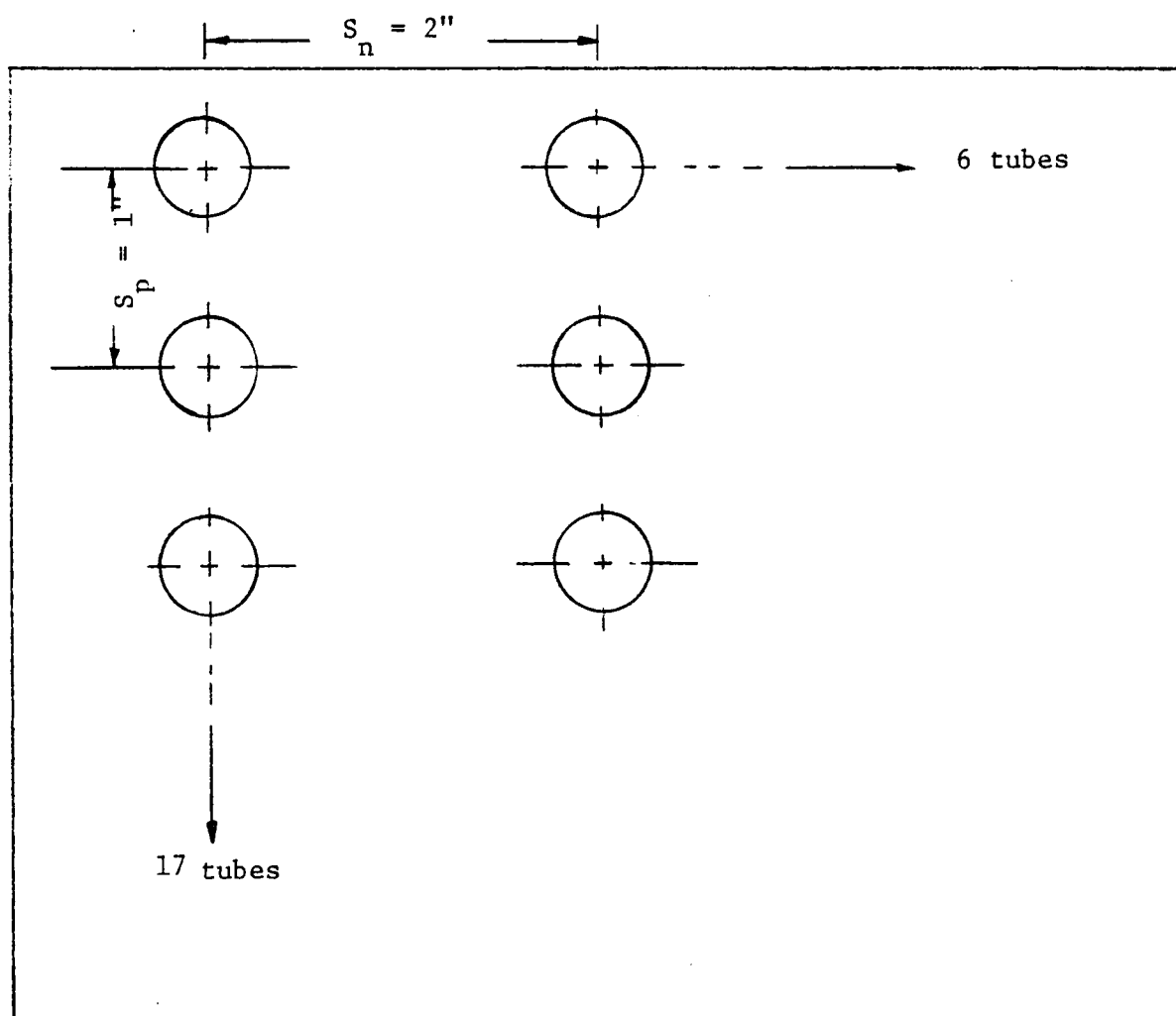


Figure 29. In-line distribution - 1 in. centers on rows and 2 in. centers on columns.

Also from equation (42)

$$\Delta P_{\text{tubes}} = (D_{C,T} L_{C,D_{F,B}} \frac{V_{\text{max}}^2}{2} \rho_f N) \frac{1}{A_B} g_c$$

where $V_{\text{max}} = 5.33 \text{ ft/s}$, $N = 144$, $C_{D_{F,B}} = 1.5$

that is,

$$\begin{aligned} \Delta P_{\text{tubes}} &= \left(\frac{0.5}{12} (2) (1.5) \frac{(5.33)^2}{2} (0.0754) (102) \right) \frac{1}{2(32.2)} \left(\frac{27.673}{144} \right) \\ &= 0.041 \text{ in. of water} \end{aligned}$$

Pressure Loss Due to Staggered Tube Distribution Using 1 in. Centers
On Rows and 2 in. Centers on Columns

The tube arrangements are as shown in Figure 30. From equation (B-4), $V_{\text{max}} = 5.33 \text{ ft/s}$

Therefore, $G_{\text{max}} = 1446.8 \text{ lbm/hr.ft}^2$

$$R_{e_{\text{max}}} = 1.35 \times 10^3$$

From (B-5), the empirical friction factor, f^1 for staggered is

$$\begin{aligned} f^1 &= \left(0.25 + \frac{0.118}{((2-0.5)/0.5)^{1.08}} \right) (1.35 \times 10^3)^{-0.16} \\ &= 0.0903 \end{aligned}$$

Using (B-2) with $N = 17$, the pressure loss is

$$\begin{aligned} \Delta P &= \frac{0.0903 (1446.8)^2 (17)}{(0.0754) (2.09 \times 10^8)} \\ &= 0.204 \text{ lbf/ft}^2 \\ &= 0.0014 \text{ lbf/in}^2 \end{aligned}$$

$$\Delta P_{\text{air}} = 0.039 \text{ in. of water}$$

Hence, from equation (43),

$$\Delta P_{F.B(\text{tubes})} = 0.039 (1.5) = 0.059 \text{ in. of water}$$

Also from equation (42)

$$\Delta P_{\text{tubes}} = 0.041 \text{ in. of water, } N = 102. \text{ For comparison of equations (42)}$$

and (43) see Table II.

