HYDRODYNAMICS OF SPOUT FLUID BEDS

By

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B.Eng. McGill University, 1981

**A THESIS SUBMITTED IN PARTIAL FULFILLMENT OF
THE REQUIREMENTS FOR THE DEGREE OF**

MASTER OF APPLIED SCIENCE

in

THE FACULTY OF GRADUATE STUDIES

(Department of Chemical Engineering)

**We accept this thesis as conforming
to the required standard**

July 1983

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ABSTRACT

Experiments were conducted in a spout-fluid bed, a modified version of a standard spouted bed, incorporating auxiliary flow in addition to central flow. The auxiliary flow was supplied in three different ways through the perforated conical distributor base. The vessel used was a cylindrical half column, 0.15 m in diameter and 1.05 m high fitted with an inlet orifice plate with a diameter of either 19.1 mm or 25.4 mm. The three solid materials studied were: polystyrene, millet and high density polyethylene (HDPE).

Aspects studied included regime maps, minimum fluid flowrate for spouting with aeration and spout-fluidization, annulus gas velocity profile, total bed pressure drop, annulus solids circulation, fountain shape and height and spout shape and diameter.

For different central flow/auxiliary flow combination, four types of flow behaviour were established: packed bed; spouting with aeration; spout-fluidization and jet in a fluidized bed.

The minimum total fluid flowrate for spouting with aeration and spout-fluidization was always found to be greater than the minimum spouting velocity. Greater fluid percolation through the annulus in the cylindrical part of the column was achieved by increasing the amount of auxiliary flow for a constant total flow. As much as a 50% increase could be realized for a 1:1 split between central/auxiliary flow compared with the case where there is no auxiliary flow.

The circulation rate of solid particles was affected little by small rates of aeration. At higher rates of aeration and for deeper bed

heights aeration led to an increase of solids circulation. However, at lower bed heights, the solid circulation rate decreased as a result of aeration.

The overall bed pressure drop for spouting with aeration under minimum conditions increases linearly with auxiliary flow.

Fountain height was found to decrease as the fraction of auxiliary flow increased. A greater total gas flow is required to reach a given fountain height for spouting with aeration. However, with aeration, the fountain was observed to have a greater solids concentration than that for pure spouting. The empirical correlation due to McNab was found to predict the average spout diameter very well under various central and auxiliary flow combinations if the sum of central and auxiliary flows is used in the correlation.

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ACKNOWLEDGMENTS

I would like to express my appreciation to Dr. N. Epstein and Dr. J.R. Grace for their thoughtfulness and advice, under whose guidance and encouragement this work was carried out. I am grateful to Dr. C.J. Lim for his helpful discussions and suggestions.

Finally, special thanks are also due to the gentlemen of the Department of Chemical Engineering workshop and stores for their enthusiastic and invaluable assistance.

CHAPTER 1

INTRODUCTION

Fluidization has been widely accepted as a solid-fluid contacting technique. Relatively fine ($d_p \leq 1 \text{ mm}$) particles are desirable for good quality fluidization operations. The application of the spouted bed technique as an alternative to fluidization for handling coarse particles ($d_p \geq 1 \text{ mm}$) has achieved increased recognition in recent years [K.B. Mathur and N. Epstein (1974)].

Fig. 1.1 illustrates schematically a typical spouted bed column with a conical base. Under conditions of spouting, the bed consists of two distinct zones: the very dilute phase central core called the spout, and the surrounding dense phase annular solids called the annulus. The upward moving particles in the spout, having reached a certain height above the bed surface, fall back onto the annulus where they slowly move downwards and, to some extent, radially inwards as a loosely packed bed. Fluid from the spout percolates through these annular solids as it travels upwards.

Various modifications to standard spouted beds have resulted in a wide range of useful applications [H. Kono (1981), K.B. Mathur and N. Epstein (1974)]. One such modification introduces fluid to the annulus directly from an external source in order to increase the fluid percolation through the annulus, especially in the lower section. In addition to supplying spouting fluid through a centrally located opening, auxiliary fluid is supplied through a flat or conical porous/perforated distributor as shown in Figures 1.2a and 1.2b. This

Fig. 1.1. Schematic diagram of a spouted bed

Fig. 1.2a Schematic diagram of a spout-fluid bed with conical base distributor.

Fig. 1.2b Schematic diagram of a spout-fluid bed with flat base distributor.

particular modified spouted bed, usually known as a "spout-fluid bed", is obtained by aerating or fluidizing the annular solids. Spout-fluid beds show better solids mixing and annular solid-fluid contact than standard spouted beds. In addition, they are better for sticky or agglomerating solid particles.

A large amount of information exists on the hydrodynamics of spouted beds and fluidized beds. Relatively little information is, however, available for either gas or liquid spout-fluid beds. Quite often, the fragmentary information obtained by one worker differs considerably from that of another. The disagreement can partly be attributed to differences in column and inlet geometry, particle and fluid properties and the operating conditions used.

The main objective of the present research is to study the changes in the hydrodynamic behaviour of a standard spouted bed as modified by the addition of auxiliary flow. Hydrodynamic characteristics studied are: the total minimum fluid flowrate and bed pressure drop for spouting with aeration; regime maps; longitudinal fluid velocity distribution in the annulus; particle circulation in the annulus; height and shape of the fountain; and the shape and diameter of the spout.

A secondary objective is to compare the data obtained with existing correlations and data where appropriate, and, wherever possible, to close the gaps in the existing fragmentary information for spout-fluid beds.

CHAPTER 2

LITERATURE REVIEW

The spout-fluid bed technique has elicited great interest in recent years as a modification to the simple standard spouted bed. Because the basic hydrodynamic features of the spouting remain unaltered, the present review is confined to those aspects of spouted bed behaviour which have direct relevance to the present research and also to the very limited information available on spout-fluid beds.

2.1 Regime Maps:

The presence of auxiliary flow yields a number of different types of bed behaviour which are generally represented by a "phase diagram" or "regime map".

Regime maps determined experimentally by Chatterjee (1974), Littman et al (1982) and Dumistrescu (1977) are shown in Figures 2.1a, 2.1b and 2.1c, respectively. These three regime maps, though they appear to be different from each other, share many common features:

- i. packed bed region,
- ii. fluidization,
- iii. spouting with aeration of the annulus, and
- iv. spouting with fluidization of the annulus, or
spout-fluidization.

A hypothetical regime map shown in Figure 2.2 could be drawn encompassing all the common flow behaviours.

Fig. 2.1a Regime map due to Chatterjee (1974), [material: glass beads, $\rho_s = 2.41 \text{ g/cm}^3$, $H = 18.5 \text{ mm}$, $d_p = 0.60 \text{ mm}$, $d_i = 5.0 \text{ mm}$, $D_c = 90 \text{ mm}$]
Fluid used: air

Fig. 2.1b Regime map due to Littman (1982)
[material: calcium carbonate particles, $\rho_s = 2.6 \text{ g/cm}^3$, $d_p = 1.8 \text{ mm}$, $D_c = 70 \text{ mm}$, $d_i = 15 \text{ mm}$, $H = 60 \text{ mm}$]
Fluid used: air

Fig. 2.1c Regime map due to Dumistrescu (1977)
 [material: polyethylene pellet, particle size: 3mm diameter, 1.5-4.0 mm long, $D = 15\text{cm}$, $d_i = 2.21\text{ cm}$, $H = 25^{\text{C}}\text{cm}$]
 Fluid used: air

Fig. 2.2 Hypothetical phase diagram

2.1.1

- (i) Packed bed condition: $Q_T < Q_{mf}$ and $Q_T < Q_{ms}$

This regime exists as long as the sum of the central flow and auxiliary flow is less than the minimum fluidization and the minimum spouting flow.

- (ii) Fluidization: $Q_T \geq Q_{mf}$

This regime requires a much higher auxiliary flow than central spouting flow. Varying the spouting flow and auxiliary flow can alter the type of fluidization, ranging from a bubbling to a slugging behaviour. In the bubbling mode, the bubbles are formed above the inlet orifice plate and rise along the spout axis. Chatterjee denoted this condition as either the unstable spout-fluid bed regime or the unstable spouting regime, while Dumistrescu called it either fluidization or bubbling.

However, if bubbles grow until their volume equivalent diameter equals or exceeds the column diameter, slugging of the upper bed material is observed. This regime is denoted by Littman as "jet in fluidized bed".

- (iii) Spouting with aeration: $U_{AH} < U_{mf}$

Chatterjee and Dumistrescu denoted this regime as spouting while Littman named it spout-fluidization with $H < H_{msf}$. Spouting with aeration, as far as its visible characteristics are concerned, is similar to spouting in a standard spouted bed. The sum of auxiliary and spouting flows is always greater than the minimum spouting flowrate, except in the result obtained by Chatterjee which showed they are the

same. In all three studies, there appear to exist limiting values for central flow, $(Q_{msa})_{min}$, and auxiliary flow, $(q_{msa})_{max}$, below and above which respectively spouting with aeration will not occur (The limiting value is denoted as point C in Figures 2.1a, 2.1b and 2.1c).

(iv) Spout-fluidization: $U_{AH} \geq U_{mf}$

Spout-fluidization is achieved if sufficient auxiliary flow is supplied to cause the annular solids at the bed surface surrounding the spout of a spouted bed to be fluidized. The behaviour of the bed at this condition resembles that of a spouted bed at $H = H_m$ where the spout instability just begins to develop. The instability is reflected by the pulsations of the spout boundary near the bed surface due to break-up of the spout into bubbles. Dumistrescu named this regime "pulsed jet".

2.2.1 Minimum Total Fluid Flow for Spouting with Aeration:

Chatterjee (1977) found that the minimum total fluid flow remains constant along the line L-M in Figure 2.1a and almost equal to the minimum spouting flow:

$$\text{i.e.} \quad Q_{T,msa} \approx Q_{ms} \quad (2.1)$$

However, a correlation obtained by statistical analysis of experimental data by Dumistrescu (1977) showed that

$$U_{T,msa} = 27.7 \frac{q_{msa}}{Q_{msa}} + U_{ms} \quad (2.2)$$

where U_{ms} was estimated from the correlation of Mathur and Gishler (1955),

$$U_{ms} = \left[\frac{d_p}{D_c} \right] \left[\frac{d_i}{D_c} \right]^{1/3} \left[\frac{ZgH(\rho_s - \rho)}{\rho} \right]^{1/2} \quad (2.3)$$

Thus the minimum total fluid superficial velocity under aeration is always greater than the minimum spouting velocity.

