RESIDENCE TIME OF SOLIDS IN CONTINUOUS FLUIDIZED BEDS

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By
DINESH KUMAR BHARADWAJ

DEPARTMENT OF CHEMICAL ENGINEERING
UNIVERSITY OF ROORKEE
ROORKEE [INDIA]
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CERTIFICATE

It is certified that the thesis entitled "Residence Time of Solids in Continuous Fluidized Beds" which is being submitted by Dinesh Kumar Bharadwaj in fulfillment of the requirements for the award of the degree of DOCTOR OF PHILOSOPHY in Chemical Engineering at the University of Roorkee, Roorkee is a record of the candidate's own work carried out by him under the supervision and guidance of the undersigned. The matter embodied in this thesis has not been submitted for the award of any other degree.

This is further certified that the candidate has worked for a period of five years since May 1970 for preparing this thesis.

S. K. Saraf

S. K. Saraf
Department of Chemical Engineering
University of Roorkee
Roorkee, U.P. 247667

AK Jagota

A. K. Jagota
School of Chemical Engineering
Oklahoma State University
Stillwater, Oklahoma 74074
U.S.A.

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NOMENCLATURE

b, c, d	Stoichiometric Coefficients
C	Dimensionless output tracer concentration
d_p	Particle diameter, mm or in.
E(t)	Exit age distribution function using ordinary time, $sec.^{-1}$
G	Gas flow rate, l/min.
G_{mf}	Gas flow rate at minimum fluidizing conditions, l/min or lb/(hr) (sq.ft)
h	Height, cm
h_D	Static bed height to bed diameter ratio
H	Hold up ratio
H(2, 1)	Hold up ratio between middle and small size particles
H(3, 2)	Hold up ratio between large and middle size particles
$p_o(d_p)$	Size distribution of feed solids
t	Time, sec. or min.
$\bar{t}$	Mean residence time of solids, sec or min
T_{GA}	$= V_f/G$, gas contact time in fluidized bed, sec or min
T_{GB}	$= V_s/G$, gas space time, sec or min
u	Velocity, cm/sec
u_{mf}	Superficial gas velocity at minimum fluidizing conditions, cm/sec
V_b	Bed volume, cm^3
V_f	Fluidized bed volume, cm^3
V_s	Volume of solids in the bed, cm^3
w	Solid feed rate, g/min
W	Weight of Solids, g
x	Mass fraction

X_B	Fractional conversion of solid B
$\bar{X}_B$	Mean conversion of exit stream of solids

GREEK SYMBOLS

ϵ_{mf}	Void fraction in a bed at minimum fluidizing conditions
θ	Dimensionless residence time
μ	Viscosity of gas, cp or lb/(hr) (ft)
ρ_f	Fluid density, g/cc or lb/cu ft
ρ_g	Gas density
ρ_s	Solid density, g/cc or lb/cu ft
τ	Mean residence time,sec or min
τ_A	$= \tau_s / T_{GA}$, dimensionless time parameter
τ_B	$= \tau_s / T_{GB}$, dimensionless time parameter
τ_s	= W/w, mean residence time of particles, sec or min

Subscripts

1, 2, 3	for small, middle and large size particles respectively.

SYNOPSIS

A thesis entitled " Residence Time of Solids in Continuous Fluidized Beds" submitted in fulfilment of the requirements for the degree of Doctor of Philosophy in Chemical Engineering by Dinesh Kumar Bharadwaj at the University of Roorkee, Roorkee in September, 1975.

Industrial fluidized beds normally use wide size distributions of solid feeds in continuous operations. In literature, practically no work has been reported about the mean residence time of different size particles in moving and fluidized beds when mixed feed containing particles of wide size distributions are used. Fluidized beds at present are designed on the assumption of equal mean residence time for each size particles. The present work investigates the residence time distributions of particles of different sizes using mixed feeds in moving and fluidized beds. The effect of air flow rate, solid feed rate, bed height and particle size distribution of solids was analysed and the findings are of great importance for the proper design of fluidized bed systems in which mixed feeds are used.

A 50 mm inside diameter column of perspex and glass ballotini beads of 0.393 mm, 0.724 mm and 0.96 mm diameter were used in the investigation. The solid feed in different size combination was introduced at the top of the bed and the solids were removed by a discharge system located mostly at the bottom of the bed, thus giving, in general, a downward flow of solids. Air was used as the fluidizing gas and studies were made on the residence times of solids in moving and fluidized beds by using stimulus-response technique with fine coating of iron particles

on glass ballotini beads as tracer.

The study of moving beds was carried out with air flow rates in the range of 0.2 u_{mf} to 0.95 u_{mf} using a double pulse tracer injection technique to eliminate the end effects. The effect of particle size was studied by introducing the tracers of different sizes in the feeds of mixed sizes. The parameters studied were: solid feed rates from 34 to 300 g/min giving a mean residence time of particles in the range of 1.1 min to 30 min and the bed height to diameter ratios were varied from 2.0 to 6.0.

The studies on fluidized beds were carried out by single pulse tracer injection technique and the air flow rates were varied from 1.2 to 6.0 u_{mf}. The effect of particle size was studied by introducing tracer particle of different sizes in the feeds of mixed sizes. The parameters studied were: solid feed rate from 34 to 300 g/min, bed hold ups corresponding to static bed height to diameter ratios from 0.5 to 4.0 and the mean residence time was varied from 1.1 to 10. min.

The tracer studies revealed interesting and useful observations and further investigation was carried out on the mixed feeds in fluidized beds by analysing the bed hold ups. The parameters investigated were air flow rate from 1.3 to 8.0 u_{mf}, solid feed rate from 120 to 500 g/min, bed hold ups from 90 to 620 g of solids corresponding to static bed height to diameter ratios from 0.5 to 4.0.

A model is proposed for fluidized beds to explain the observed behaviour at different operating conditions. Axial and radial distribution of solids in the bed were measured and modifications in the solid discharge system were made to confirm the basic postulates of the proposed model.

The effect of horizontal and vertical baffles was also studied and the changes in the hold up ratios are explained with the help of the proposed model.

For the system studied, experimental results confirm that the flow of solids in moving beds closely resembles plug flow upto u/u_{mf} of 0.95. End effects due to the solid discharge system are primarily responsible for any dispersion observed in the stimulus response experiments. There is practically no effect of bed height to diameter ratios or of solid feed rate on the flow pattern of solids in moving beds. In the experimental range investigated a dead zone of particles was observed in moving bed close to the solid discharge system. The dead zone remained virtually uninfluenced by variables like particle size and bed height but an increase in solid feed decreased the volume of the dead zone. In moving beds with continuous solid feeds of mixed sizes of particles, the flow behaviour of each size particles exhibited plug flow quite closely upto u/u_{mf} value of 0.95.

In fluidized beds with feeds of uniform sizes, experimental results confirm that the solid flow pattern resembles ideal backmix for air velocity higher than 1.5 u_{mf}. For air velocities from 1.2 to 1.5 u_{mf}, the presence of short circuiting, dead zone and plug flow was detected along with the major backmix flow. In fluidized beds using mixed size feeds the particles of each size exhibited different mean residence times and yet individually they were found to be perfectly backmixed. The mean residence time of large size particles was found to be more than that of smaller particles. This effect is quite useful for industrial applications of fluidized beds for continuous feeds containing mixed size particles. The hold up ratios, defined as the ratio of the mean residence times of any two size particles increased with air velocity upto about 2.5 u_{mf}, and further increase in air velocity resulted in a sharp decrease in the value of hold up ratios. The hold up ratios are found to be independent of the solid feed composition provided u/u_{mf} is used as the correlating parameter. A second order polynomial of the form:

$$\frac{H}{(\tau_B)^n} = b + c \ (u/u_{mf}) + d \ (u/u_{mf})^2$$

satisfactorily correlates hold up ratio H, dimensionless time parameter τ_B representing solid feed rate and u/u_{mf} for superficial gas velocities higher than 2.5 u_{mf}.

The trend of the variation of hold up ratio remains the same for solid feeds of two or three size combinations. Very sharp cut combination of sizes, however, produced poor fluidization and hold up ratio values less than unity were observed and lower hold up ratio values were found in comparison to mixed feeds of two particle sizes only.

The effect of different mean residence time for different size particles in fluidized beds is explained by proposing a model which postulates that the low velocity of air near the enclosing wall results in the formation of two thin boundary layers. The boundary layer close to the enclosing wall was richer in small size particles and the second layer next to the first boundary layer was slightly richer in large size particles. The thickness of two boundary layers was affected by the formation of large bubbles and by slugging in the bed. With the help of the model, the effect of various parameters on the hold up ratios is explained.

The vertical baffles produce smooth fluidization and the radial gradients of the particles of different sizes were not significantly influenced by the presence of vertical baffles. The horizontal baffles with smaller openings and close spacings resulted in the segregation of larger particles at the bottom and hold up ratios less than unity were obtained.

CHAPTER I

INTRODUCTION

Reactions involving chemical transformation of solid particles in an atmosphere of gas have numerous industrial applications. Examples of such reactions include roasting of sulphide ores, reduction of iron ore, nitrogenation of calcium carbide, calcination of limestone and chlorination of uranium oxide. For all these reactions, fluidized bed reactors have exhibited an overall improvement in performance as compared to fixed bed reactors. Continuous operation of solid phase in fluidized bed is an added advantage since the handling of solids becomes relatively easy.

The use of fludized beds for continuous chemical processing of solid materials can be successful only if adequate information is available about the nature of solids movement in and through the bed in order to predict residence times of solids accurately. A knowledge of solids movement within fluidized beds is also of interest to get a better understanding of flow pattern in a fluidized bed. The operation of fluidized bed as a continuous reactor in which solid feed material is introduced at the top and reacted material is withdrawn at the base requires a knowledge of the residence time distribution of the solids in the bed before kinetic calculations can be made. Fluidized beds with continuous flow of solids flowing counter-current to the gas have been used as fluo Solid system for sizing and drying of dolomite particles. For such beds also the residence time distribution of particles is of great importance.

A considerable amount of literature has appeared since 1960 on the flow behaviour of solids in fluidized beds but not much work has been reported on the solid flow behaviour in moving beds. In the work reported so far on solids mixing very little emphasis has been paid on the effect of the presence of particles of more than one size in the feed. Such feeds are always used in industrial processes and a better understanding of the flow pattern of particles of different sizes is of paramount importance. The lack of proper knowledge of the flow pattern of particles of different sizes when solid feeds of wide size distribution are used places an undue emphasis on the previous plant practices rather than on the basic principles while designing such reactors for industrial application. Practically no investigation has been reported so far on the flow pattern of particles of different sizes in a fluidized bed system using solid feeds with wide size distribution.

The proposed work was undertaken to investigate the residence time distributions of particles of different sizes using mixed feeds in moving and fluidized beds. The effect of air flow rate, solid feed rate, bed height and particle size distribution of solids has been analyzed in this work. The findings of the present investigation will have great practical value for the proper design of fluidized bed system in which mixed feeds are used.