Configuration 6

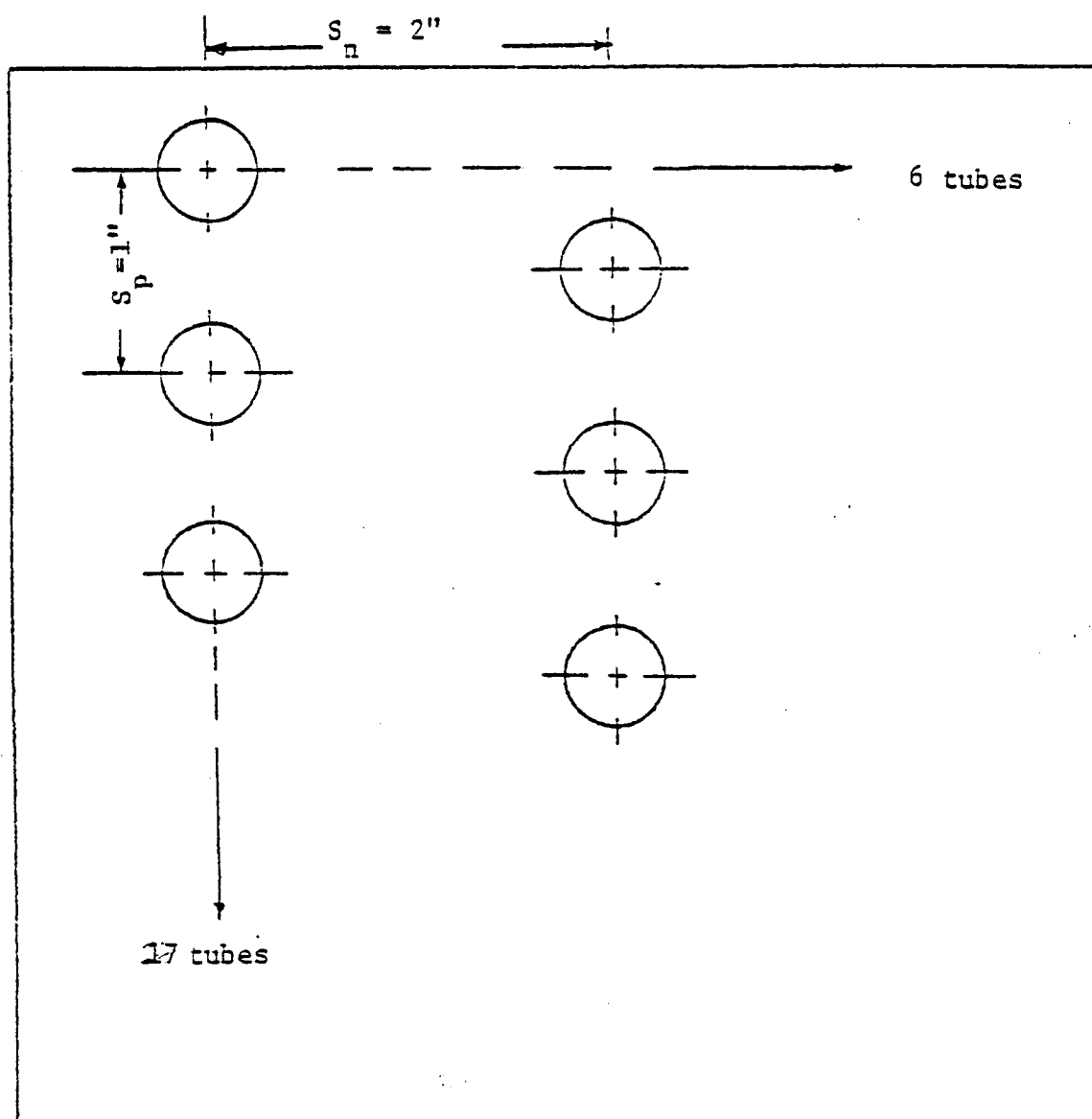


Figure 30. Staggered distribution - 1 in. centers on rows and 2 in. centers on columns.

Table II

Comparison of ΔP_{tubes} From Equation (42) & (43) for
Different Tube Arrangements

Configurations	N	Tube Arrangements	ΔP_{tubes} (in. of water) Equation (42)	$\Delta P_{\text{F.B(tubes)}}$ (in. of water) Equation (43)
1	204	In Line	0.147	0.072
2	204	Staggered	0.147	0.132
3	108	In Line	0.078	0.072
4	108	Staggered	0.078	0.066
5	102	In Line	0.041	0.021
6	102	Staggered	0.041	0.039

APPENDIX C

APPENDIX C

CALCULATION OF TOTAL PRESSURE DROP

From equation (30a), the total pressure drop is

$$\Delta P_{\text{total across tubes}} = P_{w/a} + P_{\text{wall}} + P_{\text{tube}} \quad (30a)$$

where,

$$\Delta P_{w/a} = m_s g \frac{(P_s - P_f)}{P_s A_B}$$

$$\Delta P_{\text{wall}} = 0.664 \frac{V_{\text{max}}}{A_g} \frac{\mu_{F.B}}{\mu_f} \frac{V_{\text{max}} P_s \mu_f}{H} (2(B+L)H)$$

ΔP_{tube} can be represented by either equation (42) or (43). In order to compare the effect of equation (42) and (43) on total pressure drop, each will be used to calculate the total pressure drop across the bed for different number of tubes.