Based on the work of Vuković et al (1972), Littman et al [(1974), (1976)] showed that the minimum spout-fluid flowrate varies between Q_{ms} and Q_{mf} (for $Q_{mf} \geq Q_{ms}$) and for any given bed height, $H < H_m$,

$$q_{msa} = Q_{mf} - \left(\frac{Q_{mf}}{Q_{ms}} \right) Q_{msa} \quad (2.4)$$

Since

$$Q_{T,msa} = q_{msa} + Q_{msa} \quad (2.5)$$

therefore

$$Q_{T,msa} = Q_{mf} - \left[\frac{Q_{mf}}{Q_{ms}} - 1 \right] \left[Q_{msa} \right] \quad (2.6)$$

Grbavčić et al (1976) found for a spouted bed that

$$\frac{Q_{ms}}{Q_{mf}} = \left[1 - \frac{A_A}{A_C} (1 - h)^3 \right] \quad (2.7)$$

Combining equations (2.4) and (2.7) yields

$$\frac{Q_{msa}}{Q_{mf}} = \left[1 - \frac{A_A}{A_C} (1 - h)^3 \right] \left[1 - \frac{q_{msa}}{Q_{mf}} \right] \quad (2.8)$$

However, for a deeper bed such that $Q_{ms} > Q_{mf}$, equations (2.4) to (2.8) are no longer valid.

Further revision by Littman (1983) led to:

$$\begin{aligned} Q_{T,msa} &= q_{msa} + Q_{msa} \\ &= (U_{AH} A_A)_{msa} + (U_{SH} A_S)_{msa} \end{aligned} \quad (2.9)$$

The annular velocity in a spout-fluid bed was derived separately by Littman (1976) as:

$$\frac{U_A(Z)}{U_{mf}} = 1 - \left(1 - \frac{q}{Q_{mf}}\right) \left(1 - h_{msf} \frac{Z}{H}\right)^3 \quad (2.10)$$

It was assumed that for spouting with aeration at the minimum condition,

$$U_{SH} = U_{mf}, \quad q = q_{msa} \quad (2.11)$$

$$Q_T = Q_{T,msa} \text{ with } H \leq H_{msf} \quad (2.12)$$

Substituting Equations (2.10) and (2.11) into Equation (2.9) and

assuming $A_A/A_C = 1$ yields

$$\frac{Q_{T,msa}}{Q_{mf}} = 1 - \left(1 - \frac{q_{msa}}{Q_{mf}}\right) \left(1 - h_{msf}\right) \quad (2.13)$$

H_{msf} is difficult to measure. Furthermore, it is also a function of q_{msa} , a relationship not very well understood. The validity of Equation (2.13) is therefore subject to uncertainties.

2.2.2 Minimum Total Fluid Flow for Spout-fluidization

The only correlation available for estimating the total flow required for spout-fluidization was proposed by Chatterjee (1974). It combines the fluid needed to spout a static bed at the minimum condition with the fluid needed for minimum fluidization of the annulus of that spouted bed:

$$U_{T,msf} = U_{ms} + U_{mf} (1 - \phi) \quad (2.14)$$

The function ϕ was determined empirically to be

$$\phi = 0.20 d_p^{-0.320} d_i^{0.235} H^{0.160} \quad (2.15)$$

The value of U_{ms} was obtained from the Mathur and Gishler correlation, Eq. (2.3).

2.3 Longitudinal Annular Fluid Velocity and Overall Bed Pressure Drop

The analysis of Mamuro and Hattori (1968, 1970) leads to a successful expression for the longitudinal fluid velocity profile and pressure drop in a spouted bed at its maximum spoutable bed depth:

$$\frac{U_A(z)}{U_{mf}} = 1 - (1 - z/H_m)^3 \quad (2.16)$$

$$\left. \frac{\Delta P_s}{\Delta P_f} \right|_{H_m} = 0.75 \quad (2.17)$$

For a given column geometry, solid material and spouting fluid, Grbavcic pressure gradient at any given level in the annulus is independent of bed height. Hence

$$\frac{U_{AH}}{U_{mf}} = 1 - (1 - H/H_m)^3 \quad (2.18)$$

Dividing Eq. (2.16) by Eq. (2.18) yields the annular fluid velocity at any given height, $Z \leq H \leq H_m$:

$$\frac{U_A(Z)}{U_{AH}} = \frac{1 - (1 - Z/H_m)^3}{1 - (1 - H/H_m)^3} \quad (2.19)$$

While the overall pressure drop becomes:

$$\frac{\Delta P_s}{\Delta P_f} = 1.5 h - h^2 + 0.25 h^3 \quad (2.20)$$

Equations (2.16) and (2.17) were derived based on the following boundary conditions:

- (i) $Z = 0, U_A = 0$
- (ii) $Z = H_m, U_A = U_{AHm} = U_{mf}$
- (iii) $Z = H_m, \left(\frac{-dP}{dZ}\right)_{H_m} = \left(\frac{-dP}{dZ}\right)_{mf}$
 $= (\rho_s - \rho) (1 - \epsilon_A)g$

Conditions (ii) and (iii) were discussed in depth by Epstein, Lim and Mathur (1978). For a spout-fluid bed, U_A is no longer zero at $Z=0$. Nonetheless, Equations (2.16) through (2.20) form the basis of understanding of the behaviour of a spout-fluid bed.

In an attempt to account for the auxiliary flow, Littman (1976) modified Equation (2.18) and proposed Equation (2.10).

Assuming Darcy's law in the annulus,

$$\frac{dP_A}{dZ} = k U_A(Z) \quad (2.21)$$

integrating Eq. 2.21 with U_A from Eq. (2.10), and noting that

$\Delta P_{mf} = k U_{mf} H$, he obtained for $h_{msf} \leq 1$:

$$\left. \frac{\Delta P_s}{\Delta P_f} \right|_{msa} = f(h_{msf}) + \frac{q_{msa}}{Q_{mf}} \left[1 - f(h_{msf}) \right] \quad (2.22)$$

where

$$f(h_{msf}) = 1.5(h_{msf}) - (h_{msf})^2 + 0.25 (h_{msf})^3 \quad (2.23)$$

2.4 Annular Solids Circulation Rate:

The only study in the literature on annular solids circulation rate in spout-fluid beds was made by Yang (1983). Because the spout-fluid bed was fitted with a draft tube, the rate of solids circulation depended primarily on the entrainment rate of solids below the draft tube inlet rather than along the surface of the entire spout. Yang found that at low rates of aeration, the solid circulation is essentially similar to that without aeration. However, at higher aeration rates, a substantial increase in solids circulation rate is realized with the same total gas flowrate. He concluded that a minimum aeration rate is required to increase substantially the solids circulation rate.

2.5 Fountain Shape and Height

The shape of the fountain is well defined and is roughly paraboloidal. Grace and Mathur (1978) formulated a theoretical model, based on a force-balance analysis, for the prediction of maximum height of a fountain. They also proposed a simpler approximate equation for estimating fountain height, by ignoring the drag force:

$$H_F = \epsilon_o^{1.46} \left[\frac{v_o^2 \max}{2g} \frac{\rho_s}{(\rho_s - \rho)} \right]$$

2.6 Spout Shape and Diameter

Spout shape and diameter have been determined by a number of workers from measurements using half-cylindrical columns. It was shown that there was no difference between spout shape obtained from measurements in a full column and a half-column [Mikhailik (1966), Lim and Mathur (1974)].

The empirical dimensional correlation of longitudinal average spout diameter, d_s , due to McNab (1972) is perhaps the most popular. This gives

$$d_s = \frac{1.48 G^{0.49} D_c^{0.68}}{\rho_b^{0.41}}$$

CHAPTER 3

APPARATUS AND BED MATERIALS

3.1 Apparatus

Experiments were carried out in a cylindrical half-column, 15.2 cm in diameter by 1.05 m high, equipped with a 60° included angle conical base perforated distributor as shown in Figure 3.1. The column was constructed from plexiglass, except for the front flat surface which was made of glass. Semicircular orifice plates of diameter 1.91 cm or 2.54 cm and a coarse screen could be sandwiched between the conical base and the inlet section. A straight vertical pipe of 3.18 cm I.D. and 30 cm long formed the approach section. A 2.54 cm long 'honey comb' type flow straightening bundle was fixed inside this approach section.

Bed pressure drop measurements were provided by two pressure taps. One was in the approach section, 5 cm upstream from the inlet orifice plate. The other was situated in the orifice plate itself, achieved by having a 0.75 mm hole drilled through each of the orifice plates and a small diameter stainless steel tube inserted as shown in Fig. 3.2. The other end of the tube was connected to a U-tube liquid manometer.

The conical base section is shown in Figures 3.3a and 3.3b. The perforated annulus distributor was constructed from five layers of equal thickness plexiglass plates, machined and glued to form a single piece. The number and size of distributor openings were determined such that the pressure drop across the distributor equalled 25.4 cm H₂O at a superficial gas velocity of 1 m/s. There were 35 openings, each 3 mm in

Fig. 3.1 Schematic diagram of the column

TOP VIEW

SIDE VIEW

Fig. 3.2 Diagrams of inlet orifice plate with pressure tap

Fig. 3.3a Conical base distributor section. Scale 1:2

Fig. 3.3b Sectional side view of the conical distributor. Scale 1:1.5

diameter, evenly spaced throughout the entire conical surface as shown in Fig. 3.4. The distributor assembly was actually composed of five layers of subdistributor stacked together (see Fig. 3.3b) with each subdistributor leading to a separate auxiliary gas supply line. Gas flow through each of these five lines could be controlled independently. A diffuser, was provided at the outlet end of each auxiliary gas supply line to provide uniform gas flow across the distributor openings. These diffusers were all housed inside the calming chambers (see Fig. 3.3b). This distributor design provided great flexibility, enabling the auxiliary flow to be introduced in a variety of different ways. A 75 mesh stainless steel screen was secured over the conical surface on the inside to prevent particles from entering or blocking the auxiliary air distributor openings.

Air flowrates were monitored by six orifice plate flowmeters, one for the central spouting air supply and one for each of the five auxiliary air chambers. Dimensions are given in the Appendix IV. Calibration was carried out with a dry gas meter.*

The upstream air pressure of the orifice plate flowmeter was indicated by a pressure gauge (see Fig. 3.5b). All air flowrates were corrected to 20°C and one atmosphere. A cylindrical buffer tank 25 cm in diameter and 60 cm high was connected to the main air supply line to reduce incoming pressure fluctuations. A schematic flow diagram and the picture of the experimental set-up is shown in Figures 3.5a and Fig. 3.6 respectively.

*Model AL425, by Canadian Meter Company Limited.

Fig. 3.4 Flattened out surface of the conical perforated distributor

Fig. 3.5a Schematic diagram of the experimental set-up

Fig. 3.5b Orifice plate flowmeter assembly

Fig. 3.6 Picture of the experimental set-up.

Annulus gas velocity measurements were provided by the static pressure probe shown in Fig. 3.7, the same probe as used by Lim (1975). The probe consists of two unequal lengths of stainless steel tubing with O.D. of 1.6 mm enclosed in a larger stainless steel tube housing of diameter 4.8 mm to provide rigidity and support. The two ends were fitted with 22 gauge hypodermic needles, bent at right angles 1.2 cm from the tips. The tips were vertically separated by 2 cm. Static pressure drops were measured using an inclined (15° with respect to horizontal) manometer filled with fluid of specific gravity 0.827. Minimum readability was ± 0.5 mm on a linear scale, which corresponds to a pressure drop of ± 0.11 mm H₂O.