CHAPTER 2

REVIEW OF LITERATURE AND EARLIER INVESTIGATIONS

2.1 GENERAL

Before 1942, study on fluidization was of theoretical interest only, but in 1942, fluidized beds entered the industrial scene in a big way with the advent of catalytic cracking and have since found use in many other areas. The use of fluidized beds in industry showed sometimes proud success but it also resulted in many important failures. This was mainly due to the lack of some unifying theory to explain adequately the characteristics of a fluidized bed. The first attempt to present the subject in a unifying manner was made by Leva(59) in 1959. In 1960's, the research in the field witnessed a rapid growth and the literature on many good books have appeared on the subject (24, 56, 81, 90, 106, 117) and they have contributed to a great deal in understanding the behaviour fundamentals of fluidization and the behaviour of fluidized beds. The presence of bubbles in the fluidized bed was conceived by many earlier investigators, but Davidson(23) made a significant breakthrough by presenting a bubbling bed model to account for the movement of both gas and solid and the pressure distribution about the rising bubbles.

In a recent book, Davidson and Harrison (24) have consolidated the results of most of the important investigations on fluidization and fluidized bed behaviour. The wealth of information collected by countless workers in this field has little utility unless it is presented in a form suitable for the design of industrial fluidized bed systems. In a recent book, Kunii and Levenspiel (56) have made an excellent attempt to present the design of fluidized bed systems using the fundamental principles and the bubbling bed model, a modified extension (54) of the Davidson model.

2.2 INDUSTRIAL APPLICATIONS OF FLUIDIZED BEDS

Fluidized beds have numerous applications in chemical industry both in the fields of physical operations and chemical reactions. The fluidity of solids in fluidized beds helps in the transportation of solids from one place to another. Similarly the mixing of solids is easily achieved in fluidized conditions due to the random motion of particles. Both in physical operations and chemical reactions, fluidized beds have been extensively used for heat exchange because of their unique ability to rapidly transport heat due to random motion of particles and to maintain a uniform temperature throughout the bed. The random motion of particles in fluidized condition has also been used in plastic coating of metal surfaces. Drying and sizing of powdery materials have been successfully achieved in fluidized beds. Crystal growth is also found to improve in fluidized beds and they find extensive use in crystallization. Dehumidification of air using adsorbends has also been practiced in fluidized beds.

In the field of chemical processes, fluidized bed reactors are now being used very widely for carrying out synthesis reactions, cracking and reforming of hydrocarbons, carbonization and gasification of coal, calcinating and clinkering and for many other gas-solid reactions. Details of such applications are discussed by Kunii and Levenspiel(56).

2.3 BUBBLING BED MODEL

General characteristics of fluidized beds are discussed in various books but a reasonable model to represent the flow of gas through the bed and its mechanism of contacting the solids was explained clearly by the bubbling bed model of Kunii and Levenspiel(54). Without mentioning the details of this model, it may only be pointed out that the fluidized beds have two regions, one of low solid density and the other of higher solid density. The former region is more commonly known as bubble phase and the second region of high solid density is termed as emulsion phase. Most of the gas flows through the bubble phase and most of the solids flow through emulsion phase. In the Davidson model gas bubbles were considered solid free, spherical in shape but later developments by various investigators have revealed that the bubbles have a concave shape at the bottom and this small concave portion of bubble, known as 'Wake', carries solids with it The work of Kunii and Levenspiel(54) has explained clearly the gas behaviour in the vicinity of bubbles, bubble growth, bubble velocity, size of bubbles and voidage in bubble and emulsion phases.

2.4 SOLIDS MIXING IN MOVING AND FLUIDIZED BEDS

The rapid mixing of two layers of differently coloured solids was among the earliest observation on fluidized beds and the rapid solids movement was used to explain the much better heat transfer properties of fluidized bed as compared to a fixed bed. Several techniques are used to study this rapid movement of solids. Experimental findings with such techniques are summarized in the following sections.

2.4.1 Study of Particle Trajectories

The study of solids mixing can be carried out quite easily by following the trajectories of individual particles in the fluidized bed, Massimilla and Westwater (69) and Toomy and Johnstone(102) have used high speed cinematography to study particle trajectories. In the work of Massimilla and Westwater motion pictures at 2000 frames/sec. were used to measure the movements of individual solid particles and gas bubbles in a fluidized bed. Air was used to fluidize 0.028 inch glass spheres and 200 mesh alumina in a 3.75 inch diameter glass column. The results obtained by them revealed that aggregates were very common and each aggregate moved as a unit. Particles and aggregates near the wall showed pronounced alterations of fast and slow movements both upward and downward. Individual particles also exhibited spin. The use of baffles increased the bed density and decreased the particle velocity. Kondukov et. al (51) studied trajectories of tagged radio-active particles in a 7 inch diameter glass column using air to fluidize catalyst particles of 2.81 mm size. Their results indicated that the particles move randomly in the bed, the upward motion of particles being more rapid than downward movement. They also observed that particles near the surface usually remain there for a while before dropping into the bed. Observations by these workers also showed a lateral pulsating movement of particles during downward flow which was less pronounced during upward flow. Local zones of recirculating solids were observed just above the distributor for non-uniform distribution of air. Rowe (84) conducted experiments by allowing a bubble to pass through a fluidized bed of lead glass and soda glass particles and studying the flow patterns by passing an X-ray beam through the bed and receiving it on a phosphor screen. The results obtained by Rowe showed that the particles followed a definite pattern of displacement caused by rising bubbles rather than following a completely randon motion as is assumed normally to be true for mixing operations. Gas fluidized particles of more than 100 micron size were shown to be displacable by bubbles in a predictable way. The behaviour of very fine particles was shown to be relatively more random mainly due to the effect of surface area. Kislykh and Chirkov (48) studied the distribution of probability of fluctuations in a number of solid phase particles over a height of fluidized bed. A normal distribution of the fluctuations probability was recorded in the middle of the bed and a J type Pearson distribution was recorded near the grid and near the upper bed boundary, confirming that the fluidized bed was composed of three well defined zones.

2.4.2 Models Based on Circulation Patterns

The description of the mixing process can also be given by observing the circulation rate of solids. In such experiments, intermixing is studied between two sections of the bed separated by a horizontal plane. The change of tracer concentrations in the two sections is correlated with flux of solids across the boundary. Leva and Grummer (69) have conducted such experiments using 3.36 cm and 5.16 cm diameter beds. Air was used to fluidize sand and silica gel as the solid material. Talmor and Benenati (98) have also conducted similar experiments in beds of 3.36 to 10.2 cm diameter fluidized by air with ion exchange resin as the solid feed. Katz and Zenz (121) have given a mathematical model to calculate the internal circulation rate. Lateral circulation rates were determined by Lochiel and Sutherland (66). More detailed and more realistic models have been proposed by Bailie (9) and Van Deemter (25). Certain fractions of particles move upward or downwards in different parts of the bed and these change place with time. There exists a cross flow between these sections and the surface area available for cross flow remains constant. The determination of circulation rates was made on the assumption of constant velocity of the particles and complete mixing within each horizontal region. However, due to the heterogeneity of the bed with respect to the movement of particles and questionable fit of the model to the real situations, Kunii and Levenspiel (56) have indicated that the reported values of the circulation rate parameter should only be considered as a first approximation of the flux of solids up or down a bed.

2.4.3 Models based on Residence Time Distribution (RTD) of solids

Danckwerts (22) has described methods of investigating and specifying the behaviour of continuous flow systems. The nature of the mixing process could be studied by studying the response at the exit of the bed to a tracer signal at the entrance. This method leads to a fairly simple experimental technique, and to a possibility of describing the experimental results mathematically. The properties of the tracer material should be the same as that of the inflowing material, yet, some other property, not important to the flow qualities of the parent material should be different in order to analyze the tracer concentration at the outlet stream. Many different techniques for such stimulus-response studies have been used to study the flow of solids in continuous systems. In view of the use of tracer technique in the present investigation, the techniques used by earlier investigators are described briefly.

2.4.3.1 Tracer Techniques used in RTD Studies

Morris et. al (78) in their study on moving and fluidized beds conducted experiments with glass ballotini beads (Spherical glass particles) as the solid material and the tracer material consisted of the ballotini beads with a thin coating of blue dye. Analysis of samples of the solids was effected by shining a beam of light onto a sample of particles and measuring the intensity of the reflected beam by means of a cadmium sulphide photoconductive cell. Since the cell was insensitive to blue light, the intensity of the reflected beam decreased with increase in the concentration of particles with blue dye in the sample. Wolf and Resnick (110) used the tracer of magnetic sand, actually magnetite coated with a polymer material and the analysis was carried out by a magnetic separator. Similar tracers were prepared by Heertjes et. al. (42) by impregnating a portion of the silica gel stock with colloidal $Fe_3 O_4$ and analysing these tracers by strong magnetic fields (20,000 gauss) using a vibrating apparatus. An accuracy of $\pm$ 0.3% was reported for these experiments. Tailby and Cocquerel (199) also used glass ballotini beads as the solid material and tracer material was prepared from these glass ballotini by coating them with blue dye with the help of perspex granules dissolved in chloroform. Analysis was carried out by washing the sample with a known weight of chloroform and comparing the colour of the liquid with solutions of known concentration in the spectrophotometer. Sutherland (96) observed that the use of coloured glass ballotini in a bed of uncoloured material, the colour being soluble in acetone, was very tedious, especially as the analytical technique involved dissection of the entire bed into many zones of equal volume and, therefore, used two other techniques for stimulus-response studies. In one technique he used nickel shots as tracer in a bed of copper shots. The two metals have practically identical physical properties relevant for flow.

Since nickel is magnetic, easy separation of the tracer from main feed is possible both for the analysis and for the removal of all tracer material from the bed. In the second technique, tracer was prepared by irradiating some of the bed material and analysing the tracer concentration by determining the radioactivity. Yagi and Kunii (112) used coke as the tracer material in the main feed of sand particles. Since these experiments were conducted at high velocities, the fluidizing properties of coal and sand were considered same. The analysis of coke was done by burning the samples and determining the loss in weight. Kamiya (47) used particles coated with potassium dichromate as the tracer and the amount of tracer in the exit stream was measured by dissolving it in water and analysing for colour intensity. Hayakawa et. al (41) used conductive tracers in a bed of non conducting materials and observed the change in electrical resistance.