Calculation I (Using equation (42) for P_{tube})

$$\begin{aligned} \Delta P_{\text{total}} = & \frac{1}{A_B} (m_s g \frac{(\rho_s - \rho_f)}{\rho_s} + 0.664 V_{\text{max}} \frac{\mu_{F.B}}{\mu_f} \sqrt{\frac{V_{\text{max}} \rho_f \mu_f}{H}} (2(B+L)H) \\ & + D_{TT} L_T C_{D_{F.B}} \frac{V_{\text{max}}^2}{2} \rho_f N) \end{aligned} \quad (30)$$

Using Figures 25 and 26, pages 77 and 81, with $N = 204$, $V_{\text{max}} = 8.01 \text{ ft/s}$, and other values from Appendices A and B,

$$\Delta P_{\text{total}} = \frac{1}{2(32.2)} (6141.38 + 0.68 + 49.34) \left(\frac{1}{144}\right) \text{ psi}$$

$$\Delta P_{\text{total}} = (18.33 + 0.0020 + 0.147) \text{ in. of water}$$

$$= 18.45 \text{ ins. of water}$$

Using Figure 27 and 28 with $N = 144$, $V_{\text{max}} = 8.01 \text{ ft/sec.}$

$$\Delta P_{\text{total}} = \frac{1}{2(32.2)} (6141.38 + 0.68 + 26.12) \left(\frac{1}{144}\right) \text{ psi}$$

$$\begin{aligned}\Delta P_{\text{total}} &= (18.33 + 0.0020 + 0.078) \text{ in. of water} \\ &= 18.41 \text{ in. of water}\end{aligned}$$

Using Figures 29, page 88, and 30, page 90, with $N=144$, $V_{\text{max}} = 5.33 \text{ ft/s}$

$$\Delta P_{\text{total}} = \frac{1}{2(32.2)} (6141.83 + 0.37 + 13.65) \left(\frac{1}{144}\right) \text{ psi}$$

$$\begin{aligned}\Delta P_{\text{total}} &= (18.33 + 0.0011 + 0.041) \text{ ins. of water} \\ &= 18.37 \text{ in. of water}\end{aligned}$$

Calculation II (Using equation (43) for ΔP_{tube})

$$\begin{aligned}\Delta P_{\text{total}} &= \frac{1}{A_B} \left(\frac{M_s g (\rho_s - \rho_f)}{\rho_s} + 0.664 V_{\text{max}} \frac{\mu_{F.B}}{\mu_F} \sqrt{\frac{V_{\text{max}} \rho_f \mu_f}{H}} (2(B+L)H) \right. \\ &\quad \left. + \frac{C_{D_{F.B}}}{C_{D_{\text{gas}}}} \Delta P_{\text{gas}} \right) \quad \text{C-1}\end{aligned}$$

Using Figures 25 and 26, pages 77 and 81, with $N=204$, $V_{\text{max}} = 8.01 \text{ ft/s}$, and other values from Appendices A and B,

$$\begin{aligned}\Delta P_{\text{total}} &= (18.33 + 0.0020 + 0.072) \text{ ins of water} \\ &= 18.40 \text{ ins of water (In-line)}\end{aligned}$$

$$\begin{aligned}\Delta P_{\text{total}} &= (18.33 + 0.0020 + 0.132) \text{ ins of water} = 18.46 \text{ ins of} \\ &\quad \text{water (staggered)}\end{aligned}$$

Using Figures 27 and 28, pages 84 and 86, with $V_{\text{max}} = 8.01 \text{ ft/s}$ and other values from Appendices A and B,

$$\begin{aligned}\Delta P_{\text{total}} &= (18.33 + 0.0020 + 0.072) \text{ ins of water (In-line)} \\ &= 18.40 \text{ in. of water (In line)}\end{aligned}$$

$$\begin{aligned}\Delta P_{\text{total}} &= (18.33 + 0.0020 + 0.066) \text{ in. of water} \\ &= 18.40 \text{ in. of water (staggered)}\end{aligned}$$

Using Figures 29 and 30, pages 88 and 90, with $V_{\max} = 5.33 \text{ ft/s}$ and other values from Appendices A and B,

$$\begin{aligned}\Delta P_{\text{total}} &= (18.33 + 0.0011 + 0.021) \text{ in. of water} \\ &= 18.35 \text{ in. of water (In line)}\end{aligned}$$

$$\begin{aligned}\Delta P_{\text{total}} &= (18.33 + 0.0011 + 0.039) = 18.37 \text{ in. of water} \\ &\text{(Staggered)}\end{aligned}$$

Calculation of Pressure Drops Using Larger Heat Transfer Tube Diameter

The following calculations are made for 1.5 in. (0.125 ft) diameter tube.

Pressure loss due to In-line tubes distribution using 1 in. centers on Rows and 2 in. centers on columns.