3.2 Solid Properties

Three solid materials were used in this work: polystyrene, millet and high density polyethylene (HDPE). The materials were initially screened to a relatively narrow size range before particle density, particle size, bulk density and the loosely packed voidage were measured.

3.2.1 Particle Shape and Size

The HDPE particles were nearly spherical, while the polystyrene and millet particles were assumed to be elliptical cylinders and oblate spheroids, respectively. Equivalent particle diameters were defined as the diameter of a sphere of the same volume as the particle. Two thousand polystyrene, seven hundred and fifty HDPE and three thousand millet particles were separately weighed and their volumes measured by water displacement in a 100 c.c. graduated cylinder. In the case of

Fig. 3.7 Schematic diagram of the static pressure probe

millet, which is permeable to water, the particles were first coated with a thin film of Aeroplast¹ dressing. Particle dimensions for polystyrene and millet were also obtained by measuring the size of 20 particles chosen at random for each material. Results obtained by the displacement method are used in this work.

3.2.2 Particle Density, Bulk Density and Loosely Packed Bed Voidage

Particle density was determined by means of pycnometer. The procedure for bulk density measurement used by Oman and Watson (1944), Eastwood et al (1967) and Lim (1975) was adopted. A 500 c.c. graduated cylinder was partially filled with a known weight of solid particles. The cylinder was covered at its open end and inverted, and then quickly returned to its original upright position. The volume occupied by the solids at the end of this procedure was taken as the bulk volume under loosely packed conditions. Five replicates were taken to calculate the bulk density of the material. The variation between separate tests was less than 1%.

Voidage in the loosely - packed bed condition was calculated from particle and bulk densities as follows:

$$\epsilon_A = 1 - \frac{\rho_b}{\rho_s}$$

The results obtained are presented in Table 1.

¹Brand name for a waterproof dressing made by Parke-Davis.

Table 1: Properties of Solid Particles used

Material	d_p (cm)	ρ_s g/cm ³	ρ_b g/cm ³	ϵ_A
Polystyrene 3.51 mm * 2.51 mm * 1.78 mm	0.285	1.04	0.59	0.43
Millet	0.228	1.14	0.68	0.40
HDPE	0.394	0.93	0.58	0.38

CHAPTER 4

CHOICE OF EXPERIMENTAL CONDITIONS

4.1 Points of Entry for Auxiliary Air Flow

The distributor assembly allowed the auxiliary air to be supplied in thirty-two (2^5) different ways (including the case when there was no auxiliary flow). However, only three configurations were used.

Configuration #1 - Auxiliary air was supplied at all levels (i.e. at each of the five subdistributors shown in Figures 3.3c and 3.4).

Configuration #2 - Auxiliary air was supplied through the upper two chambers only.

Configuration #3 - Auxiliary air was supplied through the two lowest chambers only.

For each configuration, the air flow to each chamber was adjusted in proportion to the number of openings from that calming chamber. In other words, the air velocity through each opening was the same for all chambers that were in operation.

The above configurations represented the normal case (configuration #1) and two extreme cases (configuration #2 & #3).

4.2 Central Flow and Auxiliary Flow Combination

Proper choice for combinations of the central flow and auxiliary flow for spouting with aeration was difficult without knowing how the various flow regimes depended on these flows. Therefore, regime maps were first obtained so that appropriate flow combinations could be selected.

CHAPTER 5

REGIME MAPS AND MINIMUM FLUID FLOWRATE FOR SPOUTING WITH AERATION AND SPOUT FLUIDIZATION

5.1 Experimental Technique

Two methods were used to obtain the flow regimes

Method #1 - The central air flow was increased/decreased continuously while maintaining the auxiliary air flow constant.

Method #2 - The auxiliary air flow was decreased/increased stepwise while maintaining the central air flow constant.

Several initial trial runs were made. It was found that the flow regimes were quite distinct and could be identified visually. Moreover, the transition from one regime to another was accompanied by either a sudden increase/decrease in bed pressure drop or increase/decrease in the magnitude of pressure drop fluctuation. These observations thus formed the basis for distinguishing the different flow regimes.

5.1.1 Transition From Spouting with Aeration to Fluidized Bed or Fixed Packed Bed

For a given bed height,* bed material and inlet orifice diameter, spouting was initially obtained by supplying sufficient central spouting air. The auxiliary flow was then supplied and maintained constant while

*Bed height was measured to the top of the annulus under spouting conditions at a spouting flowrate of 1.1 Q_{ms}.

the central air flow was gradually decreased (method #1) until a point was reached where a sudden increase in bed pressure drop and the collapse of the spout were noted. This point, was taken as the condition for minimum spouting with aeration. If $Q_{ms} > Q_{mf}$, fluidization of the upper bed solid material was obtained after the transition; on the other hand, if $Q_{ms} < Q_{mf}$, direct transition to fixed packed bed conditions was obtained. The auxiliary flow was supplied in the range from no flow to a value where fluidization of the upper bed annular solids was attained. Thus, a set of points representing the relative influences of central and auxiliary flows was established.

5.1.2 Transition from Spouting with Aeration to Spout Fluidization

In this case, method #2 was employed. Spouting was established initially, the central air flow was then kept constant while the auxiliary flow was increased stepwise. Two situations were observed:

- (I) If the auxiliary flowrate was increased stepwise, a point was reached where spout instability was obtained. This marks the onset of spout fluidization as described already in Section 2.1.1 (iv). The transition was accompanied by fluctuation of bed pressure drop. The auxiliary flow for this transition was denoted as q^* .
- (II) If the auxiliary flowrate was increased stepwise and, at each step, the column was manually given a shake or two, the bed would behave in one of two different ways:

- (i) spout instability was attained at an auxiliary flowrate q^{**} which was lower than q^* . The spout would remain unstable.
- (ii) Spout instability was attained at an even lower auxiliary flowrate, q^{***} , but stability was restored immediately.

It was apparent that under condition II(i) in the absence of shaking, the bed was in a state of pseudo-stability provided that $q^{**} < q < q^*$, where

$$q^{***} < q^{**} < q^*$$

Therefore, the transition point was taken as the maximum value of q^{***} or minimum value of q^{**} at which instability could not be sustained.

Several initial spouting flowrates in the range of $1.0 Q_{ms}$ to $1.4 Q_{ms}$ were used to obtain different experimental points. It was found that at an auxiliary flowrate of q^{**} , spout instability could also be attained, even without any physical disturbance, if the bed was kept at this condition for several minutes. The physical disturbance only provided a faster transition to instability.

5.1.3 Transition from Spout-fluidization to Jet in Fluidized Bed

Auxiliary air was introduced to a spouted bed until spout-fluidization was obtained as described in the previous section. Once this state was attained, the central air flow was decreased slowly and the bed was observed to gradually undergo a transition state where the condition of slugging* and spout-fluidization alternated. A point

*The term slugging as used here refers to the condition of a submerged jet in fluidized bed with slugging of the solid material in the upper section of the bed. Please refer to section 5.2 for greater detail.

was reached, upon further reducing the central air flow, where the bed was in a completely slugging mode. The transition was accompanied by an increase in the magnitude of pressure fluctuation due to the rising and collapsing of the bed surface as slugs break through.

5.2 Results and Discussion

5.2.1 Regime Maps:

A typical phase diagram is shown in Fig. 5.1. The ordinate and abscissa denote the auxiliary flowrate and central flowrate normalized with respect to minimum fluidization flowrate, Q_{mf} . The transition points determined, based on the criteria described in Section 5.1, have been used to define the boundaries of the various flow regimes. Four fairly distinct flow regimes have been delineated: Spouting with aeration (SA), Spout-fluidization (SF), Submerged jet in fluidized bed (JF), and Packed bed (P).

The physical appearance of these flow regimes is shown in Fig. 5.2.

For the jet in fluidized bed regime there were two states of fluidization. At higher auxiliary flow, slugging of the solid material in the upper section of the bed was observed, with a slugging frequency of the order of one to two Hz, similar to that in conventional slugging. This state was denoted as JF(I). At lower auxiliary flows a subregime, designated JF(II), was noted where bubbles broke the bed surface. The transition between JF(II) and JF(I) was gradual and could

Fig. 5.1 Regime map [system: polystyrene, $d_i = 1.91$ cm, $H = 60$ cm]

Points designation:

G - Fluidization at minimum condition

A - Spouting at minimum condition

C - Co-ordinates for spout-fluidization at minimum condition,

i.e. $(\frac{q_{msa}}{Q_{mf}})_{max}$ and $(\frac{Q_{msa}}{Q_{mf}})_{min}$ respectively.

Fig. 5.2 Physical appearance of various flow regimes [(P)- fixed packed bed, (SA)- spouting with aeration, JF(I) & JF(II)- submerged jet in fluidized bed, (SF)- spout-fluidization, (T)- transition between JF(I) and (SF)]

not be established from measurement of the changes in pressure drop. No annular solids movement was observed for JF(II). However, for the JF(I) state, solids mixing resulted from the periodic rise and fall of the slugs.

In spout-fluidization, stable spouting was not achieved. The annular solids at the upper section of the bed were being fluidized. The spout diameter near the bed surface appeared to shrink and expand in a periodic manner. However, on closer examination by means of a Strobotac* and Super-8 mm movie film taken at 64 frames per second, the spout was observed to be not continuous; instead it broke into large bubbles near the bed surface. For millet, the formation of the bubbles occurred much earlier, extending from a few centimeters above the inlet orifice plate. The "spout" was in fact a succession of rising bubbles. The frequency of bubble break-up at the bed surface determined by the Strobotac was found to be in the range of 6 to 7 Hz. A similar range was also determined by Heil (1983) and by Rowe (1983).

The periodic formation of bubbles caused the spout to appear to fluctuate in diameter. The appearance of this regime is shown in Fig. 5.3.

The shape and structure of the fountain was entirely different from that observed for stable spouting. The fountain was formed by solids material periodically ejected by the bursting bubbles. A maximum height was achieved at the burst of the rising bubbles and a minimum immediately thereafter. Therefore, the height of the fountain oscillated between a maximum and a minimum value with the same frequency

*Trade name for strobe light generator.

Fig. 5.3 Physical appearance of spout-fluidization, [polystyrene, $H = 30\text{cm}$, $d_i = 1.91\text{ cm}$, $Q_T = 9\text{ l/s}$, $r_1 = 1$]

as the frequency of bubble eruption. (See Fig. 5.3).

The movement of annular solids near the bed surface assumed a slip-stick type behaviour, and the solids movement gradually became more continuous with decreasing height. The behaviour of spout-fluidization was similar to that of a standard spouted bed whose height equals the maximum spoutable bed depth.

The transition from spout-fluidization to submerged jet in fluidized bed of the type JF(I) was gradual. On decreasing the central flow, the frequency of bubble formation decreased. This allowed bubbles to grow larger in size before reaching the bed surface. A further decrease of central flow resulted in bubbles whose size was equal to or greater than the column diameter, and slugging of the upper bed material was thereby obtained. The transition between these two regimes is illustrated for millet and polystyrene in Figure 5.4, and is also shown as transition state (T) in Fig. 5.2. The sequence of transition from spout-fluidization to submerged jet in fluidized bed of the type JF(I) is shown in Fig. 5.5.