2.4.3.2 Solids Mixing from RTD

Among the earliest studies on solids mixing by RTD were made by Tailby and Cocquerel (100) on the residence time distribution of solids in a system consisting of glass beads fluidized with air. The variables studied were aspect ratio (length to diameter ratio of the fluidized bed), the gas flow rate and the solid feed rate through the bed. They studied both the co-current and the counter-current flow of gas and solids and observed that the increase in aspect ratio as well as in solid feed rate increases the tendency to plug flow. Increase of air rate was found to be less significant and showed only a slight tendency towards perfect mixing. Their work was mainly of a qualitative nature. Sutherland (96) compared the performance of tapered and non-tapered beds, both circular and annular in cross sections and confirmed the reduction in vertical mixing rates of solids in deep fluidized beds of dense materials by the taper of the retaining walls. They also found that the effect of tapering was confined to gas flow rates less than 30% above the minimum fluidizing value and to beds of over 2 feet in height. The inclusion of a non tapered insert in the bed did not reduce the vertical mixing rates. In his further study Sutherland (83) found that solid mixing was caused mainly by bubbles only. Bubbles appeared at 1.3 u_{mf} and for deep beds mixing was found to occur at the top portion of the bed for air rates as low as 0.83 u_{mf} due to the velocity increase at the bed top because of gas expansion. He also found that for very fine particles where u_{mf} was much lower than the bubble velocity, slow mixing occured even at velocities higher than 1.3 u_{mf} and determined that for good particle movement, depths of the beds should be at least ten times the bubble diameters and diameter of bed should be at least more than two times bubble diameter.

Yagi and Kunii (112) extended the study of RTD to carry-overs and overflow systems in fluidized bed and concluded that beyond 1.3 u_{mf} close agreement between theory and experiment justified the assumption of complete mixing of particles in the fluidized bed. These results were also corroborated by Kamiya (47). It was further shown by Yagi and Kunii that the average residence times for a given particle size was the same in both the carry over and the over flow streams. Another important study on the RTD of solids in moving and fluidized beds was made by Morris et. al. (78). At high gas flow rates of more than 1.5 u_{mf} for the system studied by Morris et. al., the fluidized bed could be treated as a perfectly mixed system with some solids by passing. Solids by passing was found more significant at small bed heights and it increased with the decrease in solid flow rates. A maximum value for solids by-passing of 11% was observed at a bed height of 9 inch and a bulk solids flow rate of 0.75 ft^3/hr. At low gas flow rates, below 1.3 u_{mf} the solids in the bed tended progressively toward piston flow. No significant effect of solids flow rate on the residence time distribution of solids was detected over experimental range examined. It was also shown that the solids mixing in moving and fluidized beds could not be described by a simple diffusional mechanism and it was concluded by them that solids diffusivities reported in literature have little meaning when applied to moving and fluidized beds.

The transport and the residence time of particles in a continuously fed rectangular fludized bed with low bed heights compared to reactangular dimension of the bed using silica gel particles was examined by Heertjes et. al. (42). For a low depth rectangular fluidized bed with a net flow of solids stream, the transport of the particles was described by a model employing perfect mixing and piston flow in a binomial distribution function. Alfke et. al. deduced from their experiments that the mixing of the solids could not be described by a single mixing coefficient, since the bed consisted of two distinct layers, a top layer and the rest of the bed. The total bed transport was described by a constant quasi-diffusion coefficient and the extra convective transport in the top layer was also described by a constant quasi diffusion coefficient.

Solids mixing by use of RTD data was also reported by Bowling and Watts (16), Turyayev and Tzailingold (103). Wolf and Resnick (111) have extended the study of RTD of solids to multicompartment fluidized beds as a function of solid and gas flow rates. The use of baffles was also studied by them.

2.4.3.3 Mathematical Representation of the Proposed Models

The various models proposed for describing the solids mixing in moving and fluidized beds were mathematically represented by the following methods:

2.4.3.3.1 Holdback and Segregation

The concept of holdback and segregation was introduced by Danckwerts to describe the deviation from piston flow and perfect mixing. Other workers (16, 78, 100) have also used this representation to describe the mixing behaviour. Such concepts are useful only for qualitative representation of the mixing process. Equal values of holdback and segregation may represent widely different conditions of solids flow pattern and these parameters, moreover, do not provide information in readily usable form for design calculations. Such models are, therefore, not very useful.

2.4.3.3.2 Partial Perfect Mixing

Cholette and Cloutier (29) have described models based on perfect mixing in a part of bed combined with piston flow, short circuiting, dead space etc. for the rest of the bed. For fluidized1 beds many other investigators (6, 27, 45, 109) have found that the mixing of solids could be described by perfect mixing conditions. Wolf and Resnick (110$, however, pointed out that basic knowledge of the mixing process could not be obtained via this approach.

2.4.3.3.3 Perfect Mixers in Series and Parallel with backmixing

The use of these models has been described by Zaloudik (118) and Moo Young and Chan (76). Shinnar et. al (89) have given equations to represent such models as a result of the method of equal variances. However, solids mixing has not been reported in literature in the form of such models because experimental curves do not match very well with the theoretical curves.

2.4.3.3.4 Diffusional Model for Solid Movement

Many investigators adopted the idea of random movement of particles or eddies. For the description a diffusional coefficient or mixing coefficient was used. The differential equation using diffusion coefficient and its solution were first given by Danckwerts (22) and the equations using mixing coefficient have been described by Houghton (38). Van Deemter (25) stated that in his circulation model for long mixing times the solids, mixing can be described by an effective diffusion coefficient. However, Morris et. al. (78) and Tailby and Cocquerel (100) found that the diffusional model could not be applied to their experimental results. May (71) used beds of 10 m height and 1.5m diameter and reported the values of axial diffusivities of solids. The little deviation from the diffusional model observed by him were attributed to circulatory movement of solids. Jinescu et. al (46) found that the mixing coefficient varied with the position in the bed with time.

The influence of gas velocity on mixing coefficient was reported by various workers (17, 26, 46, 65, 68, 77) but Alfke et. al (2) observed that the changes in gas velocity had no effect on mixing coefficients. The variation of mixing coefficient proportionally with some power of diameter was reported by de Groot (26) and May (71). Mori and Nakamura (77), Alfke et. al. (2) and Morris et. al (78) estimated the effect of bed height on diffusion coefficient. Brötz (17) showed that the mixing coefficient are inversely proportional to the particle size. Mori and Nakamura (77) have correlated mixing coefficient to bubble diameter and Alfke et. al (2) correlated it to solid feed rate.

The basic assumptions in the diffusion model are that the system is homogeneous and that the movement of particles is perfectly random. The physical behaviour of the fluidized bed, however, shows that these assumptions do not hold. The bed is not homogeneous as shown by porosity measurements (10, 21, 104) and particle movement studies (15, 52). Rowe et. al (87) also observed that for particles larher than 100 micron the movement is not random but a repetition of a distinct pattern. For particles below 100 micron the movement is caused not only by bubbles but by eddy diffusion as well. However, in uniformly bubbling beds the possibility exists that the repetition of the movement is so fast that the random movement is approximated.

2.4.5 Solids Mixing through Axial Temperature Distribution

An indirect method to find the axial diffusion coefficient of solids was used by various investigators (55, 59, 91). The axial temperature distribution was measured in long narrow fluidized beds, with lower section heated and upper section cooled and vice-versa. By neglecting the heat capacity of the flowing air the effective axial thermal conductivity were determined and with the assumption that the heat transport was caused by the movement of solids alone, axial diffusion coefficients for solids were determined.

2.4.6 Lateral Mixing of Solids

To evaluate the radial diffusion coefficients, lateral mixing of solids was studied by Brötz (17), Gabor (31) and Mori and Nakamura (77). Rectangular beds were separated by a vertical partition plate and both sections of the bed were fluidized by air. With a sudden removal of the partition plate, the rate of approach to uniformity of solids was measured and the radial diffusion coefficients correlated. The results obtained show an increase of radial diffusion coefficient with gas velocity. The radial diffusion coefficient were found to be much smaller in magnitude than the axial diffusion coefficient. But rather wide spread in the reported data suggests that more work need to be done to prove the applicability of diffusional model to the flow of solids in fluidized beds.

2.4.7. Mixing Caused by Bubbles

From the findings of Sutherland (96) and Rowe and Sutherland (83) mixing of solids could be related to the mixing caused by rising gas bubbles. These findings suggested that the rate of circulation of solids could be estimated from the number and size of bubbles passing through the bed. Gabor (30) measured the movement of particles around a rising bubble in a transparent two dimensional bed fluidized slightly above the incipient condition. and concluded from his results that the mixing of solids in the wake of a rising bubble was fairly good. Kunii et. al (55) measured radial dispersion coefficients of solids and they found that this was dependent on the effective size of bubbles in the bed.

Rowe et. al (87) and Toei et. al (93) have described in detail the various techniques for following particle motion in a gas fluidized bed. The mixing of solids described by most as a flux of particles across a horizontal plane in the bed and the extent of this mixing is expressed in terms of the excess gas flow rate over the minimum fluidizing gas flow rate and the wake fraction and mean particle size. Many investigators pointed out that the bubbles induce groups of particles to move upwards in the bubble wake and downwards everywhere else in the bed. Baeyens and Geldart (7) have discussed quantitatively, the particle motion around a rising bubble. It was found that the velocity of the wake particles was equal to the rise velocity of the bubble and an average value for the velocity of particles carried upwards by drift action amounted to 0.38 times the bubble velocity. The wake and drift fraction decreased with increasing particle size. De Groot (26) have described the scale-up rules for gas fluidized bed reactors from the theory of solids mixing caused by bubbling in the bed.

2.4.8. <u>Variation in Particle Size Distribution</u>

Particle size distribution in fluidized beds have been studied as a function of bed height by Urabe et. al (104). A steady state technique using sand of wide size distribution at high gas velocities was used in a tube 14 cm in diameter. Solids were fed continuously to the bed, fines were elutriated with the gas and an over flow tube discharged solids keeping the bed height constant. It was concluded by them that the size distribution was roughly constant within the main zone of constant voidage, however, the upper falling density zone became progressively richer in fines. In addition, the main zone had less fines at high velocities, indicating that fines were more rapidly eliminated at higher velocities. The conclusions were corroborated by Thomas et. al (99) who used 10.2 cm diameter bed and by Hiraki et. al (44) who used microspherical catalysts in rectangular beds. The results of Hiraki et. al, showed homogeneity of size distribution in the main region of the bed and falling density zone at any bed height.

At the time of writing this thesis, the author has found one short paper by Chechetkin et. al (19) describing the effect of the solid phase particles on the intensity of their mixing in the fluidization of a polydisperse material. It was established by them that in a bed of polydisperse material, there was a separation of particles of the solid phase with respect to size through the height of the bed for velocities ranging in between 1.1 to 1.3 u_{mf}. The experimental results for air velocities higher than 1.3 u_{mf} showed that the average residence time of the coarse particles was higher than that of the fine particles. Beyond 2.4 u_{mf} the intensity of vertical mixing increased giving less separation of the two particle sizes. However, details of experimental set up and the technique used are not described by them. Baskakov et. al (11) conducted an experimental study on the effect of fluidizing air velocity on this quality of separation of solid materials in a continuously operating laboratory apparatus with two fluidized beds one above the other with and without the incorporation of

a third ......

solid material as a packing. It was shown by these investigators that the appropriate ratio for the cross sectional areas of the two zones, proper discharge organization and with proper air velocity in both zones, practically complete separation of the products could be achieved.