The tubes distribution are as shown in Figure 31,

$$\begin{aligned}V_{\max} &= \frac{V_{\infty} S_n}{S_n - d} = \frac{(4 \text{ ft/s})(2/12 \text{ ft})}{(2/12 - 0.125)} = 7.95 \text{ ft/s} \\ R_{e_{\max}} &= \frac{V_{\max} d}{\nu_f} = \frac{7.95 \text{ ft/s} (0.125 \text{ ft})}{0.165 \times 10^{-3} \text{ ft}^2/\text{s}} = 6.02 \times 10^3 \\ f^1 &= \left(0.044 + \frac{0.08 S_p/d}{((S_n - d)/d)^{0.43 + 1.13d/S_p}} \right) R_{e_{\max}}^{-0.15} \\ &= \left(0.044 + \frac{0.08 (2/1.5)}{((2-1.5)/1.5)^{0.42 + 1.13(1.5)/2}} \right) (6.02 \times 10^3)^{-0.15} \\ &= 0.13 \\ G_{\max} &= \rho V_{\max} = 0.0754 \text{ lbm/ft}^3 (7.95 \text{ ft/s}) \left(\frac{3600 \text{ sec}}{1 \text{ hr.}} \right) \\ &= 2158 \text{ lbm/hr.ft}^2\end{aligned}$$

The contribution of bed weight per unit area, wall and heat transfer tubes effect to the total pressure is shown on Table III.

Configuration 7

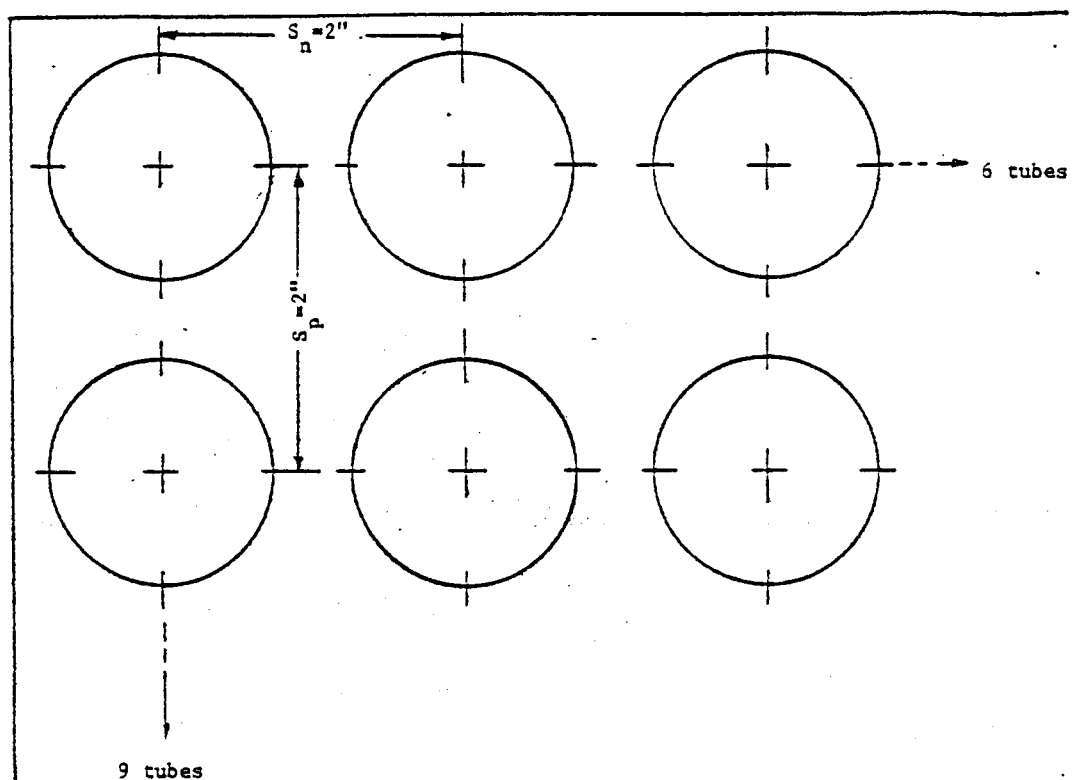


Figure 31. In-line distribution 2 in. centers on both sides.

Table III

Contribution of Bed Weight Per Unit Area, Wall and
Heat Transfer Tubes Effect to the Total Pressure

Con- figu- rations	N	Tube arrange- ments	$V_{\max}$ ft/s	I $\Delta P_{w/a}$ (in. of water)	ΔP_{wall} (in. of water)	ΔP_{tube} (in. of water)	ΔP_{total}
1	204	In-line	8.01	18.33	0.0020	0.147	18.45
2	204	Staggered	8.01	18.33	0.0020	0.147	18.45
3	108	In-line	8.01	18.33	0.0020	0.078	18.41
4	108	Staggered	8.01	18.33	0.0020	0.078	18.41
5	102	In-line	5.33	18.33	0.0011	0.041	18.37
6	102	Staggered	5.33	18.33	0.0011	0.041	18.37
II							
1	204	In-line	8.01	18.33	0.0020	0.072	18.40
2	204	Staggered	8.01	18.33	0.0020	0.132	18.46
3	108	In-line	8.01	18.33	0.0020	0.072	18.40
4	108	Staggered	8.01	18.33	0.0020	0.066	18.40
5	102	In-line	5.33	18.33	0.0011	0.021	18.35
6	102	Staggered	5.33	18.33	0.0011	0.039	18.37