Regime maps of other workers discussed in Chapter 2 were determined from experiments in full columns. The visible characteristics of the spout were found to be distinct in each flow regime in the present half-column work. The inability of previous workers to observe changes in spout shape between the different regimes no doubt contributed to difficulties in interpreting flow behavior. Visual observation and pressure drop measurement were of equal importance in the present work. The half-column geometry thus provided important observations which

millet

polystyrene

Fig. 5.4 Transition state between spout-fluidization and submerged jet in fluidized bed of the type JF(I)

Fig. 5.5 Transition from spout-fluidization to submerged jet in fluidized bed of the type JF (I) [polystyrene, $q_a = 4.5$ l/s]

would have been impossible in a full column. Regime maps are shown in Figures 5.1 and Figures 5.6 through 5.13 for the various systems investigated.

The hypothetical phase diagram of Fig. 2.2 predicted the existence of these flow regimes fairly well. The three materials studied yielded similar patterns (Figures 5.1 and Figures 5.6 through 5.13). The line A-C has a negative slope greater than one, indicating that the sum of auxiliary and central flow necessary for spouting with aeration is greater than that required for spouting in a standard spouted bed. This indicates a modified fluid flow pattern in the bed, with a greater portion of the total fluid passing through the annular solids as the proportion of auxiliary flow is increased.

In all cases line B-E has a positive slope which indicates that penetration of the spout through the bed surface in a fluidized state requires greater central spouting flow; otherwise, slugging is achieved. Polystyrene particles, which have an elliptical cylindrical shape, show a greater tendency to interlock and bridge. They are thus more likely to show slugging behaviour. This is reflected in the smaller slope of line B-E compared with those for millet and HDPE.

The transition points separating spout-fluidization and spouting with aeration were obtained at the highest possible auxiliary flow that does not cause fluidization of the annular solids at the bed surface. For polystyrene particles, the boundary shown by line C-D has an initial slope of zero and the slope becomes greater than zero with increasing central flow. Slopes of the line C-D for both HDPE and millet particles are always greater than zero. A similar trend was noted by Littman (1982), Chatterjee (1974) and Dumistrescu (1977). This may occur

Fig. 5.6 Regime map [system: polystyrene, $d_i = 1.91$ cm, $H = 30$ cm]

Fig. 5.7 Regime map [system: polystyrene, $d_i = 1.91$ cm, $H = 45$ cm]

Fig. 5.8 Regime map [system: HDPE, $d_1 = 1.91$ cm, $H = 30$ cm]

Fig. 5.9 Regime map [system: HDPE, $d_1 = 1.91$ cm, $H = 35$ cm]

Fig. 5.10 Regime map [system: millet, $d_i = 1.91$ cm, $H = 30$ cm]

Fig. 5.11 Regime map [system: millet, $d_i = 1.91$ cm, $H = 45$ cm]

Fig. 5.12 Regime map [system: polystyrene, $H = 45$ cm, $d_i = 2.54$ cm]

Fig. 5.13 Regime map [system: millet, $H = 30$ cm, $d_i = 2.54$ cm]

because there is a decrease of annulus gas velocity for an increasing central flow, (Lim (1975)), so that more auxiliary flow can be introduced before spout-fluidization is achieved.

Spouting with aeration, as far as its visible characteristics are concerned, is similar to spouting in a standard spouted bed. This regime is stable and the extent of this regime depends on several factors. The regime area for polystyrene was greater than for HDPE and millet. HDPE and millet particles both have very low maximum spoutable bed height - H_m of 45 and 48 cm respectively for an orifice diameter of 1.91 cm - whereas the corresponding H_m for polystyrene particles is 90 cm. With a greater orifice diameter, $d_i = 2.54$ cm, the line A-C shifts a little to the right, since a greater minimum spouting flow is then required [see Figures 5.12 and 5.13].

An increased orifice diameter results in a lower H_m value. This in turn reduces the regime area for spouting with aeration. The value of H_m has a strong effect on this regime area. For the case of millet and HDPE at $d_i = 2.54$ cm, this regime is so restricted that $(\frac{q_{msa}}{Q_{mf}})_{\max}$ is less than 0.2, a value too low to accurately monitor the auxiliary flow. For this reason, determination of annulus gas velocity, solid circulation, fountain height, pressure drop and spout diameter could not be accomplished for millet and HDPE.

5.2.2 Minimum Fluid Flowrate:

The experimental data plotted as line A-C in Fig. 5.1 and Fig.

5.6 through Fig. 5.13 showed that the minimum total fluid flow rate followed a linear relationship and can be represented by

$$\left(\frac{q_{msa}}{Q_{mf}}\right) = K \left[1 - \left(\frac{Q_{mf}}{Q_{ms}}\right) \left(\frac{Q_{msa}}{Q_{mf}}\right) \right] \quad (5.1)$$

This linear equation terminates at point C due to the formation of another flow regime. Equation (5.1) is therefore valid for

$$\left(\frac{q_{msa}}{Q_{mf}}\right) < \left(\frac{q_{msa}}{Q_{mf}}\right)_{\max} \quad (5.2)$$

$$\left(\frac{Q_{msa}}{Q_{mf}}\right) < \left(\frac{Q_{msa}}{Q_{mf}}\right)_{\min} \quad (5.3)$$

values for Q_{ms} and Q_{mf} were determined experimentally. Total minimum fluid flow at a given bed height under this condition, noting Equation (2.5), is then given by

$$\frac{Q_{T,msa}}{Q_{mf}} = K + \left[1 - K \frac{Q_{mf}}{Q_{ms}} \right] \frac{Q_{msa}}{Q_{mf}} \quad (5.4)$$

Earlier work by Vukovic (1972) and Littman (1974, 1976) indicated that

$$\left(\frac{q_{msa}}{Q_{mf}}\right) = 1 - \left(\frac{Q_{mf}}{Q_{ms}}\right) \left(\frac{Q_{msa}}{Q_{mf}}\right)$$

Eq. (2.4) was valid for $Q_{ms} < Q_{mf}$ and it also showed that the sum of central flow and auxiliary flow for spouting with aeration lies between Q_{ms} and Q_{mf} i.e.

$$Q_{ms} < Q_{T,msa} < Q_{mf}$$

However, for $Q_{ms} > Q_{mf}$, the apparent intercept on the ordinate axis of line A-C on a plot of $(\frac{q_a}{Q_{mf}})$ versus $(\frac{Q}{Q_{mf}})$ of Figure 5.1 and Figures 5.6 through 5.13 was found to be always greater than unity. Equation (5.1) is therefore an improvement over the original equation (2.4) in incorporating a parameter K to account both for cases where $Q_{ms} < Q_{mf}$ and $Q_{ms} > Q_{mf}$. The values of K and the slopes of the line A-C for the various runs are summarized in Table 2. Since the slope is greater than unity in all cases, it is confirmed that the sum of the auxiliary flow and central flow required for spouting with aeration is always equal to or greater than the minimum spouting velocity, even when $Q_{ms} > Q_{mf}$.

From Table 2, values of the slope for a given material appear to be insensitive to bed height and to inlet orifice diameter.

The coordinate of point C is the limiting value for spouting with aeration, and is the minimum condition for spout-fluidization. The sum of $(\frac{q_{msa}}{Q_{mf}})_{max}$ and $(\frac{Q_{msa}}{Q_{mf}})_{min}$ equals the normalized minimum fluid flowrate for spout-fluidization, and at this condition, the given bed height is termed the "maximum bed height for spout-fluidization", H_{msf} . H_{msf} differs from the maximum spoutable bed height, H_m , in its dependence on

Table 2: Values of Empirical Constant K and Slopes of the Line A-C for Different Systems Investigated.

Material	H (cm)	d _i (cm)	Slope	K
Polystyrene	30	1.91	1.30	1.27
Polystyrene	45	1.91	1.27	1.38
Polystyrene	60	1.91	1.25	1.47
HDPE	30	1.91	1.28	1.46
HDPE	35	1.91	1.24	1.52
Millet	30	1.91	1.54	1.39
Millet	45	1.91	1.77	1.75
Polystyrene	45	2.54	1.29	1.50
Millet	30	2.54	1.36	1.61

the value of $(\frac{q_{msa}}{Q_{ms}})_{\max}$. In other words, every bed height could be

termed as H_{msf} at a given $(\frac{q_{msa}}{Q_{mf}})_{\max}$.

Values of $(\frac{q_{msa}}{Q_{mf}})_{\max}$ and $(\frac{Q_{msa}}{Q_{mf}})_{\min}$ appear in Table 3. As shown in Table 3, the value of $(\frac{q_{msa}}{Q_{mf}})_{\max}$ appears to be sensitive to H_m .

Polystyrene particles having a high value of H_m , has a higher $(\frac{q_{msa}}{Q_{mf}})_{\max}$ than HDPE and millet, which have much lower H_m value.

Equation (5.1), together with Equations (5.2) and (5.3), provide an easy means for estimating the minimum fluid flowrate for spouting with aeration. However, general correlations encompassing a wide range of experimental data under various conditions are needed for estimation of K , $(\frac{q_{msa}}{Q_{mf}})_{\max}$ and $(\frac{Q_{msa}}{Q_{mf}})_{\min}$.

Table 3: Normalized Values of Central and Auxiliary Flow at Minimum Spout-Fluidization Conditions and the Maximum Spoutable Heights for the Systems Investigated.

Material	H (cm)	d _i (cm)	$\left(\frac{q_{msa}}{Q_{mf}}\right)_{\max}$	$\left(\frac{Q_{msa}}{Q_{mf}}\right)_{\min}$	H _m (cm)
Polystyrene	30	1.91	0.75	0.40	90
	45	1.91	0.83	0.45	
	60	1.91	0.78	0.56	
HDPE	30	1.91	0.30	0.91	45
	35	1.91	0.30	0.99	
Millet	30	1.91	0.40	0.65	48
	45	1.91	0.20	0.87	
Polystyrene	45	2.54	0.54	0.74	80
Millet	30	2.54	0.18	0.88	36

CHAPTER 6

LONGITUDINAL ANNULAR GAS VELOCITY AND OVERALL BED

PRESSURE DROP FOR SPOUTING WITH AERATION

6.1 Measurement Technique

6.1.1 Longitudinal Annular Gas Velocity:

Annular gas flow in the cylindrical section was determined from measurement of the vertical static pressure gradient along the annulus. Lim (1975) found that this measurement technique was reliable. The static pressure probe shown in Fig. 3.7 was supported from above half-way between the spout-annulus interface and the column wall. No significant radial pressure gradient was found in the cylindrical section. The longitudinal positions of the probe extended from 2 cm below the bed surface to the cylindrical-conical junction. Measurements were taken at 10 cm intervals, reduced to 5 cm near the conical-cylindrical junction due to a more rapid change in vertical gas velocity in this section.

Gas flow measurement in the conical section was complicated by the uncertainty in the voidage of this section. It is generally accepted that the voidage in this section is lower due to particles realigning and compacting as they move downward (Eljas (1975)). Gas flow in the conical region was not determined in this work.