2.4.9 Miscellaneous Studies on Solids Mixing

Many investigators have studied the fluidity of the solid particles in fluidized beds in terms of viscosity of the bed. The viscosity was defined and measured in a similar manner as for a homogeneous fluid. In a recent paper by Bessant and Botteril (13) non-newtonian properties of a 200 micron sand fluidized by air are reported. On this basis they determined the velocity as a function of channel aspect ratios and slip across the distributor.

Regular circulation of the particulate phase was induced in a shallow fludized bed by means of an uneven distribution of fluidizing fluid by Merry and Davidson (72). Geldart (32, 33) recently showed that differences of opinions regarding the fluidizing behaviour of powders could be explained by postulating that there were four types of powders. Data on the fluidization of one particular powder could then be generalized to other powders within its group and he proposed tentative criteria to define the boundaries of each group. The ideas have been quantified by Baeyens and Geldart (8).

2.5 BAFFLES IN FLUIDIZED BEDS

Internal baffles have a wide application in fluidized bed systems as heat transfer surfaces as means of promoting smooth fluidization and as aids to successful scale up. Various types of baffles have been used by various workers and in a recent paper, Grace and Harrison (37) have compared the use of different types of baffles on the design of fluidized beds and they recommended the use of vertical rods over fixed Backings or other baffles of more complex geometry for achieving smooth fluidization.

CHAPTER 3

DEVELOPMENT OF THESIS OBJECTIVES

The chemical reactions of solid particles reacting with a fluidizing gas can be represented by

$$A\ (\text{gas}) + b\ B\ (\text{Solid}) \rightarrow \begin{cases} \text{gaseous products} \\ \text{Solid product} \\ \text{gaseous and solid product} \end{cases}$$

In operations with continuous feed of solids the average conversion depends on two factors:

(i) The rate of reaction of single particles in the reactor environment

(ii) The residence time distribution (RTD) of solids in the reactor.

Generally two types of kinetic models have been used to calculate conversion of solids in the reactor. In the continuous reaction model diffusion of gaseous reactent is considered rapid enough in comparison to the chemical reaction, while in the shrinking core model the diffusion of gaseous reactent is considered so slow that the reaction zone is restricted to a thin front which advances from the outer surface into the particle. Both these models represent closely many of the real systems and design of a reactor for gas-solid reactions based upon these models are adequate for such systems.

In the shrinking core model, the reaction front advances from the outer surface into the solid leaving behind a layer of completely converted and inert solids called the ash or product layer. For gas A to react with solid B, it has to first diffuse through surrounding gas film of the particle and then diffuse through the blanket of product layer to reach the reaction front.

3.1 CONVERSION OF SOLIDS OF UNIFORM SIZES

When one or other of the above mentioned resistances control the overall reaction rate, kinetic expressions have been derived (56) between conversion X_B and t/τ, where τ is the total time of conversion of the particle. The total time of conversion for any particle depends among other things on the particle diameter. For diffusion controlling step τ is proportional to the square of particle size, whereas, it is directly proportional to particle size for reaction controlling step. In either case, larger diameter particles require more time for total reaction as compared to the smaller diameter particles.

For design calculations of a reactor for gas-solid reactions, equation 3.1 should normally be used to calculate average conversion from the bed for feeds of single size particles (54, 56)

$$1 - \bar{X}_B = \int_0^\infty (1 - X_B)_{\text{particle}} \; E(t).dt. \qquad \ldots \text{(eq. 3.1)}$$

In the moving bed reactors plug flow is normally assumed in which case $\bar{X}_B$ is same as X_B while in the fluidized bed reactors, complete backmixing of solid particles is assumed so equations are derived between $\bar{X}_B$ and X_B by substituting $\frac{1}{\bar{t}}$ exp $(-t/\bar{t})$ for E(t)

For a given particle size and material, τ is determined experimentally and used in the available equations (54, 56) to calculate X_B of particles for a given reaction time t. Using the assumptions of plug flow and backmix flow respectively for the moving and fluidized bed reactors, design calculations can easily be made.

The validity of the assumption of plug flow in moving beds has been discussed earlier in Section 2.4.3.2. The findings of Morris et. al. (78) have reported that for gas velocities close to u_{mf} values, the RTD shows some deviation from plug flow and it was attributed to variation in solid velocity across the diameter of the column. An attempt to describe the velocity profile mathematically and to use the derived relation to characterize the experimental data was abandoned by them due to an observed change in velocity profile with change in bed depth. They concluded that this was largely due to the influence of the offtake system employed by them.

The assumption of backmix flow in fluidized bed has enough evidence in literature as discussed earlier in Section 2.4.3.2.

3.2 <u>CONVERSION OF SOLIDS OF NONUNIFORM SIZE</u>

The design calculations of reactors using non-uniform sizes of solid feeds can be done if the residence time distribution of each size particles is known in the bed. With no carry-over of particles, the design calculations until now have been made on the assumption of equal mean residence times of each size particles in the bed and the average conversions are calculated (54, 56) from the following relationship:

$$1 - \bar{X}_B = \int_0^{dp_{max}} (1 - X_B(dp)) \quad po\ (dp)\ d(dp) \qquad \text{..... (eq. 3.2)}$$

The assumption of equal mean residence time of each size particle in the bed means that for a given value of space time, if a solid feed of wide size distribution is processed through a moving or a fluidized bed, the smaller particles in the feed will display in general much higher fractional conversions than the large particles. A preferred situation would be to increase the average time of stay of larger particles as compared to smaller particles by some appropriate method. When the present investigation was undertaken no work was reported on the residence time distribution of the flow behaviour of each size particle in a feed of wide size distributions. Since in almost all industrial application feeds of mixed sizes are used in moving and fluidized beds, a study on the flow behaviour of each sized particle, which was neglected so far, was undertaken.

3.3 THESIS OBJECTIVES

For accurate designs of industrial moving and fluidized beds, a critical experimental evaluation of the assumptions as described in sections 3.1 and 3.2 was proposed while planning the experimental work of the present investigation. The nature of experiments and the range of experimental work are described in the rest of this chapter.

3.3.1 Studies on Moving Beds

Morris et. al (78) observed slight deviation to plug flow in moving beds near the incipient fluidizing conditions and attributed it to the offtake system. In the present investigation the study of moving beds was, therefore, proposed by removing the effect of offtake system and for this aim the following two methods were, considered:

(i) to design an offtake system of a very simole geometry so that the end effects do not extend beyond a certain height of the bed.

(ii) to use double tracer injection system.

In this technique the response of one tracer injection at some height from bed bottom is considered as an imperfect tracer input signal and the response of a second tracer input at another bed height is considered the actual response signal of the first imperfect tracer input signal. The bed height of the first tracer injection being common to the second tracer injection, the end effects caused due to offtake system are thus eliminated by using the mathematical techniques of parameter estimation by the method of weighted moments, pattern search in time domain or the transfer function. The details of these methods are available in literature (3,4,80).

Apart from extensions of the work of Morris et. al., it was also proposed to investigate the effect on the residence time of each sized particle in the moving beds using mixed feeds. The use of tracers of different sizes in feeds of mixed sizes was considered appropriate for this work. Again, the double tracer injection system was planned for this study as well.

3.3.2 Studies on Fluidized Beds

As discussed earlier in Sections 3.1 and 3.2, the assumption of complete backmixing in fluidized beds has been verified by many investigators but the residence time distribution of each size particles in a bed using feeds with wide size distribution was not studied by earlier investigators. The only work of this kind was done in the fluidized beds with carry-overs and over-flows (67, 88, 108, 115, 120). In the proposed investigation this work was planned with the aim of verifying the validity of the assumption of identical mean residence times for particles of different sizes in the fluidized bed. It was also planned to develop condditions, if possible, to obtain lower mean residence time for smaller particles and higher mean residence times for larger particles.

3.3.3 Variables Proposed to be Investigated

The following variables were proposed to be studied in the present investigation of solids flow behaviour in the moving and fluidized beds:

(i) Particle size

(ii) Feed size distribution

(iii) Gas velocity

(iv) Solid feed rate

(v) Solid hold up,

CHAPTER - 4

EXPERIMENTAL SET UP AND PROCEDURES

4.1 GENERAL

Experiments carried out in the present investigation are divided into four main categories:

(a) Stimulus-response studies in moving and fluidized beds.

(b) Studies relating to hold-ups of different size particles in fluidized bed by vacuum removal technique.

(c) Axial and radial distribution of solids in fluidized bed.

(d) Effect of baffles on hold-up ratios in fluidized beds.

In all these experiments, main components used in the experimental set up were:

(i) Unit for supply of air at constant pressure

(ii) Main column for fluidization

(iii) Equipment for feeding and removing solids at controlled rates.

(iv) Sampling system for solids.

(v) Apparatus for vacuum removal of solids from the bed directly.

(vi) System for the study of radial and axial distributions.

(vii) Baffles for use in fluidized bed.

These components are discussed in detail in the following sections:

4.2 STIMULUS-RESPONSE STUDIES

Experiments for stimulus-Response technique were conducted both for moving and fluidized beds. The details of set-up used are described below:

4.2.1 Flow Diagram of Experimental Set-up for Stimulus Response Studies

A schematic flow diagram of the experimental set-up is given in Fig. 4.1. Air is compressed by the compressor to a pressure of 5 kgf/cm^2 and sent to a surge-tank (ST). The surge tank, apart from removing pressure fluctuations, acted as a moisture trap and before the start of experiment, moisture was removed every day from the Surge tank. The compressed air from the Surge tank passed through a pressure regulator which regulated the entry pressure for rotameters to a constant value. Three romaters in parallel were used for measuring the entire range of air flow during the experiments but only one rotameter was used at a time. The pressure at the rotameter outlet was controlled by simultaneously adjusting valves (G5) or (G6) or (G7) and (G4) or (G3). The air entered the main column (MC) through a honeycomb distributor (D1) and a sintered glass disc (SGD) of mean pore size of 50 micron. The solids were fed to the main column from the hoppers (H1) and (H2) and the solid feed rate from hopper (H1) was controlled by orifice (O1) and valve (V3) and from hopper (H2) was controlled by orifice (O2) and valve (V4). The solid discharge rate from the column was controlled by valve (V1) at the outlet.