$$\begin{aligned}
 \Delta P &= \frac{f_f^1 G_{\max}^2 N}{P_f (2.09 \times 10^8)} \left(\frac{\mu_w}{\mu_b} \right)^{0.14} \\
 &= \frac{0.13 (2158)^2}{(0.0754)(2.09 \times 10^8)} \frac{(9)}{(9)} \\
 &= 0.35 \text{ lbf/ft}^2 \\
 &= 0.0024 \text{ lbf/in}^2
 \end{aligned}$$

$$\Delta P = 0.07 \text{ in. of water}$$

that is,

$$\Delta P_{\text{air}} = 0.07 \text{ in. of water}$$

Using equation (43), the pressure in the fluidized bed due to heat transfer tubes is

$$\Delta P_{\text{F.B(tubes)}} = \frac{C_{D_{\text{F.B}}}}{C_{D_{\text{gas}}}} P_{\text{gas}}$$

$$R_{e_{\text{F.B}}} = \frac{V_{\max} D_T \rho_f}{\mu_{\text{F.B}}} = \frac{(7.95)(0.125)(0.0754)}{5.61 \times 10^{-5}} = 1335.6$$

$$\therefore C_{D_{\text{F.B}}} \approx 1.0 \text{ (from Figure 7, page 21),}$$

$$R_{e_{\text{air}}} = \frac{7.95 (0.125)(0.0754)}{1.2321 \times 10^{-5}} = 6081.4$$

$$\therefore C_{D_{\text{air}}} \approx 1.0$$

Hence,

$$\begin{aligned}
 \Delta P_{\text{F.B (tubes)}} &= (1)(0.07) \text{ in. of water} \\
 &= 0.07 \text{ in. of water}
 \end{aligned}$$

Using equation 42,

$$\begin{aligned}
 \Delta P_{\text{tubes}} &= (D_T L_T C_{D_{\text{F.G}}} \frac{V_{\max}^2 \rho_f N}{2}) \frac{1}{A_B}, \text{ where } N = 9 \times 6 = 54 \\
 &= ((0.125)(2)(1.0) \frac{(7.95)^2}{2}) (0.0754)(54) \frac{1}{2} \left(\frac{1}{32.2} \right) \left(\frac{1}{144} \right) \text{ psi}
 \end{aligned}$$

$$\Delta P = 0.096 \text{ in. of water}$$

Pressure Loss Due to Staggered Tube Distribution Using 2 in. Centers
on Rows and 2 in. Centers on Columns

The tube arrangements is as shown in Figure 32.

$$V_{\max} = 7.95 \text{ ft/s}, R_{e_{\max}} = 6.02 \times 10^3, G_{\max} = 2157.95 \text{ lbm/hr.ft}^2$$

$$f^1 = (0.25 + \frac{0.118}{((2-1.5)/1.5)^{1.08}}) (6.02 \times 10^3)^{-0.16}$$

$$= 0.158$$

$$\Delta P = \frac{f^1 G_{\max}^2 N}{P (2.09 \times 10^8)} \left(\frac{\mu_w}{\mu_b} \right)$$

$$= \frac{0.158 (2158)^2 (9)}{0.0754 (2.09 \times 10^8)}$$

$$= 0.42 \text{ lbf/ft}^2$$

$$= 0.0029 \text{ lbf/in}^2$$

i.e. $\Delta P = 0.081 \text{ in. of water}$

Using equation (43),

$$\Delta P_{F.B(\text{tubes})} = \frac{C_{D_{F.B}}}{C_{D_{\text{gas}}}} \Delta P_{\text{gas}} = 0.081 \text{ in. of water}$$

Using equation (42)

$$\Delta P_{\text{tubes}} = 0.091 \text{ in. of water}, N = 51$$

Pressure Loss Due to In-Line Tubes Distribution Using 3 ins. Centers
On Both Sides

The tube distributions are as shown in Figure 33.

$$V_{\max} = 7.95 \text{ ft/s}, R_{e_{\max}} = 6.02 \times 10^3, G_{\max} = 2158 \text{ lbm/hr.ft}^2$$