The static pressure technique relies on the assumption that the state of the annulus is the same as in a loosely packed bed of the same material. The static pressure gradient in a loosely packed bed is a function of the gas velocity. For a given solid material, a calibration

curve of static pressure gradient versus gas velocity obtained in a loosely packed bed was used in this work to determine the local gas velocity in the annulus. Calibrations were obtained in the same half-column with bed height greater than the maximum spoutable bed height, H_m . Measurements made at two radially different positions showed that there was insignificant radial variation of static pressure drop. The calibration results are summarized in Appendix I.

6.1.2 Overall Bed Pressure Drop

An inclined U-tube manometer filled with liquid of specific gravity 0.827 provided the measurement of overall bed pressure drop. The frictional pressure drop due to particulate material at a given mass flowrate was obtained by the procedure proposed by Epstein and Mathur (1974). For the pressure tap located upstream of the inlet orifice plate, positive values of solids-free pressure drop were obtained with increasing central air flow. On the other hand, negative values of solids-free pressure drop were obtained for the pressure tap located at the orifice plate itself. Since the frictional pressure drop is a positive quantity, the cancelling effect with the negative solids-free pressure drop results in a narrower range of manometer fluid linear travel, allowing lighter specific gravity fluids to be used. Thus readability was enhanced. In this work, only readings from the pressure tap located at the orifice plate were used.

6.2 Results and Discussion

6.2.1 Longitudinal Annular Gas Velocity:

Figure 6.1 shows that the superficial gas velocity in the annulus

Fig. 6.1 Effect of auxiliary gas flow on annulus gas superficial velocity [$Q_T / Q_{ms} = 1.25$, $d_i = 1.91 \text{ cm}$]

increased with bed level and that the increase was more rapid in the lower section. A comparison with results of annulus superficial gas velocity obtained by Lim (1975) is also shown in Fig. 6.1. The difference between his results and the results obtained in this work is probably due to differences in the dimensions and geometry of the polystyrene particles used. Annulus gas velocity was also found to increase with increasing auxiliary flow. The increase was more substantial at the lower section of the annulus. The percentage increase of annulus gas velocity due to auxiliary flow is shown in Fig. 6.2. As much as a fifty percent increase could be realized near the conical-cylindrical junction when the auxiliary flow comprised half of the total flow supplied. It was noted, however, that the percentage increase due to auxiliary flow rapidly dropped off with increasing bed level. The results indicate that additional external fluid supplied directly into the annulus of a standard spouted bed can be accommodated without upsetting the basic behaviour of spouting. However, there is a limit to how much auxiliary flow can be introduced before spout instability sets in resulting in spout-fluidization.

Figure 6.3 shows that, as in a standard spouted bed, the annulus gas velocity at a given bed level was independent of the overall bed height.

A plot of U_A/U_{AHm} versus Z/H is shown in Figure 6.4. The values of U_{AHm} were obtained by extrapolating U_A to H_m . It was found that at H_m , $U_{AHm} < U_{mf}$, a result reported earlier by Lim (1975). Experimental data were compared with the only existing equation derived by Littman (1976) for a flat-base column, Eq. 2.10. However, U_{mf} is replaced by

Fig. 6.2 Percent increase of annulus gas velocity due to auxiliary flow [$Q_T / Q_{ms} = 1.25$, $d_i = 1.91 \text{ cm}$]

Fig. 6.3 Annulus gas superficial velocity for different heights [system: polystyrene, $d_i = 1.91$ cm, $Q_T/Q_{ms} = 1.25$]

Fig. 6.4 Relationship between U_A/U_{AHm} and Z/H for different values of r_a [$d_i = 1.91$ cm]

U_{AHm} at $Z=H_m$ so that Equation (2.10) becomes

$$\frac{U_A(Z)}{U_{AHm}} = 1 - (1 - \frac{q}{Q_{mf}})(1 - h_{msf} \frac{Z}{H})^3 \quad (6.1)$$

For $q/Q_{mf} = 0$, equation (6.1) reduces to the modified Mamuro Hattori equation,

$$\frac{U_A(Z)}{U_{AHm}} = 1 - (1 - Z/H_m)^3 \quad (6.2)$$

The maximum bed height for spout-fluidization is always smaller or equal to the maximum spoutable bed depth, i.e. $H_{msf} \leq H_m$. (Values of H_m are given in Table 3). If H_{msf} is replaced by H_m , equation 6.1 is found to correlate the data for polystyrene particles with $H = 60$ cm ($h_{msf} = h = 0.67$) quite well but underpredicts the data for HDPE particles with $H = 35$ cm ($h_{msf} = h = 0.78$). However, at $H_{msf}/H_m = 0.92$ or $h_{msf} = 0.85$, a better fit is achieved for HDPE particles.

In Figure 6.5, the assumption of $H_{msf} = H_m$ or $h_{msf} = 0.75$ again caused Equation 6.1 to underpredict the data. At $H_{msf}/H_m = 0.83$ or $h_{msf} = 0.90$, however, Equation 6.1 produces a better fit. For the systems investigated, the relationship between maximum bed height for spout-fluidization and inlet orifice diameter and auxiliary flow could not be established. Therefore the assumption of H_{msf} being independent of auxiliary flow for the cases studied is also subjected to error. It is apparent that h_{msf} is an adjustable parameter whose value needs to be interpreted more precisely. Furthermore, the equation for H_{msf} (Littman

Fig. 6.5 Relationship between U_A/U_{AHm} and Z/H for different values of r_a [$d_i = 2.54$ cm, $H = 60$ cm]

1976, 1983) was derived based on a flat-base column. The applicability of such an equation may be acceptable for predicting the annulus gas velocity profile of the cylindrical part but not the conical part of the present column.

The effect of auxiliary gas entry position on annulus gas velocity is shown in Figure 6.6. Configuration #2 (see Section 4.1) was found to yield greater annulus gas velocity than configuration #3. In configuration #3, for which the auxiliary flow entry position was close to the spout, the observed lower annulus gas velocity was likely due to gas bypassing directly into the spout.

The effect of inlet orifice diameter on annulus gas velocity is shown in Fig. 6.7. As in standard spouted beds, the greater the orifice plate opening, the greater the portion of spouting gas which flows into the annulus, thereby contributing to greater overall annulus gas velocity.

The effect of increasing the fraction of auxiliary flow on the annular gas flow at different bed levels is shown in Fig. 6.8. The fraction of gas flowing through the annulus increases almost linearly with an increase of auxiliary flow. The total flow in the annulus was always found to be greater than the auxiliary flow supplied indicating that there is no net cross-flow of fluid from the annulus to the spout in the cylindrical section for the conditions studied. However, for configuration #3, the tendency of fluid cross-flow from the annulus to the spout may be high, especially in the cone region near where the auxiliary flow is being introduced.

Fig. 6.6 Relationship between U_A/U_{mf} and Z/H for different auxiliary gas flow entry configurations [polystyrene, $d_i = 2.54$ cm, $H = 60$ cm, $Q_T/Q_{ms} = 1.25$]

Fig. 6.7 Relationship between U_A/U_{mf} and Z/H for different inlet orifice diameters [polystyrene, $H = 60$ cm, $Q_T/Q_{ms} = 1.25$]

Fig. 6.8 Effect of the fraction of auxiliary flow on the fraction of gas which flows through the annulus (polystyrene, $H= 60 \text{ cm}$, $d_i= 1.91 \text{ cm}$, $Q_T= 8 \text{ l/s}$)

6.2.2 Overall Bed Pressure Drop

The pressure gradient for an incipiently fluidized bed having voidage ϵ_A is given by

$$\frac{-dP}{dZ} = (\rho_s - \rho)(1 - \epsilon_A)g \quad (6.3)$$

The pressure drop per unit bed height for the present experiments was fitted by

$$\frac{\Delta P}{\Delta H} = f(U) \quad (6.4)$$

where $f(U)$ is a 5th order polynomial given in Appendix I.

Under minimum fluidization condition, Equation 6.4 becomes

$$\frac{\Delta P_f}{\Delta H} = f(U_{mf}) \quad (6.5)$$

Values of $(\rho_s - \rho)(1 - \epsilon_A)$ and $f(U_{mf})$ are shown in Table

4. It is seen that $f(U_{mf}) < (\rho_s - \rho)(1 - \epsilon_A)g$. Values of overall pressure drop for spouting with aeration normalized with respect to pressure drop for fluidized bed i.e. $(\frac{\Delta P_s}{\Delta P_f})$ are plotted versus q_{msa}/Q_{mf} in Figure 6.9 for the various systems investigated. Comparison of the overall pressure drop data was made with equation 2.22 with different h_{msf} values. The data indicate that overall pressure drop shows a linear relationship with q_{msa}/Q_{mf} , and the rate of increase is not as rapid as predicted by Eq. 2.22. Furthermore, extrapolation of these data to $q_{msa}/Q_{mf} = 1$ do not coincide with the ordinate equal to unity as is predicted. Physically, for $(q/Q_{mf}) > (q_{msa}/Q_{mf})_{max}$ the bed is in the submerged jet in fluidized bed regime. This seems to indicate that different correlations may be required to treat each regime separately.

Table 4: Comparison of the Values of $(\rho_s - \rho)(1 - \epsilon_A)g$ and $f(U_{mf})$

Material	$\frac{dP}{dZ} = (\rho_s - \rho)(1 - \epsilon_A)g$ (mm H ₂ O/m)	$\left(\frac{\Delta P_f}{\Delta H}\right) = f(U_{mf})$ (mm H ₂ O/m)	$\frac{\Delta P_f / \Delta H}{(\rho_s - \rho)(1 - \epsilon_A)g}$
Polystyrene	594	445	0.75
HDPE	578	488	0.84
Millet	6.86	650	0.95

$(\Delta P_s / \Delta P_f)_{msa}$

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Fig. 6.9 Relationship between $(\Delta P_s / \Delta P_f)_{msa}$ and q_{msa} / Q_{mf} for different inlet orifice diameters, bed heights and auxiliary flow configuration [polystyrene]

CHAPTER 7

PARTICLE VELOCITY IN THE ANNULUS

7.1 Measurement Technique

Particle velocity in the annulus was determined by visually following and timing a marked particles at the annulus wall interface over a distance of 10 cm, starting at 5 cm below the bed surface. Particle velocity was found to vary at different positions along the perimeter of the column. It was found to be lowest at the junction of the flat front surface and the curved surface. This was primarily caused by the wall effect due to the convergence of the two surfaces. To minimize this problem and simulate full-column behaviour, measurements were made near the mid-section of the curved surface, far from the junctions.

Under spout-fluidization, movement of particles in the upper part of the annulus was of a stick-slip type and this measurement method could not be employed reliably. At deeper bed heights, the fountain was observed to sway from side to side, affecting the continuity of particle movement in the annulus. The situation was corrected by placing an inverted half-funnel at the top of the fountain.