C	AIR COMPRESSORS
CB	COLLECTING BOTTLE
D1	HONEY COMB DISTRIBUTOR
SGD	SINTERED GLASS DISC
MC	MAIN COLUMN
G	GLOBE VALVES
H1, H2	HOPPERS
M	MANOMETER
N	NORGREN PRESSURE REGULATOR
O1, O2	ORIFICE PLATES
P1, P2	PRESSURE GAUGES
R1, R2, R3	ROTAMETERS
SD	SAMPLING DEVICE
ST	SURGE TANK
V1, V2	TWO AND THREE WAY GLASS STOP COCKS
V3, V4	TWO WAY GLASS STOP COCKS
T	TRACER STORAGE

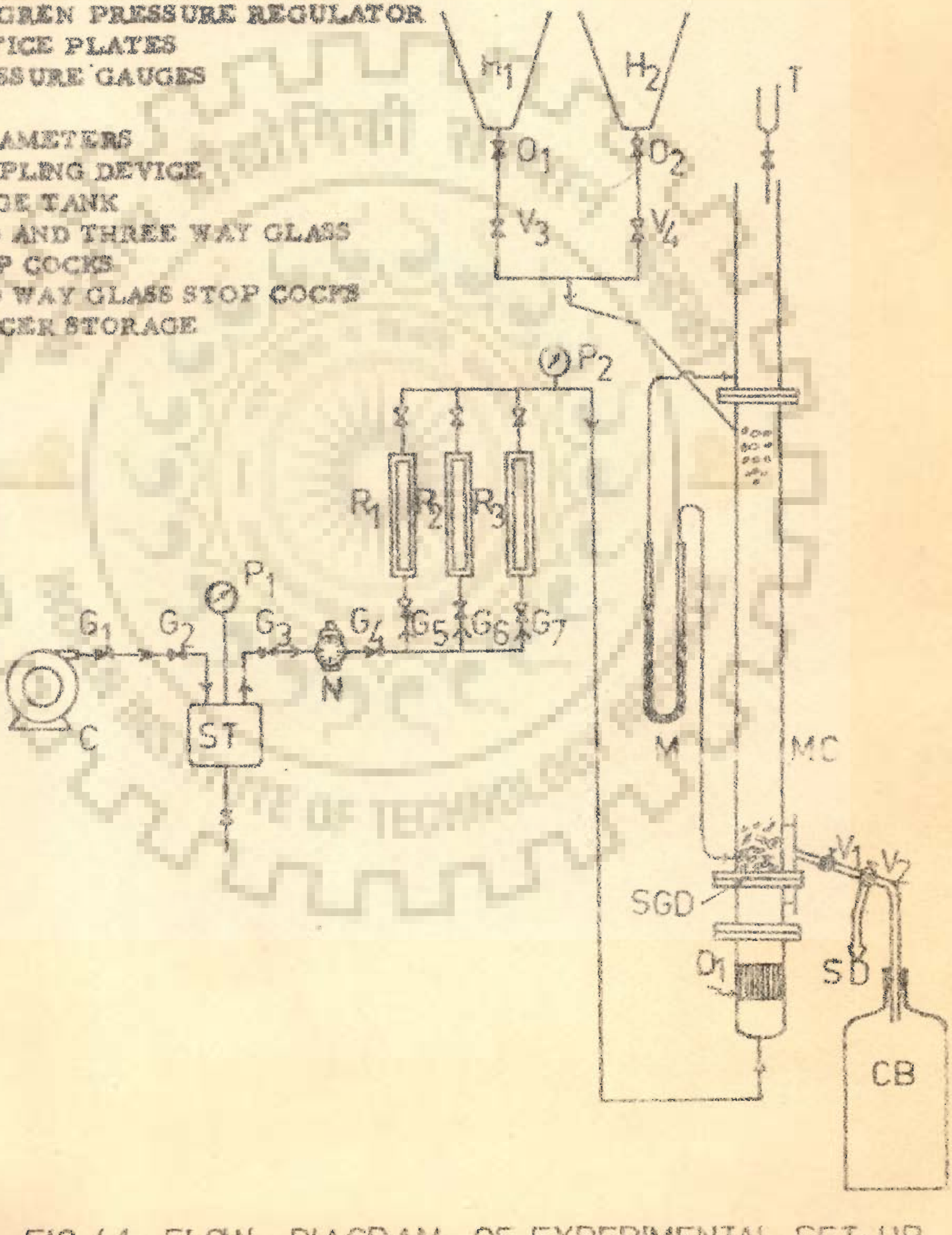


FIG. 4.1 FLOW DIAGRAM OF EXPERIMENTAL SET-UP

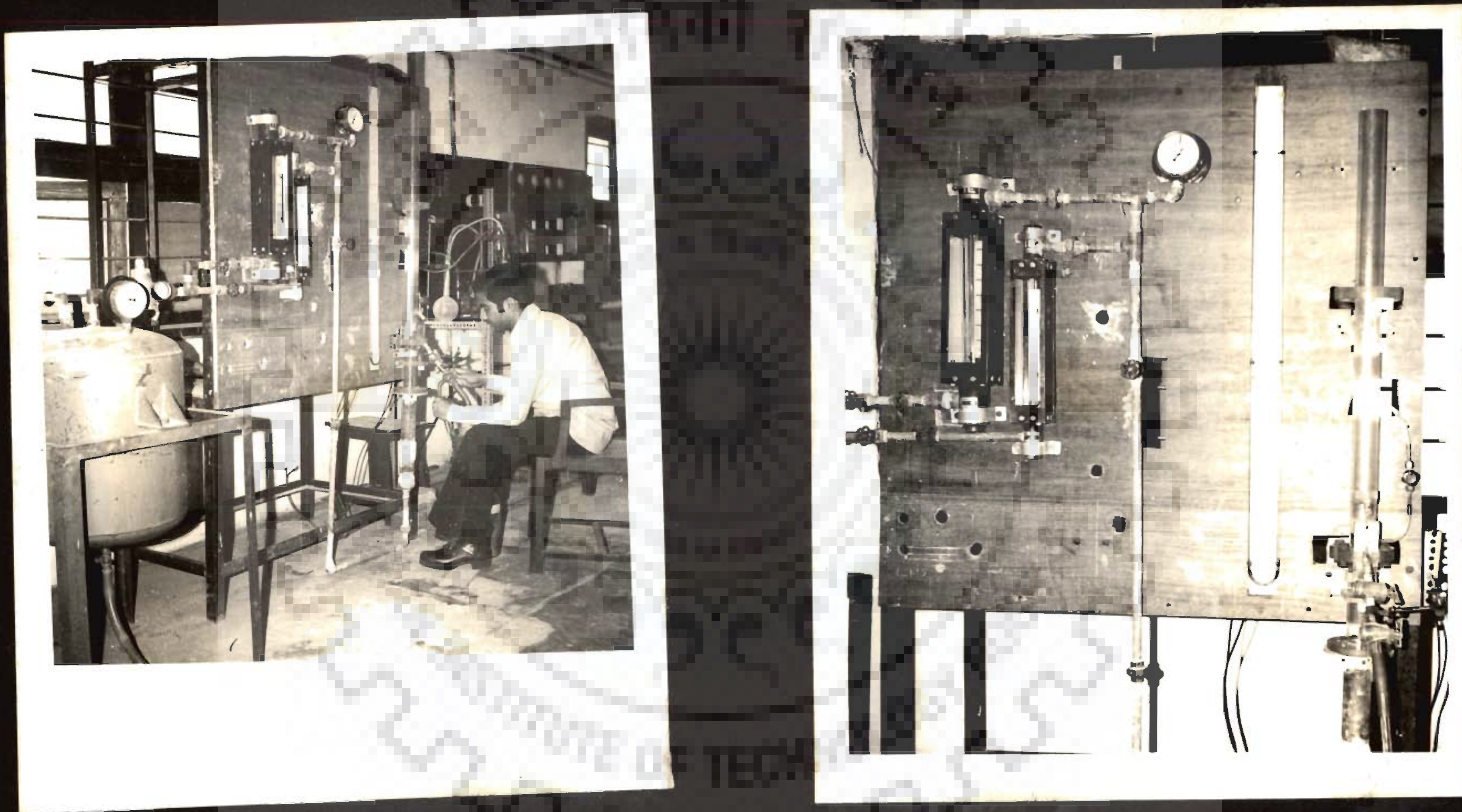
The solids leaving the column were collected in a collecting bottle (CB) through a three way stop cock (V2). The samples of solids leaving the column were collected at sampling tube (SD) by suitably adjusting the position of stop-cock (V2). A general layout of experimental set up is given in Photograph 4.1. The details of solid discharge and sampling system are given in photograph 4.2.

4.2.1.1 Air Compressor (C)

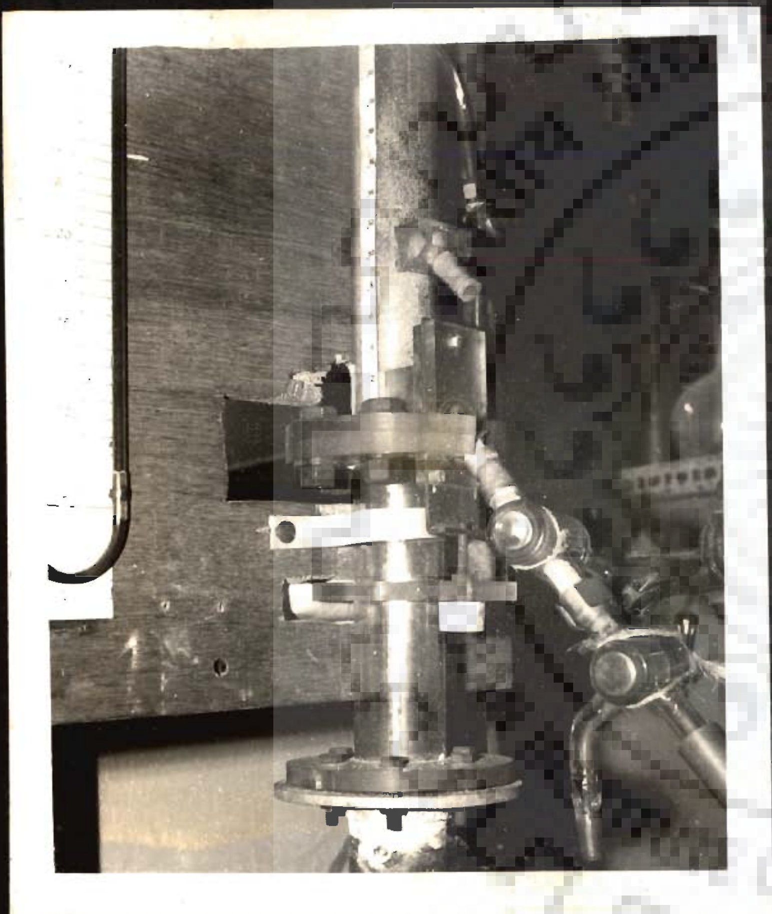
A single stage reciprocating compressor driven by a 5 H.P. motor and manufacture by M/S Anand Industries Corporation, Kanpur (India) was used for the supply of air. The specifications of the compressors are COMAIR brand, Model D.C. V-4, No. V599, displacement volume 20.4 cubic feet per minute at S.T.P. The compressor could supply air at a pressure of 12 kgf/cm^2 but for the present experiments the cut-off of the compressor was set at 5 kgf/cm^2.