Configuration 8

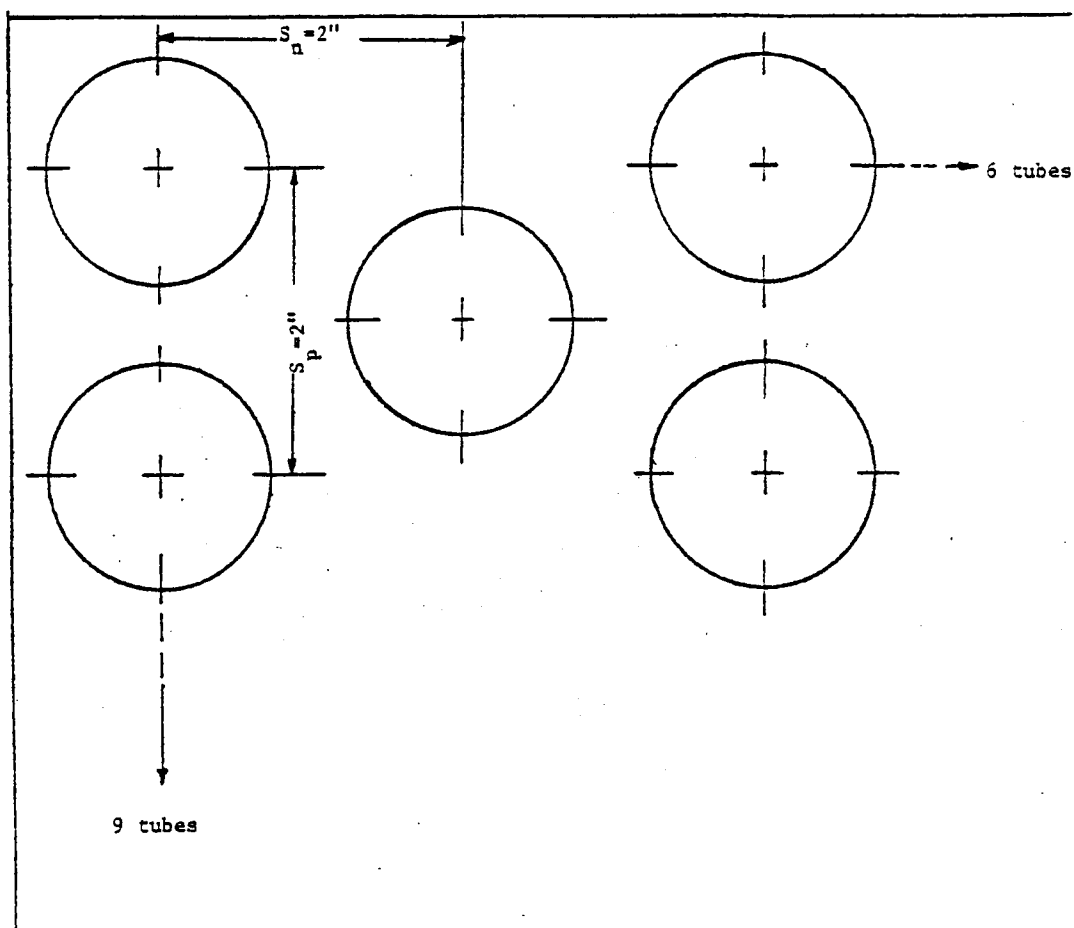


Figure 32. Staggered distribution - 2 in. centers on both sides.

Configuration 9

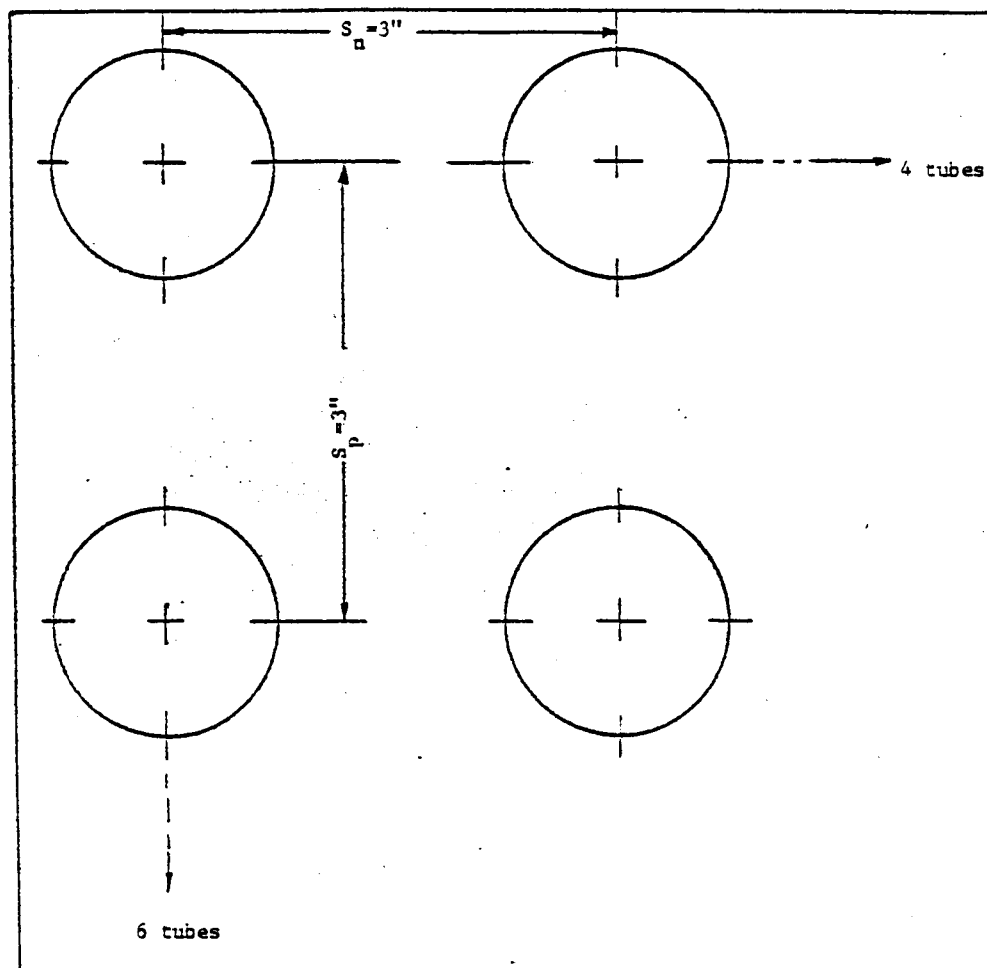


Figure 33. In-line distribution - 3 in. centers on both sides.