7.2 Results and Discussion

The solid mass circulation rate is expressed as

$$Sr = A_A * V_p * (1 - \epsilon_A) * \rho_s \quad (7.1)$$

with ϵ_A and ρ_s being constants for a given material. For spouting with aeration, the changes in spout cross-sectional area in the upper section

of the bed were found to be insignificant compared with the annulus cross-sectional area. The maximum percentage change in the annulus cross-sectional area was less than 3% for different central and auxiliary flow combinations. Particle velocity was therefore used as a representation of the overall solids circulation rate in the bed.

Particle velocity was found to increase with bed height as shown in Fig. 7.1. This occurs because for a deeper and longer spout, there is a greater possibility of particles being entrained at the spout-annulus interface and carried over into the fountain where they rain back onto the annulus.

The effect of auxiliary flow on particle velocity at various bed heights is shown in Fig. 7.2. For deeper beds, there was an increase of particle velocity with an increase of auxiliary flow. However, for shallower beds, the reverse was found to be true. One possible explanation for the observed effect at shallow bed height is that the sizeable increase of annulus gas velocity due to auxiliary flow tends to oppose the downward moving annular solids, even though the auxiliary flows may induce a loosening or freer motion of the annular solids. The increase of the annulus gas flow diminishes with bed level. Hence for deeper beds, the loosening effect coupled with the sheer bulk weight of the annular solids may result in more particles being entrained at the spout-annulus interface, thereby contributing to greater solids circulation.

The effects of auxiliary flow entry positions on the particle velocity are shown in Figures 7.3a and 7.3b. For configuration #2 (see

Fig. 7.1 Relationship between particle velocity, V_p , and Q_T/Q_{ms} for different bed heights [polystyrene, $d_i = 1.91$ cm, $q = 0$]

Fig. 7.2 Relationship between particle velocity, V_p , and the proportion of auxiliary flow r_a for different bed heights [polystyrene, $d_i = 1.91$ cm, $Q_T/Q_{ms} = 1.4$]

Fig. 7.3a Relationship between particle velocity, V_p , and fraction of auxiliary flow with configuration #2 [polystyrene, $H = 60$ cm, $d_i = 1.91$ cm]

Fig. 7.3b Relationship between particle velocity, V_p , and fraction of auxiliary flow with configuration #3 [polystyrene, $H=60$ cm, $d_i=1.91$ cm]

Section 4.1), particle velocity in the annulus was found to increase with increasing auxiliary flow. However, with configuration #3, there was an initial slight decrease of particle velocity before it increased with increasing auxiliary flow. A greater increase of particle velocity in the case of configuration #2 than configuration #3 was most likely due to more loosening of the annular solids. In configuration #3, a greater percentage of auxiliary flow bypassed directly into the spout. However, with increasing auxiliary flow, solids velocity was observed to increase substantially beyond a certain value of auxiliary flow. As noted by Yang (1983), there is a minimum aeration required to significantly increase solid circulation. From Figures 7.3a and 7.3b, these values, expressed as the ratio of auxiliary flow to total flow rate, were found to be approximately 0.25 and 0.30 respectively.

For the case of HDPE (see figures 7.4), addition of auxiliary flow was found to decrease the solids circulation. For configuration #2, there was a slight increase of particle velocity at higher auxiliary flow. For both configurations, the overall solids circulation rate was less than that without any aeration.

It may therefore be noted that the solids circulation rate for a given material and bed geometry is affected by the following factors:

Factors promoting increased circulation:

- higher gas flowrate in the spout
- loosening of annular solids due to auxiliary flow.
- increased bed height

Factor leading to reduced circulation:

- higher gas flow in the annulus.

Fig. 7.4a Relationship between particle velocity and fraction of auxiliary flow with configuration #2 [HDPE, $H = 30$ cm, $d_i = 1.91$ cm]

Fig. 7.4b Relationship between particle velocity and fraction of auxiliary flow with configuration #3 [HDPE, $H = 30$ cm, $d_i = 1.91$ cm]

CHAPTER 8

FOUNTAIN SHAPE AND HEIGHT AND SPOUT SHAPE AND DIAMETER

8.1 Measurement technique

Fountain height was measured directly as the distance from the bed surface to the top of the fountain. Spout diameter was measured directly at an interval of 10 cm starting from the bed surface.

8.2 Results and Discussion

8.2.1 Fountain Shape and Height

Fig. 8.1 shows that the height of the fountain increases with total gas flow, and that a shallower bed shows a greater fountain height for a given Q_T/Q_{ms} ratio.

The effect of increasing the proportion of auxiliary flow on the height of the fountain is shown in Fig. 8.2. Fountain height decreases with increasing fraction of auxiliary flow. This may be primarily due to redistribution of fluid between the spout and annulus - less fluid flows through the spout. Because of lower fluid flow through the spout, the particle velocity in the spout is also reduced, resulting in a lower fountain height. (Grace and Mathur 1978).

For a desired fountain height, a greater Q_T/Q_{ms} ratio is needed for the case of spouting with aeration than for pure spouting. The shape of the fountains is, however, quite different for the two cases. With aeration, the fountain shows little tendency to overdevelop and is

Fig. 8.1 Effect of Q_T/Q_{mf} on H_F for different values of bed heights [polystyrene, $q=0$, $d_i=1.91$ cm]

Fig. 8.2 Relationship between fountain height and the fraction of auxiliary flow with configuration #1 for different bed heights [polystyrene, $Q_T/Q_{ms}=1.4$, $d_i=1.91$ cm]

usually well defined. It was also observed to have a greater solids concentration. This preliminary observation proved to be of significant importance since greater solids concentration could be achieved at the same fountain height, but at the expense of the extra flow required.

The effect of auxiliary flow entry position is shown in Figures 8.3a and 8.3b. There was no significant difference in the fountain shape and height.

8.2.2 Spout Shape and Diameter

The spout assumed different shapes depending on the solid properties. Spout shapes are shown in Fig. 8.4. The spout diameter for millet particles was nearly constant along its length (Fig. 8.4a). For polystyrene particles, the spout diameter was also observed to be nearly constant except near the bed surface where the spout showed some divergency (Fig. 8.4b). For the case of HDPE particles, a continuously divergent spout shape was observed (Fig. 8.4c). A marked change of spout shape near the entrance region was noted at high auxiliary flow with configuration #3. The close proximity of the auxiliary flow entry position to the spout-annulus interface created a significantly "distorted" spout shape (Fig. 8.4d).

Predicted spout diameters due to McNab's empirical correlation are given in Table 5.

A comparison of the values shown in Table 5 with the results of spout diameter (see Appendix VI) under various central and auxiliary flow combinations indicated that McNab's correlation predicted the data very well if the sum of central and auxiliary mass flows was taken as the mass flowrate.

Fig. 8.3a Effect of the fraction of auxiliary flow with configuration #2 on fountain height, H_F , [polystyrene, $H = 60$ cm, $d_i = 1.91$ cm]

Fig. 8.3b Effect of the fraction of auxiliary flow with configuration #3 on fountain height, H_F , [polystyrene, $H = 60$ cm, $d_i = 1.91$ cm]

Fig. 8.4 Sketches showing spout shapes for various solid materials investigated.

Table 5: Predicted Values of Average Spout Diameter Using McNab's Empirical Correlation [Polystyrene particles].

Q	d _s	Q	d _s
3.5	2.60	7.0	3.66
4.0	2.78	7.5	3.78
4.5	2.94	8.0	3.90
5.0	3.10	8.5	4.02
5.5	3.25	9.0	4.14
6.0	3.39	9.5	4.25
6.5	3.53	10.0	4.35

CHAPTER 9

CONCLUSIONS AND SUGGESTIONS FOR FURTHER WORK

9.1 Conclusions

At various central flow and auxiliary flow combinations, four different flow regimes of behaviour were established: Packed bed; spouting with aeration; spout-fluidization and submerged jet in a fluidized bed. These flow regimes shared many similarities to previously determined regime maps. For the case of spout-fluidization, spout instability is developed as a result of bubble formation in the spout near the bed surface.

The minimum total fluid flowrate for spouting with aeration and spout-fluidization is always found to be greater than the minimum spouting flowrate. An improved correlation for predicting minimum fluid flowrate for spouting with aeration is proposed. Greater fluid percolation through the annulus, especially in the lower section, could be achieved by increasing the amount of auxiliary flow. An increase of as much as 50% over the corresponding pure spouting case could be realized near the conical-cylindrical junction for a 1:1 split between central and auxiliary flow. Annulus gas velocity data were in good agreement with an equation due to Littman (1976) if U_{mf} is replaced by U_{AHm} and if h_{msf} is treated as an adjustable parameter.

Solids circulation rate was found to be little affected by small rates of aeration. At higher rates of aeration and deeper bed heights, an increase of solids circulation was obtained. However, at lower bed height, the solids circulation rate was found to decrease.

The overall bed pressure drop for spouting with aeration under minimum conditions increases linearly with increasing auxiliary flow.

Fountain height was found to decrease with increasing fraction of auxiliary flow. For spouting with aeration, solids in the fountain were observed to have a greater concentration than pure spouting at a given fountain height. The empirical correlation due to McNab was found to predict the average spout diameter very well under various central and auxiliary flow combinations if the sum of central and auxiliary flows is used in the correlation.

9.2 Recommendations For Further Work

Further solid materials suitable for spout-fluid beds that will extend the regime of spouting with aeration should be investigated.

A correlation for K , together with factors that affect the values for H_{msf} , $(\frac{q_{msa}}{Q_{mf}})_{\max}$ and $(\frac{Q_{msa}}{Q_{mf}})_{\min}$, should be sought.

The regime of spout-fluidization with respect to gas distribution, frequency and causes of bubble formation, solids circulation and pressure drop should receive further attention.