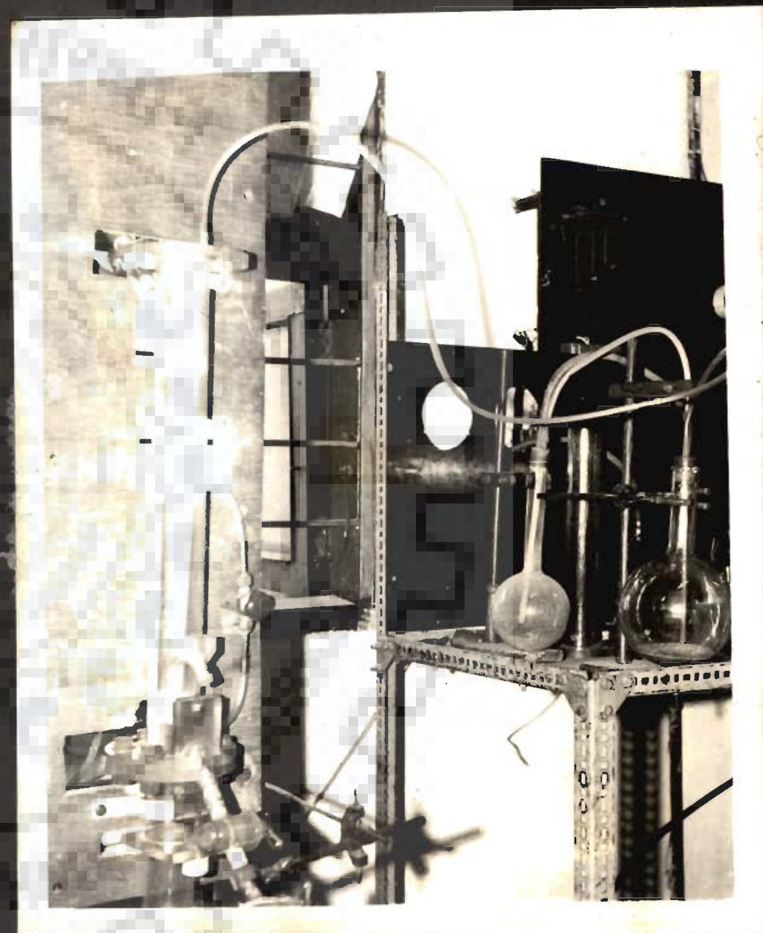
4.2.1.2 Surge-Tank (ST)

For trapping the moisture and damping pressure fluctuations a surge tank was provided. A cylindrical tank with torispherical heads and having a total volume of about 500 litres was used for this purpose. It was hydraulically tested for a pressure of 15 kgf/cm^2. The tank pressure was indicated by a bourdon-type pressure gauge having a range of 0-14 kgf/cm^2. The pressure in the Surge tank was always close to 5 kgf/cm^2.



PHOTOGRAPH 4.1 GENERAL LAYOUT OF EXPERIMENTAL SET-UP





PHOTOGRAPH 4.2 SOLID DISCHARGE AND SAMPLING SYSTEM

4.2.1.3 Pressure Regulator (N)

The pressure of air from the surge tank was further regulated by pressure regulator. The regulator was supplied by Norgren Co., make RO2 - 400 - RGC - BD. The outlet pressure of the regulator was indicated by a pressure gauge with range 0 -2.1 kgf/cm^2. For any inlet pressure, the outlet pressure could be adjusted by a knob on the pressure regulator and the regulator outlet pressure was kept at 1.5 kgf/cm^2 for all the experiments.

4.2.1.4 Rotameters

Three rotameters were used in parallel to cover the entire air flow rate range of 0 to 600 litres per minute at S.T.P. The specifications of the theee rotameters used for the experimental work are :

(i) Rotameter (R 1)

Make IE, Brooks tube size R-3M - 18 Graduations. 500-3500

Flow range: 0-40 LPM at S.T.P.

(ii) Rotameter (R 2)

Make IE, Brooks tube size R-8M-25 -2.

Graduations 0.25 to 3.00

Flow range 15 LPM to 160 LPM at STP

(iii) Rotameter (R 3)

Make IE Brooks tube size BR -$\frac{1}{2}$ - 27 G 10

Graduations 2500 to 25000

Flow range: 80 LPM to 500 LPM at S.T.P.

The calibration was done by two separate dry gas meters, one for rotameter (R 1) and the other for rotameters (R 2) and (R 3). The specifications of the dry gas meters used for calibrations are given below:

(i) Dry gas meter = (G M1)

Make UGI (Meters) Ltd., London

Graduations 1 cuft. per revolution

Capacity 200 cuft. per hour

(ii) Dry gas meter (GM 2)

Make UGI (Meters) Ltd., London

Graduations 2 cu. ft. per revolution

Capacity 400 cu. ft. per hour.

4.2.2 Air Distribution

$\frac{1}{2}$ inch diameter standard galvanized iron pipe was used to carry compressed air from the rotameter to a 30 cm long, 2" diameter galvanized iron pipe section fitted just below the main column. A 15 cm long and 5 cm diameter honey-comb was inserted in this pipe section for making the air flow direction vertically upwards in the entire pipe cross section. The position of the honey comb was 10 cm from the calming section and it was supported at the bottom and top by corss-wire support. Honey comb was made from paper pipes (Sipping straws) used for drinking soft drinks. The paper pipes were glued to each other in hexagonal cross-section and then cut to the desired specifications. The galvanized iron pipe section was connected to a 20 cm long 5 cm diameter calming secti on tube with the help of a flange.

The calming section was connected to the main column as shown in Fig. 4.2. Air from calming section (CS) entered the main column through a sintered glass disc (SGD) having pore sizes less than 0.05 mm. The sintered glass disc was fixed to the inside of the main column with the help of 'araldite' adhesive. Care was taken to ensure that 'araldite' did not block the pores of the sintered glass disc and that there was no leakage of air around its periphery.

4.2.3 Main Column (MC)

The column used for all fluidization and moving bed experiments was a 50 mm internal diameter perspex column. It was connected to the calming section with the help of special perspex flanges with an arrangement for solid removal. The details of solid-removal system are given in Fig. 4.3. Two perspex rings (PR) were used as sleeves for the main column and calming section column and these were fixed by 'araldite' adhesive to special flanges (F) in Fig. 4.4. The flanges were provided with five holes for 12 mm sizes bolts and were specially cut at one side to accommodate a brass sliding plate (BSP) and the holders for sliding plate of solids. The main purpose of the brass sliding plate was to provide a leak proof mechanism as close to the column as possible for regulating the solids withdrawal rate. The sliding plate holder (SPH) and perspex tube holder (PTH) were made from perspex sheet and they were provided with grooves for rubber 'O' rings as shown in Fig. 4.5. The sliding plate holder, brass sliding plate and the

BSP BRASS SLIDING PLATE
CS CALMING SECTION
D_1 HONEY COMB
HS HOLDER FOR SCREW
MS M. S. PIPE
OR O' RING
PTH PERSPEX TUBE HOLDER
PT PERSPEX TUBE
PTO PERSPEX TUBE OUTLET
S SCREW
SGD [illegible]RED GLASS DISC
TM TAPPING FOR MANOMETER

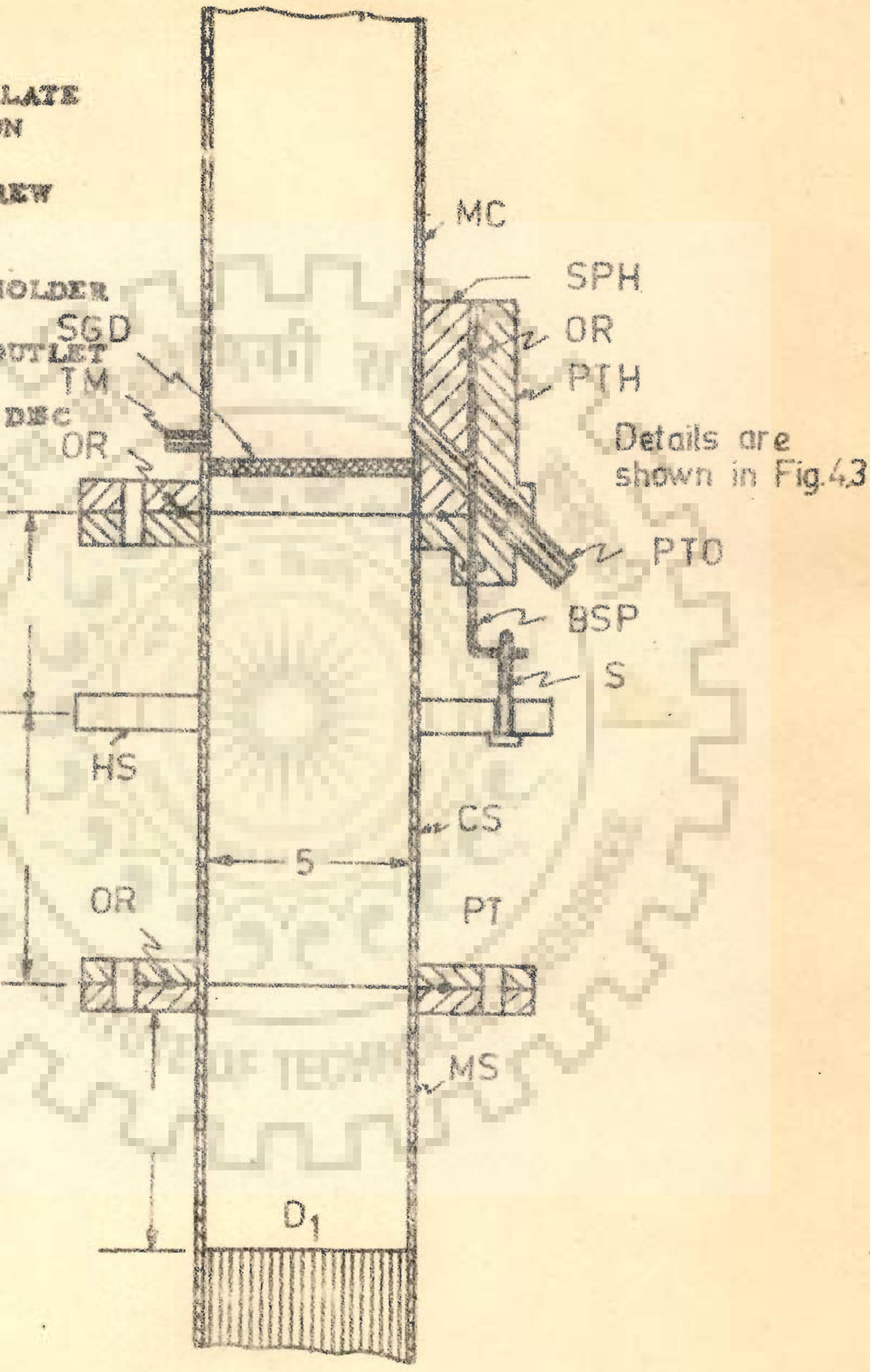
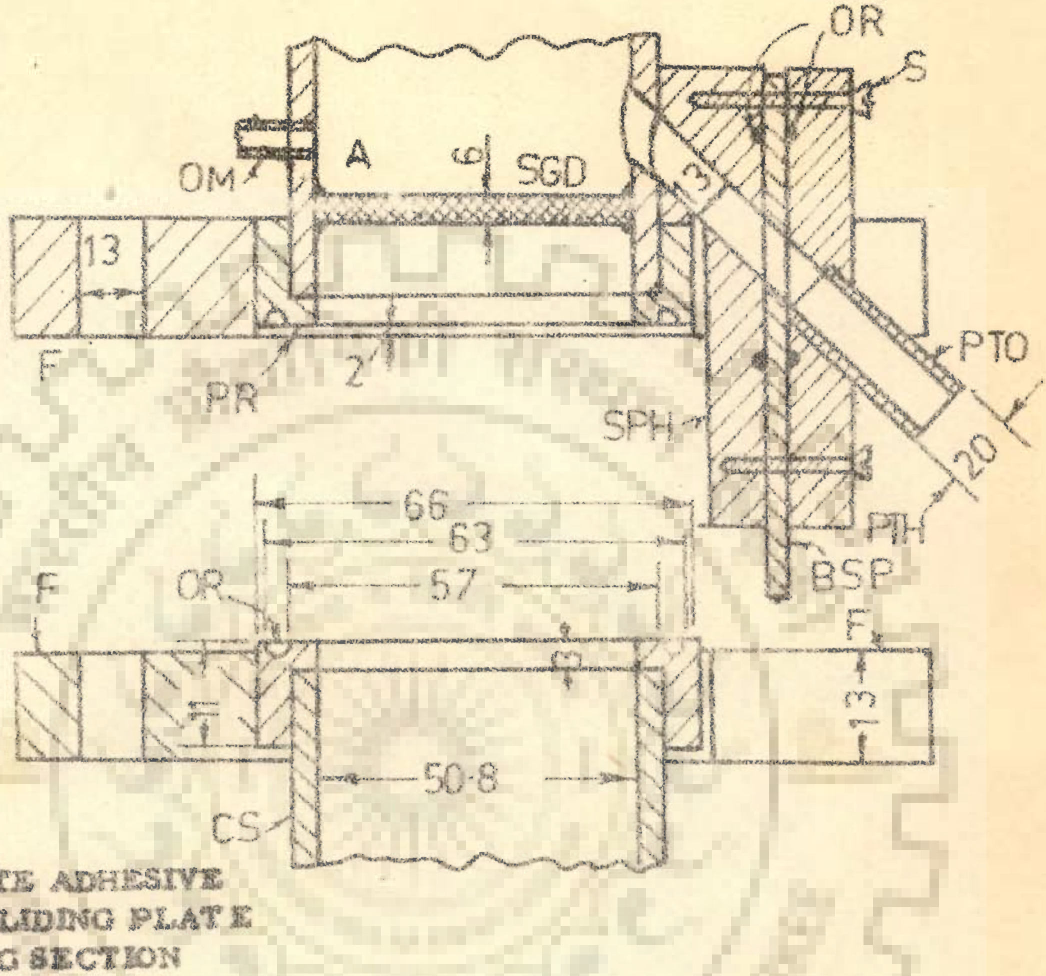