$$\begin{aligned}
 f^1 &= \left(0.044 + \frac{0.08 \frac{Sp}{d}}{\left(\frac{(S_n - d)}{d} \right)^{0.43 + 1.13d/Sp}} \right) R_{e_{\max}}^{-0.15} \\
 &= \left(0.044 + \frac{0.08 \left(\frac{3}{1.5} \right)}{\left(\frac{(3 - 1.5)}{1.5} \right)^{0.43 + 1.13(1.5)/3}} \right) (6.02 \times 10^3)^{-0.15} \\
 &= 0.055
 \end{aligned}$$

$$\begin{aligned}
 \Delta P &= \frac{f^1 G_{\max}^2 N}{\rho_f (2.09 \times 10^8)} = \frac{0.055 (2158)^2 6}{(0.0754) (2.09 \times 10^8)} \\
 &= 0.098 \text{ lbf/ft}^2 = 0.019 \text{ in. of water}
 \end{aligned}$$

that is, $\Delta P_{\text{air}} = 0.019 \text{ in. of water}$

Using equation (43) $\Delta P_{\text{F.B(tubes)}} = 0.019 \text{ in. of water}$

Using equation (42), $\Delta P_{\text{tubes}} = 0.043 \text{ in. of water}$, $N = 24$.

Pressure Loss Due to Staggered Tubes Distribution Using 3 in. Centers on Both Sides

The tube distributions are as shown in Figure 34.

$$V_{\max} = 7.95 \text{ ft/s}, R_{e_{\max}} = 6.02 \times 10^3, G_{\max} = 2157.95 \text{ lbm/hr.ft}^2$$

$$f^1 = \left(0.25 + \frac{0.118}{\left(\frac{(S_n - d)}{d} \right)^{1.08}} \right) R_{e_{\max}}^{-0.16} = 0.091$$

$$\Delta P = \frac{f^1 G_{\max}^2 N}{\rho_f (2.09 \times 10^8)} \left(\frac{\mu_w}{\mu_b} \right) = 0.161 \text{ lbf/ft}^2$$

$$\Delta P = 0.031 \text{ in. of water}$$

Using equation (43),

$$\Delta P_{\text{F.B tubes}} = 0.031 \text{ in. of water}$$

Using equation (42), $\Delta P_{\text{tubes}} = 0.039 \text{ in. of water}$, $N = 22$.

For comparison of equations (42) and (43), see Table IV.

Configuration 10

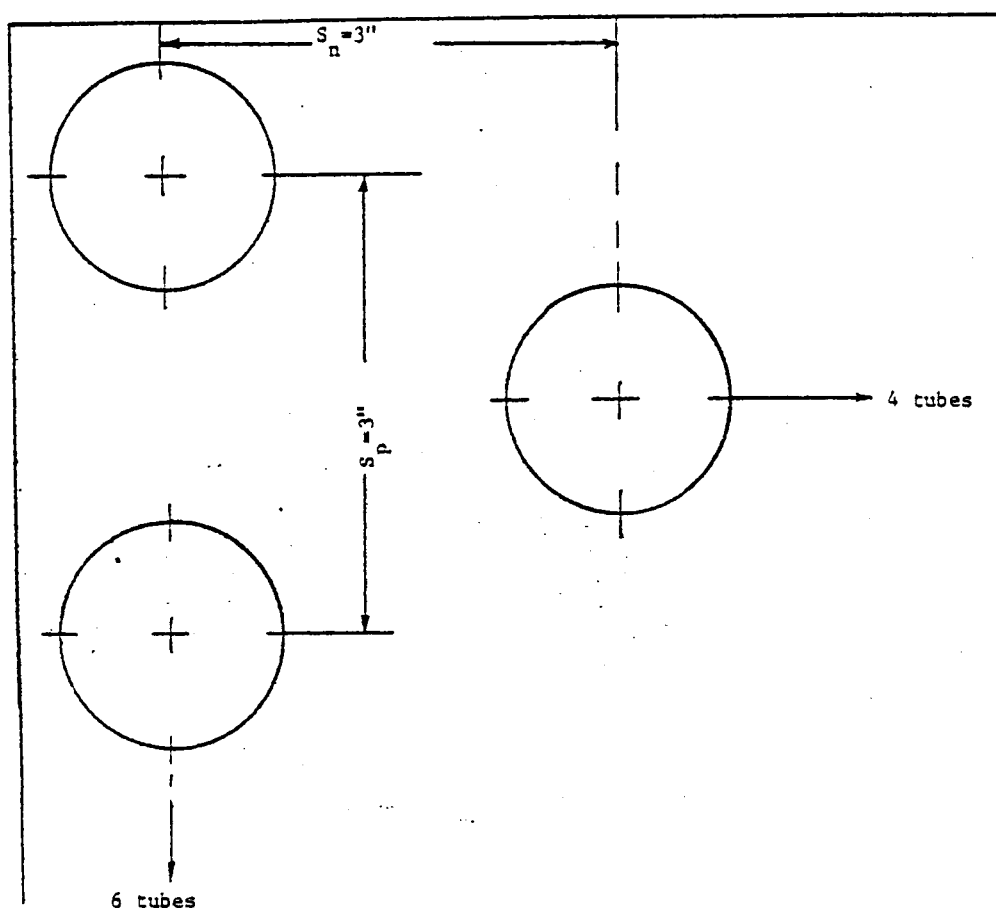


Figure 34. Staggered distribution - 3 in. on both sides.