NOTATION

A_A	- Cross-sectional area of the annulus	(cm ²)
A_c	- Cross-sectional area of the column	(cm ²)
D_c	- Column diameter	(cm)
G	- Gas mass flowrate	(g/cm ² s)
H	- Bed height	(cm)
H_F	- Fountain height	(cm)
H_m	- Maximum spoutable bed height	(cm)
H_{msf}	- Maximum height for spout-fluidized bed	(cm)
K	- Empirical constant	
Q	- Central spouting flow	(l/s)
Q_{ms}	- Minimum spouting flowrate	(l/s)
Q_{mf}	- Minimum fluidization flowrate	(l/s)
Q_{msa}	- Central spouting flow under the condition of minimum spouting with aeration	(l/s)
$Q_{T,msa}$	- Sum of central spouting flow and auxiliary flow under the condition of minimum spouting with aeration, = $Q_{msa} + q_{msa}$	(l/s)
Q_T	- Sum of central spouting flow and auxiliary flow	(l/s)
S_r	- Solid circulation rate	(g/s)
U	- gas superficial velocity	(cm/s)
$U_A(Z)$	- Annulus gas velocity at bed level Z	(cm/s)
U_{AH}	- Annulus gas velocity at $Z = H$	(cm/s)
U_{AHm}	- Annulus gas velocity at $H = H_m$	(cm/s)

U_{mf}	- Minimum fluidization velocity	(cm/s)
U_{ms}	- Minimum spouting velocity	(cm/s)
U_{SH}	- Gas superficial velocity in the spout at $Z = H$	(cm/s)
$U_{T,msf}$	- Minimum fluid superficial velocity for spout-fluidization	(cm/s)
$U_{T,msa}$	- Minimum fluid superficial velocity for spouting with aeration	(cm/s)
V_p	- particle velocity in the annulus	(cm/s)
Z	- Height coordinate, $Z = 0$ at base	(cm)
d_i	- diameter of inlet orifice	(cm)
d_p	- Particle diameter, equi-volume sphere diameter	(cm)
d_s	- Spout diameter	(cm)
g	- Acceleration of gravity	(cm/s ²)
h	$= H/H_m$	
h_{msf}	$= H/H_{msf}$	
$\dot{h}_{msf}$	$= 1 - (1 - h_{msf})^3$	
q	- Auxiliary flow	(l/s)
q_{msa}	- Auxiliary flow under the conditions of minimum spouting with aeration	(l/s)
q_a	- Auxiliary flow through all five subdistributor chambers	(l/s)
q_u	- Auxiliary flow through upper two subdistributors only	(l/s)
q_l	- Auxiliary flow through lower two subdistributors only	(l/s)

$$r_a = q_a / Q_T$$

$$r_l = q_l / Q_T$$

$$r_u = q_u / Q_T$$

$$\frac{\Delta P}{\Delta H} = f(U), \text{ pressure drop per unit bed height}$$

$$\Delta P_f - \text{Pressure drop of the fluidized bed} \quad (\text{cm H}_2\text{O})$$

$$\Delta P_{mf} - \text{Pressure drop of the bed at minimum fluidization conditions} \quad (\text{cm H}_2\text{O})$$

$$\Delta P_s = \text{Overall pressure drop of the spout-fluid bed} \quad (\text{cm H}_2\text{O})$$

$$\frac{\Delta P_s}{\Delta H} - \text{Pressure drop of the spout-fluid bed per unit bed height}$$

$$\frac{\Delta P_f}{\Delta H} - \text{Pressure drop of the fluidized bed per unit bed height}$$

$$\left(\frac{dP}{dz}\right) - \text{Differential bed pressure drop}$$

$$\varepsilon_A - \text{Voidage in the annulus}$$

$$\varepsilon_o - \text{Spout voidage at bed surface level}$$

$$\rho - \text{Fluid density} \quad (\text{g/cm}^3)$$

$$\rho_b - \text{Solids bulk density} \quad (\text{g/cm}^3)$$

$$\rho_s - \text{Density of solid particle} \quad (\text{g/cm}^3)$$

$$v_{o \text{ max}} - \text{Particle velocity along axis of column at bed surface} \quad (\text{cm/s})$$

$$\emptyset - \text{Empirical constant,}$$

$$\emptyset = 0.20 d_p^{-0.320} d_i^{0.235} H^{0.160}$$

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A P P E N D I C E S

Appendix I

Pressure Drop-gas Velocity Equations for Loosely-packed Beds of the Solids Material Investigated.

Material	a ₀	a ₁	a ₂	a ₃	a ₄	a ₅
Polystyrene	-71.1197	191.7383	-153.3241	61.5574	-11.7256	0.8557
HDPE	191.6519	-361.1952	300.2185	-114.0607	20.6864	-1.4434
Millet	44.6259	- 18.0998	- 5.5045	7.3752	- 1.7256	0.1230

Materials	b ₀	b ₁	b ₂	b ₃	b ₄	b ₅
Polystyrene	24.1395	-3.4035	1.8885 E-1	-4.9307 E-3	6.2246 E-5	-3.0515 E-7
HDPE	2.3423 E-1	8.518 E-3	7.2766 E-4	3.0771 E-6	-4.4759 E-8	-8.5587 E-11
Millet	-17.1777	1.8143	-4.9902 E-2	3.8694 E-6	1.9004 E-5	-1.9886 E-7

Equation:

$$U_A \text{ (cm/s)} = a_0 + a_1 \Delta H + a_2 \Delta H^2 + a_3 \Delta H^3 + a_4 \Delta H^4 + a_5 \Delta H^5$$

$$\Delta H \text{ (cm)} = b_0 + b_1 U_A + b_2 U_A^2 + b_3 U_A^3 + b_4 U_A^4 + b_5 U_A^5$$

Where ΔH is the value of the differential linear movement of the manometer fluid in cm of a 15° inclined manometer filled with liquid of specific gravity 0.827.

The value of ΔH corresponds to 2 cm of bed level.

SOLIDS-FREE PRESSURE DROP-CENTRAL FLOW CORRECTION CURVE

APPENDIX III

CALIBRATION CURVES FOR ORIFICE PLATE FLOWMETERS

Fig. III.1 Calibration curve for orifice flowmeter for the central airflow line

Fig. III.2 Calibration curve for orifice flowmeter for each auxiliary flow line

Air flowrate correction to account for different air temperature and inlet upstream pressure:

- (i) Correction factor, f , for orifice flowmeter for the central air flow line.

$$f = 0.6164 \left[\frac{T_N + 273}{P_N + 101.325} \right]^{1/2}$$

- (ii) Correction factor, f , for orifice flowmeter for each of the auxiliary air flow lines

$$f = 0.6945 \left[\frac{T_N + 273}{P_N + 101.325} \right]^{1/2}$$

Where

T_N - air temperature ($^{\circ}\text{C}$)

P_N - inlet upstream pressure (kPa)

APPENDIX IV

DIMENSIONS FOR ORIFICE PLATE FLOWMETERS

Dimensions for orifice flowmeters, (mm.)

	a	b	c	d
Central air-flow line	25.4	12.7	25.4	12.7
Auxiliary air flow lines	12.7	6.35	12.7	6.35

Material of construction : brass

APPENDIX V

EXPERIMENTAL CONDITIONS DESIGNATION

The run number is expressed in the form of

Run N1-N2-N3

Where the values of N's are as follow:

- N1 - Refers to the combination of bed material, bed height and inlet orifice plate diameter (see Table V.1).
- N2 - Refers to the dependent variable that is being determined. (see Table V.2).
- N3 - Refers to auxiliary flow configuration (see Table V.3).

Table V.1 Values of N1 for Different Combination of Bed Material, Bed Height and Inlet Orifice Plate Diameter.

N1	Materials	H (cm)	di (cm)
1	Polystyrene	30	1.91
2	Polystyrene	45	1.91
3	Polystyrene	60	1.91
4	HDPE	30	1.91
5	HDPE	35	1.91
6	Millet	30	1.91
7	Millet	45	1.91
8	Polystyrene	45	2.54
9	Polystyrene	60	2.54
10	Millet	30	2.54

Table V.2 Values of N2 for Different Dependent Variables Investigated.

N2	Dependent Variables
1	Regime map
2	Gas Velocity in the annulus
3	Overall bed pressure drop
4	Particle velocity
5	Fountain height
6	Spout diameter

Table V.3 Values of N3 for Different Auxiliary Flow Configuration
(see Section 4.1)

N3	Auxiliary Flow Configurations
1	#1
2	#2
3	#3

APPENDIX VI
EXPERIMENTAL DATA

RUN# 3-2-1

Q	8	6	5	4
q _a	0	2	3	4
Z	U _A	U _A	U _A	U _A
58	53.67	54.24	56.10	57.09
48	51.21	52.22	53.95	55.53
38	47.09	49.89	51.93	54.53
28	41.18	45.14	48.42	51.64
23	35.84	41.18	45.44	49.75
18	29.29	39.51	43.02	48.13

RUN# 9-2-1

Q	8.5	6.5	5.5	4.5
q _a	0.0	2.0	3.0	4.0
Z	U _A	U _A	U _A	U _A
58	56.85	57.90	58.15	58.80
48	55.35	56.24	56.53	58.35
38	52.16	54.24	55.39	56.96
28	47.23	50.77	53.24	55.53
23	42.71	48.27	51.93	54.96
18	38.96	46.19	49.80	54.39
15	34.88	44.53	48.27	53.81

RUN# 2-2-1

Q	7.5	5.5	4.5
q _a	0	2	3
Z	U _A	U _A	U _A
43	48.85	52.00	54.16
38	46.07	50.45	52.59
33	44.15	48.23	50.96
28	40.76	46.15	49.16
23	35.70	41.95	46.43
18	28.38	39.65	43.75

RUN# 5-2-1

Q	10.5	8.5
q _a	0	2
Z	U _A	U _A
33	62.65	64.39
28	59.98	63.52
23	55.40	60.58
18	50.61	56.95
15	45.91	53.82

RUN# 9-2-2

Q	8.5	6.5	5.5
q _u	0	2	3
Z	U _A	U _A	U _A
58	56.85	58.80	59.00
48	55.35	56.81	57.94
38	52.16	54.81	57.09
28	47.23	51.93	55.80
23	42.71	49.89	53.95
18	38.96	47.83	53.09
15	34.88	46.34	51.93

RUN# 9-2-3

Q	8.5	6.5	5.5
q _u	0.0	2.0	3.0
Z	U _A	U _A	U _A
58	56.85	57.00	57.66
48	55.35	53.39	55.50
38	52.16	53.09	55.25
28	47.23	49.30	52.51
23	42.71	46.00	50.48
18	38.96	43.02	47.23
15	34.88	39.94	42.41

RUN# 1-3-1

q_a	ΔP_s
0	5.7
2	7.7
3	8.2

RUN# 1-3-2

q_u	ΔP_s
2	7.8
3	9.1

RUN# 1-3-3

q_l	ΔP_s
2	8.3
3	9.1

RUN# 3-3-1

q_a	ΔP_s
0	15.4
2	19.1
3	20.8
4	21.9

RUN# 3-3-2

q_u	ΔP_s
2	18.8
3	20.8
4	22.8

RUN# 3-3-3

q_l	ΔP_s
2	19.4
3	20.6
4	22.7

RUN# 2-3-1

q_a	ΔP_s
0	9.6
2	13.2
3	14.6
4	17.0

RUN# 8-3-1

q_a	ΔP_s
0	18.25
2	19.40
3	19.50

RUN# 9-3-1

q_a	ΔP_s
0	23.65
2	25.85
3	26.37

RUN# 4-4-2

q_u	Q	V_p			
0	8.5	4.22	3.94	4.29	4.17
	9.5	4.76	4.98	4.90	4.85
	10.5	5.59	5.38	5.55	5.65
1	8.0	3.73	3.48	3.51	3.72
	9.0	4.29	4.55	4.18	4.31
	10.0	4.98	5.03	5.00	4.93
2	7.0	3.95	3.98	4.02	3.98
	8.0	4.41	4.46	4.57	4.22
	9.0	5.10	4.67	4.83	5.00

RUN# 4-4-3

q_1	Q	V_p			
1	8	3.60	3.68	3.75	3.66
	9	4.20	4.44	4.61	4.42
	10	4.93	5.29	4.93	5.15
2	7	3.38	3.15	3.26	3.24
	8	3.94	3.82	3.91	3.89
	9	4.55	4.57	4.63	4.57
	10	5.15	5.18	5.03	5.24