FIG.4.2 CONNECTION OF MAIN COLUMN TO HONEY COMB DISTRIBUTOR



A ARALDITE ADHESIVE
BSP BRASS SLIDING PLATE
CS CALMING SECTION
F SPECIAL FLANGE
OR O' RING
PR PERSPEX RING
PTH PERSPEX TUBE HOLDER
PTO PERSPEX TUBE OUTLET
SPH SLIDING PLATE HOLDER
S SCREW
SGD SINTERED GLASS DISC
OM OPENING FOR MANOMETER

FIGURE 4·3 DETAILS OF FLANGE AT OUTLET CONNECTION

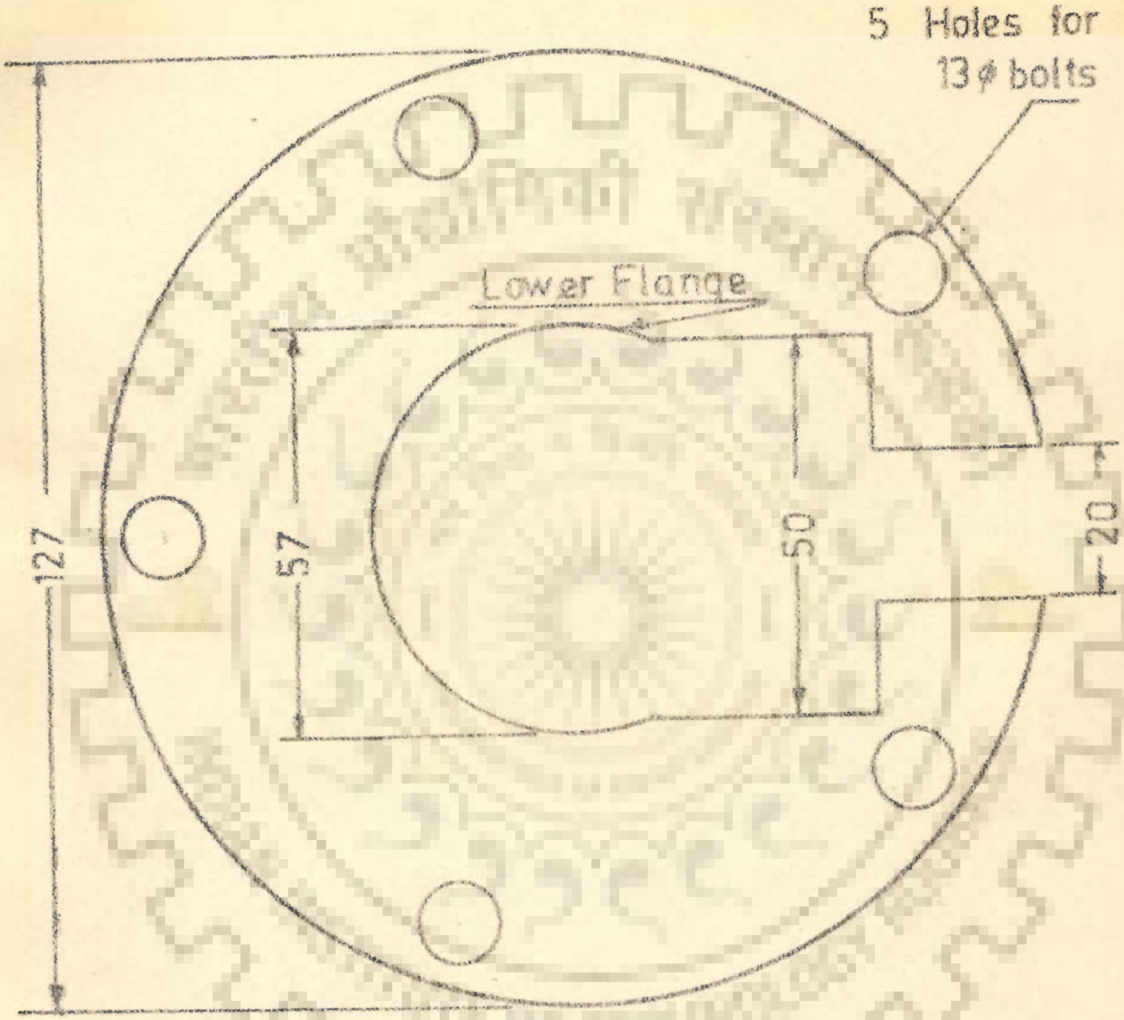


FIG. 4.4 SPECIAL FLANGE FOR SOLID REMOVAL SYSTEM

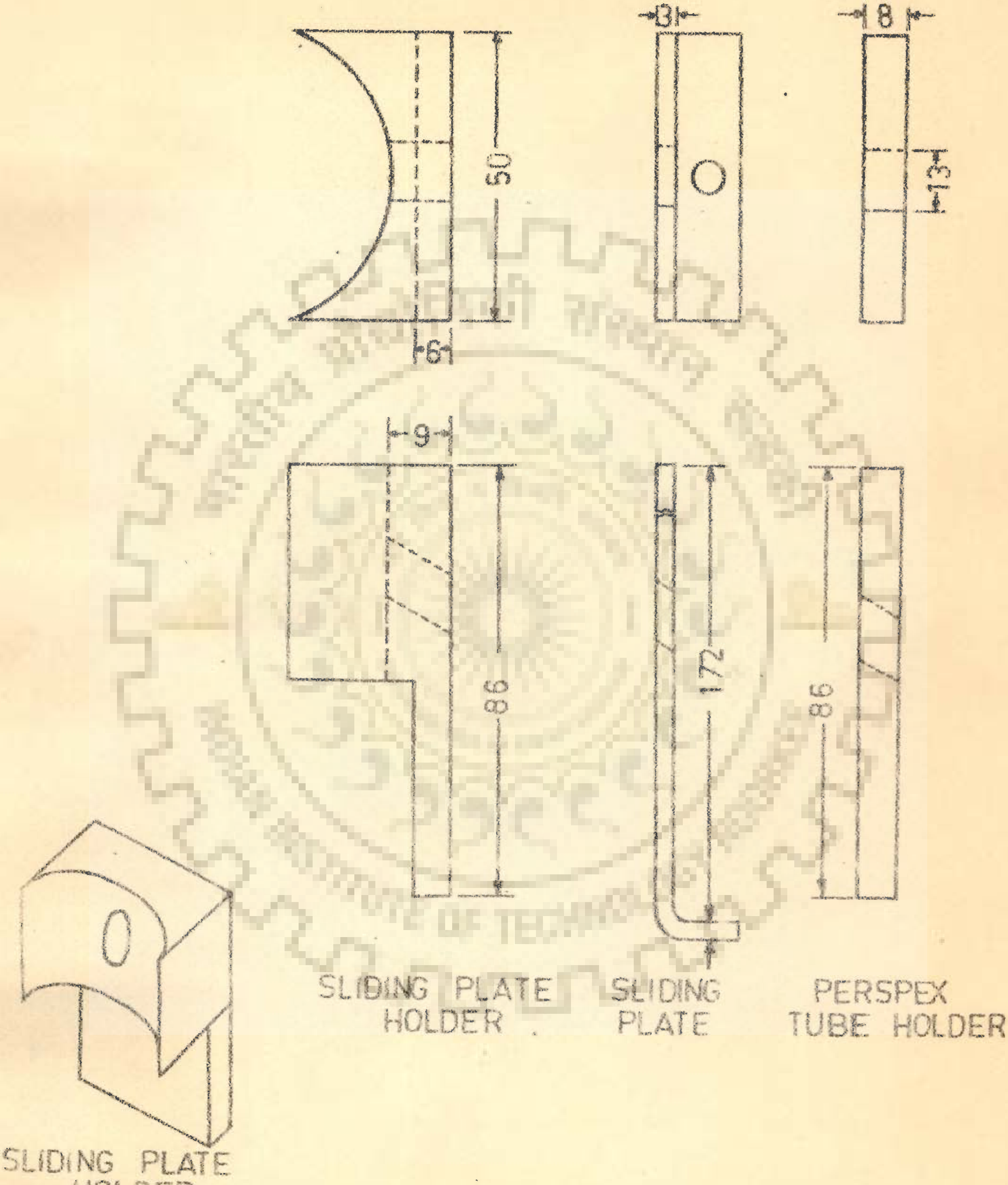


FIG. 4.5 DETAILS OF OUTLET CONNECTIONS

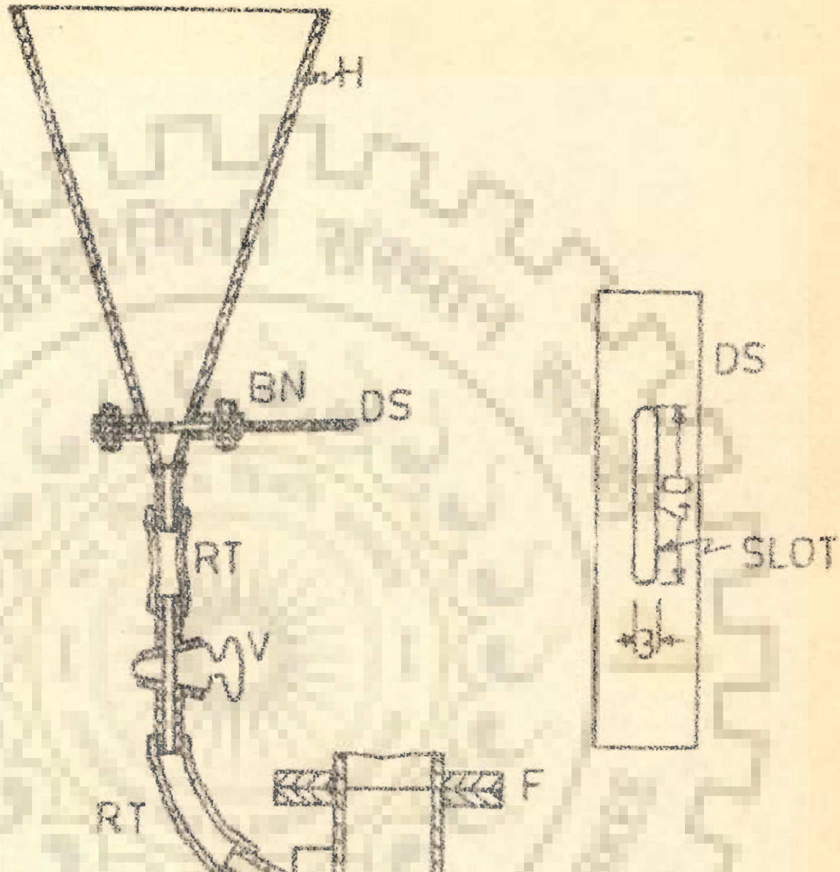
perspex tube holder were held in position with the help of screws which on tightening gave a leak proof joint due to the presence of 'O' rings (OR) as shown in Fig. 4.3. A 13 mm diameter hole was drilled at an angle of 45° in the perspex tube holder, the brass sliding plate, sliding plate holder and the main column just above the sintered glass disc for solids withdrawal. The hole was enlarged to 20 mm diameter in the outer portion of perspex tube holder to fit a perspex tube of 20 mm outer diameter and 13 mm inner diameter. All the perspex joints viz. main column (MC) and calming section (CS) to perspex rings (PR), main column to sliding plate holder (SPH), perspex rings (PR) to flanges (F) and perspex tube outlet (PTO) to perspex tube holder (PTO) were carefully made with the help of methylene dichloride. A very fine layer of methylene dichloride was applied on the surfaces to be joined and then these surfaces were pressed together to give leak proof joints free of any pin holes due to entrapped air. The perspex tube outlet (PTO) was connected to a two way glass stop cock (VI) with the help of polythene pressure tubing. This stop cock (VI) was then connected to a three-way stop cock (V 2) using polythene pressure tubing as shown in Fig. 4.1. One end of the 3-way stop cock was provided with a glass joint of size B-14 for the collection of out going solid samples. The other end of the glass stop cock was connected to a collecting bottole with the help of pressure rubber tubings and a rubber cork. A small perspex tube of size 6 mm with a bore of 1.5 mm was fixed in the main column at a distance of 2 mm. above the sintered glass disc for manometer connection to measure the pressure drop across the bed.

The brass sliding plate was made from a brass plate and its surface was finely machined. The surfaces of perspex tube holder and sliding plate holder were also smooth for easy vertical movement of brass sliding plate controlled by a screw arrangement as shown in Fig. 4. 2.

4. 2. 4 Feeding Arrangement

The main column was 60 cm long. An additional column of 30cm of length was fitted to the main column by use of another flange. At a distance of 3 cm below the upper flange of the main column a perspex block was attached to the main column. To this block two perspex tubes of 13 mm size were fixed at an angle of 45° with horizontal. These perspex tubes were connected to hoppers (H1 and H2 in Fig. 4. 1) through two way stop cock. Details of this system are given in Fig. 4. 6. Each hopper had a capacity to contain 10 kilograms of glass ballotini beads. This amount was sufficient for one run.

The hopper was made from mild steel plate and at its bottom a disc with slot (DS) was fitted as shown in Fig. 4. 6. The disc could be moved inward or outward, thereby the area of the orifice coming under the solid material could be adjusted to regulate the feed rate of solids. The pulse tracer was introduced through another two way stop cock fitted at the top of the main column (T in Fig. 4. 1).



BN BOLTS AND NUTS
DS DISC WITH SLOT
F FLANGE
H HOPPER
MC MAIN COLUMN
PB PERSPEX BLOCK
RT RUBBER TUBING
PT PERSPEX TUBE
V GLASS STOP COCKS

FIG.4.6 DETAILS OF FEED ARRANGEMENT

4.2.5 Equipment Erection and Testing

The main column was clamped on a wooden board, care was taken to keep the column perfectly verticle with the help of a plump-bob. All the fittings were then tightened properly and c hecked for leaks by using soap solution. The rotameter valves and fittings were checked for leaks at an air pressure of 4 kgf/cm^2 by closing the valve (G 4) in Fig. 4.1. The valve (G8) was then closed completely and all other valves kept open. A constant condition of pressure in pressuge gauge (P 2) ensured leak-proof joints. Leaks in the calming section and the main column were checked only by soap solution.

Manometer (M in Fig. 4.1) was connected by polythene tubings. Water was used as manometer fluid and slight colour was added to it for clarity of readings.

4.2.6 Tracer Material