Table IV

Comparison of ΔP_{tubes} from Equation (42) and (43) for
Different Tube Arrangements

Configurations	N	Tube Arrangements	ΔP_{tubes} (in. of water) Equation (42)	$\Delta P_{\text{F.B. (tubes)}}$ (in. of water) Equation (43)
7	54	In-line	0.096	0.07
8	51	Staggered	0.091	0.081
9	24	In-line	0.043	0.019
10	22	Staggered	0.039	0.031

Total Pressure Drop From Equation (30)

I.

Using equation (42) for ΔP_{tube} ,

$$\Delta P_{\text{total}} = \frac{1}{A_B} \left(M_s g \frac{(\rho_s - \rho_f)}{\rho_s} + 0.664 V_{\text{max}} \frac{\mu_{F.B}}{\mu_f} \sqrt{\frac{V_{\text{max}} \rho_f \mu_f}{H}} (2(B+L)H) \right. \\ \left. + D_{T.T} L_{T.T} C_{D_{F.B}} \frac{V_{\text{max}}^2}{2} \rho_f N \right)$$

Using configuration 7 (N=54) and 8 (N=51) and other values from Appendix A,

$$\Delta P_{\text{total}} = \frac{1}{2(32.2)} (6141.38 + 0.67 + 32.2) \left(\frac{1}{144} \right) \text{ psi}$$

$$\Delta P = (18.33 + 0.0020 + 0.096) \text{ ins. of water}$$

$$= 18.43 \text{ in. of water (In-line)}$$

$$\Delta P_{\text{total}} = 18.42 \text{ in. of water (staggered)}$$

Using configurations 9 (N=24) and 10 (N=22)

$$\Delta P_{\text{total}} = (18.33 + 0.0020 + 0.039) \text{ in. of water} = 18.37 \text{ in. of water (In-line)}$$

$$\Delta P_{\text{total}} = 18.38 \text{ in. of water (staggered)}$$

II

Using equation (43) for ΔP_{tubes} ,

$$\Delta P_{\text{total}} = \frac{1}{A_B} \left(\frac{m_s g (\rho_s - \rho_f)}{\rho_s} + 0.664 V_{\text{max}} \frac{\mu_{F.B}}{\mu_f} \sqrt{\frac{V_{\text{max}} \rho_f \mu_f}{H}} (2(B+L)H) \right. \\ \left. + \frac{C_{D_{F.B}}}{C_{D_{\text{gas}}}} \Delta P_{\text{gas}} \right)$$

Using configurations 7 (N=54) and 8 (N=51)

$$\begin{aligned}\Delta P_{\text{total}} &= (18.33 + 0.0020 + 0.07) \text{ in. of water} \\ &= 18.40 \text{ in. of water (In-line)}\end{aligned}$$

For staggered arrangements,

$$\begin{aligned}\Delta P_{\text{total}} &= (18.33 + 0.0020 + 0.081) \text{ in. of water} \\ &= 18.41 \text{ in. of water (Staggered)}\end{aligned}$$

Using configurations 9 (N=24) and 10 (N=22)

$$\begin{aligned}\Delta P_{\text{total}} &= (18.33 + 0.0020 + 0.019) \text{ in. of water} \\ &= 18.35 \text{ in. of water (In-line)}\end{aligned}$$

Likewise, for staggered arrangements,

$$\begin{aligned}\Delta P_{\text{total}} &= (18.33 + 0.0020 + 0.031) \text{ in. of water} \\ &= 18.36 \text{ in. of water (staggered)}\end{aligned}$$

The contribution of bed weight per unit area, wall and heat transfer tubes effect (using $1\frac{1}{2}$ in. tubes) to the total pressure is shown on Table V.

Table V

Contribution of Bed Weight Per Unit Area, Wall and Heat
Transfer Tubes Effect to the Total Pressure

Con- figur- ations	N	Tube Arrange- ments	$V_{\max}$ ft/s	$\Delta P_{w/a}$ (in. of water)	ΔP_{wall} (in. of water)	ΔP_{tube} (in. of water)	ΔP_{total}
I							
7	54	In-line	7.95	18.33	0.0020	0.096	18.43
8	51	Staggered	7.95	18.33	0.0020	0.091	18.42
9	24	In-line	7.95	18.33	0.0020	0.043	18.38
10	22	Staggered	7.95	18.33	0.0020	0.043	18.38
II							
7	54	In-line	7.95	18.33	0.0020	0.07	18.40
8	51	Staggered	7.95	18.33	0.0020	0.081	18.41
9	24	In-line	7.95	18.33	0.0020	0.019	18.35
10	22	Staggered	7.95	18.33	0.0020	0.031	18.36

VITA

Ganiyu Bolanle Awelenje was born in Lagos, Nigeria on July 24, 1947. He attended elementary school in this city and was graduated from Zumratul Islamiyyah Grammar School in December, 1967. He worked from August 1968 to December 1972 at the Nigerian Law School and National Bank of Nigeria, Lagos, Nigeria. In January 1969, he entered Federal School of Science, Lagos as a part-time student in Pure Mathematics, Applied Mathematics and Physics and graduated in June, 1971.

In spring of 1973, he entered Tennessee Technological University, Cookeville, Tennessee and received his Bachelor of Science degree with distinction in Mechanical Engineering in June, 1976. The following September, he entered the Graduate School of the University of Tennessee, Knoxville and received a Master of Science degree in Mechanical Engineering in June, 1978.

The author is a member of American Society of Mechanical Engineers and American Society of Heating, Refrigeration and Air Conditioning Engineers.

He is married to the former Faozat Sulola Mumuni of Lagos, Nigeria and has a 2½ year old son, Azeez Olusola Awelenje.