RUN# 3-4-2

q_u	Q	V_p			
2	6.0	4.29	4.27	4.31	4.31
	7.0	5.00	4.95	5.10	4.90
	8.0	5.71	5.38	5.38	5.71
3	5.0	4.55	4.55	4.63	4.41
	6.0	5.18	5.24	5.13	5.29
	7.0	5.56	5.85	5.85	5.92
4	4.5	5.41	5.62	5.43	5.58
	5.0	5.85	5.56	5.81	5.81
	6.0	6.17	6.21	6.29	6.21

RUN# 3-4-3

q_1	Q	V_p			
2	6.0	4.07	4.00	4.08	4.01
	7.0	4.67	4.72	4.59	4.72
	8.0	5.29	5.38	5.49	5.35
3	5.0	4.27	4.15	4.10	4.26
	6.0	4.78	4.81	4.83	4.78
	7.0	5.32	5.32	5.62	5.56
4	4.5	5.35	5.26	5.29	5.13
	5.0	5.62	5.29	5.59	5.52
	6.0	5.75	5.75	6.06	5.78

RUN# 3-4-1

q _a	Q	V _p			
0	7	3.48	3.75	3.50	3.47
	8	4.00	4.29	4.39	4.18
	9	4.78	4.88	4.81	4.88
2	5.5	3.89	3.95	3.79	3.85
	6.5	4.67	4.52	4.72	4.67
	7.5	5.41	5.21	5.10	5.21
3	4.5	4.27	4.39	4.35	4.41
	5.5	5.08	4.85	4.88	4.85
	6.5	5.65	5.59	5.71	5.55
4	4.0	5.13	4.95	5.15	4.90
	4.5	5.43	5.26	5.18	5.35
	5.5	5.88	5.99	5.88	5.81

RUN# 2-4-1

q _a	Q	v _p			
0	7	2.74	2.68	2.63	2.63
	8	3.12	3.08	3.10	3.11
	9	3.64	3.68	3.39	3.47
2	5.5	2.86	2.97	2.89	2.90
	6.5	3.55	3.39	3.30	3.37
	7.5	3.62	3.84	3.78	3.95
3	4.5	3.12	3.08	3.11	3.09
	5.5	3.58	3.61	3.57	3.64
	6.5	3.94	3.92	4.03	4.13
4	3.5	3.62	3.40	3.39	3.57
	4.5	4.00	4.10	3.83	4.12
	5.5	4.44	4.42	4.37	4.29

RUN# 2-4-2

q _u	Q	v _p			
3	4.5	3.05	3.14	3.07	3.00
	5.5	3.68	3.55	3.64	3.70
	6.5	4.03	4.05	4.33	4.24

RUN# 2-4-3

q ₁	Q	v _p			
3	4.5	2.77	2.86	2.73	2.90
	5.5	3.38	3.30	3.38	3.36
	6.5	3.79	3.72	2.76	3.69

RUN# 1-4-1

q_a	Q	V_p			
0	6	1.45	1.43	1.43	1.45
	7	1.83	1.86	1.78	1.83
	8	2.15	2.26	2.20	2.16
2	4.5	1.44	1.49	1.45	1.43
	5.5	1.98	1.92	1.94	1.88
	6.5	2.39	2.32	2.37	2.33
3	3.5	1.47	1.48	1.41	1.39
	4.5	1.85	1.99	1.94	1.91
	5.5	2.35	2.42	2.44	2.38
3.5	3.0	1.34	1.32	1.30	1.36
	4.0	1.87	1.83	1.81	1.79

RUN# 1-4-2

q_u	Q	V_p			
2	4.5	1.50	1.40	1.45	1.41
	5.0	1.73	1.75	1.74	1.75
	6.0	2.24	2.22	2.35	2.25

RUN# 1-4-3

q_l	Q	V_p			
2	4	1.45	1.43	1.43	1.45
	5	1.83	1.86	1.78	1.83
	6	2.15	2.26	2.20	2.16

RUN# 1-5-2

q_u	Q	H_F	
2	4.5	8.5	8.5
	5.0	13.5	13.0
	6.0	23.0	22.5
3	3.5	6.0	5.0
	4.5	15.0	14.5
	5.5	24.0	24.0

RUN# 1-5-3

q_l	Q	H_F	
2	4.5	9.5	9.0
	5.0	14.0	13.5
	6.0	23.0	23.0
3	3.5	6.5	7.0
	4.5	16.0	15.5
	5.5	25.5	25.0

RUN# 2-5-1

q_a	Q	H_F	
0	6.5	5.0	4.5
	7.0	9.0	9.5
	7.5	14.0	14.0
	8.0	19.0	18.5
	8.5	22.5	23.5
	9.0	28.0	28.5
2	5.5	9.0	9.0
	6.5	17.0	17.5
	7.5	27.0	26.5
3	4.5	7.0	7.5
	5.5	16.5	17.0
	6.5	26.0	25.5
4	3.5	6.0	6.0
	4.5	14.5	15.0
	5.5	24.0	24.0

RUN# 2-5-2

q_u	Q	H_F	
3	4.5	7.0	7.0
	5.5	16.0	15.0
	6.5	25.0	24.0

RUN# 2-5-3

q_l	Q	H_F	
3	4.5	7.5	7.0
	5.5	17.0	16.0
	6.5	26.5	26.0

RUN# 3-5-2

q_u	Q	H_F	
2	6.0	8.5	8.5
	7.0	16.0	16.5
	8.0	25.0	25.5
3	5.0	6.0	6.0
	6.0	15.0	15.0
	7.0	24.0	24.5
4	4.5	8.5	8.5
	5.0	13.0	13.5
	6.0	22.5	22.5

RUN# 3-5-3

q_l	Q	H_F	
2	6.0	8.0	8.0
	7.0	16.5	16.5
	8.0	26.5	27.0
3	5.0	6.0	6.0
	6.0	15.0	15.0
	7.0	25.0	25.5
4	4.5	8.5	8.5
	5.0	13.5	13.5
	6.0	23.5	24.0

RUN# 1-5-1

q_a	Q	H_F	
0	5.5	7.5	7.5
	6.0	12.5	12.0
	6.5	17.0	16.5
	7.0	22.5	21.5
	7.5	27.0	26.5
	8.0	30.0	30.0
2	4.5	8.5	9.0
	5.5	18.0	18.0
	6.5	27.0	26.5
3	3.5	7.0	6.5
	4.5	16.0	15.5
	5.5	25.0	25.0
3.5	3.0	5.0	5.0
	4.0	14.0	14.5
	4.5	18.5	19.0

RUN# 3-5-1

q_a	Q	H_F	
0	7.0	3.0	3.0
	8.0	10.0	10.0
	9.0	19.5	19.0
2	5.5	5.0	5.0
	6.5	14.5	15.0
	7.5	20.0	20.5
4	4.0	6.5	6.5
	4.5	9.0	9.0
	5.5	19.0	19.5

RUN# 1-6-1

Z		30	20	10
Q	q _a	d _s		
6.0	0	3.50	3.50	3.50
7.0		3.75	3.75	3.75
8.0		4.00	4.00	3.75
4.5	2	3.50	3.25	3.00
5.5		3.75	3.50	3.50
6.5		4.00	4.00	3.50
3.5	3	3.00	3.00	2.75
4.5		3.75	3.50	3.25
5.5		4.00	4.00	3.50

RUN# 1-6-2

Z		30	20	10
Q	q _u	d _s		
4.5	2	3.25	3.25	3.00
5.0		3.50	3.25	3.00
6.0		4.00	3.75	3.25
3.5	3	3.00	3.00	2.75
4.5		3.50	3.25	3.00
5.5		4.00	3.75	3.50

RUN# 1-6-3

Z		30	20	10
Q	q ₁	d _s		
4.5	2	3.00	3.00	3.00
5.0		3.50	3.50	3.25
6.0		4.00	4.00	3.50
3.5	3	3.25	3.25	3.00
4.5		3.75	3.50	3.50
5.5		4.00	4.00	3.75

RUN# 2-6-1

Z		45	35	25	15
Q	q _a	d _s			
7.0	0	4.00	3.75	3.50	3.50
8.0		4.00	4.00	3.75	3.75
9.0		4.25	4.25	4.25	4.25
5.5	2	4.00	3.75	3.50	3.50
6.5		4.25	4.00	4.00	3.75
7.5		4.50	4.25	4.00	4.00
4.5	3	4.00	3.75	3.75	3.50
5.5		4.00	4.00	4.00	3.75
6.5		4.50	4.50	4.25	4.00
3.5	4	4.25	4.00	3.50	3.50
4.5		4.00	4.00	3.75	3.50
5.5		4.25	4.25	4.00	4.00

RUN# 2-6-2

Z		45	35	25	15
Q	q _u	d _s			
4.5	3	4.00	3.75	3.50	3.25
5.5		4.00	4.00	3.75	3.50
6.5		4.50	4.25	4.00	3.75

RUN# 2-6-3

Z		45	35	25	15
Q	q ₁	d _s			
4.5	3	4.25	4.00	3.75	3.50
5.5		4.25	4.00	4.00	3.75
6.5		4.50	4.25	4.00	4.00

RUN# 3-6-2

Z		60	50	40	30	20
Q	q _u	d _s				
6.0	2	4.50	4.25	4.00	3.75	3.75
7.0		4.50	4.25	4.00	4.00	4.00
8.0		4.50	4.50	4.25	4.25	4.25
5.0	3	4.50	4.25	4.00	3.75	3.50
6.0		4.25	4.25	4.00	4.00	4.00
7.0		4.50	4.50	4.50	4.25	4.25
4.5	4	4.75	4.50	4.00	3.75	3.50
5.0		4.50	4.50	4.00	4.00	3.75
6.0		4.75	4.75	4.25	4.00	4.00

RUN# 3-6-3

Z		60	50	40	30	20
Q	q ₁	d _s				
6.0	2	4.25	4.25	4.00	3.75	3.75
7.0		4.25	4.00	4.00	4.00	4.00
8.0		4.50	4.50	4.25	4.25	4.25
5.0	3	4.25	4.00	3.75	3.75	3.75
6.0		4.25	4.25	4.00	4.00	4.00
7.0		4.50	4.50	4.75	4.25	4.25
4.5	4	4.50	4.25	4.00	3.75	3.75
5.0		4.50	4.50	4.25	4.00	4.00
6.0		4.75	4.50	4.50	4.25	4.00

RUN# 3-6-1

Z		60	50	40	30	20
Q	q _a	d _s				
7.0	0	4.00	4.00	3.75	3.50	3.50
8.0		4.00	4.00	4.00	4.00	4.00
9.0		4.50	4.50	4.25	4.25	4.25
5.5	2	4.50	4.00	3.75	3.75	3.50
6.5		4.25	4.25	4.00	4.00	4.00
7.5		4.50	4.50	4.25	4.25	4.25
4.5	3	4.25	4.00	3.75	3.50	3.50
5.5		4.25	4.25	4.00	3.75	3.75
6.5		4.50	4.50	4.25	4.25	4.25
4.0	4	4.50	4.25	4.00	3.75	3.50
4.5		4.50	4.25	4.00	3.25	3.75
5.5		4.50	4.25	4.00	4.00	4.00