The tracer was prepared from glass ballotini beads (spherical glass particles) and covering them with a fine layer of below 300 mesh iron powder. In about 200 g of glass ballotini beads 2 milli litre of glue was added and mixed properly to provide sticky surface for iron particles. Nearly 2 grams of iron powder was then sprinkled over these glass ballotini beads. They were well mixed by hand and left for drying for nearly 48 hours. It was found that the fine iron particles were converted to iron oxide and were quite firmly stuck to the glass ballotini beads. Uncoated particles were separated from coated particles by using a strong magnet. The coated particles were separated into different sizes by sieves and were kept for use as tracer material. The density of glass ballotini beads was found to change,, due to the coating of iron oxide particles, from 2.48 to 2.50, which was considered negligible. The effect of coating on the size and flow characteristics of ballotini beads was quite negligible because of extremely fine coating.

4.2.7 Experimental Procedure

The apparatus was first standardised by determining the minimum fluidizing velocity and comparing it with available equations in literature. Minimum fluidizing velocity was determined by plotting the pressure drop as a function of velocity. The glass ballotini beads were separated into three main sizes:

(i) -36, + 40 mesh (dp = 0.393 mm)

(ii) - 22, + 25 mesh (dp = 0.724 mm)

(iii) -16, + 18 mesh (dp = 0.96 mm)

Stimulus-response studies were planned into the following categories :

(i) Moving bed studies for single sized feeds

(ii) Fluidized bed studies for single sized feeds

(iii) Studies on mixed feeds.

In all these studies the procedure adopted was as follows:

4.2.7.1 Air Supply at Constant Rate

The air compressor (C in Fig. 4.1) was started and the Surge Tank (ST) was cleaned of accumulated moisture and was maintained at the required pressure of 5.0 kgf/cm^2. Pressure at the rotameter outlets was maintained at 1.5 kgf/cm^2. Depending on the desired air flow rate only one rotameter inlet valve was kept open, adjusted and the other two valves were kept completely closed. Steady pressure at the outlet of rotameter for a particular air flow rate was maintained by simultaneously operating valves (G8) and (G9) shown in Fig. 4.1. After adjusting the pressure and air flow rate, air was temporarily stopped by closing the valve (G4). The valve (G4) was opened only after the steady state conditions of the solid flow were maintained in the main column.

4.2.7.2 Feeding of Solids

The metal disc with slot (DS in Fig. 4.6) below the hopper (H1 or H2) was adjusted by trial and error to give the required flow rate of solids. Solids used in all the runs were glass ballotini beads supplied by M/s Haldyn Glass Works (Pvt) Ltd., Bombay. The metal disc with slot arrangement gave best results for maintaining constant solid feed rates. Earlier a star feeder and a screw conveyor arrangement for regulating solid feed rates were also tried but they did not give the desired result. The glass ballotini beads often blocked the clearance between screw as well as star feeder and its body, and this changed the speed of the driving motor, thus making it difficult to maintain the solid feed rates at constant valve. The orfice arrangement and gravity feeding was found to be very reliable and accurate. There was also no need for fixing a vibrator because the flow ofballotini beads was quite smooth. The feeding could be completely stopped or started with the help of glass stop cock (V 3 or V4) and the adjusted solid feed rates were found to be reproducible.

For experiments with mixed feeds, more hoppers (one hopper for each ballotini bead size) were used to feed the ballotini beads. In the earlier stages of this investigation premixing of different size ballotini beads to prepare the feed before introduction to the main column through a single hopper was tried but this had to be abondoned due to a marked tendency of segregation of different sized bellotini beads while flowing in the hopper.

4.2.7.3 Maintenance of steady solid flow

After adjusting constant feed rate of solids, the discharge rate of solids was regulated by adjusting the two way glass stop cock (V1 in Fig. 4.1). The regulation of steady discharge rate of solids was not done by sliding brass plate because the adjustment was more time-consuming. The velocity of solid in the outlet section was large enough to neglect its effect on residence times. The solids were continuously collected in the collecting bottle (CB in Fig. 4.1). The adjustment of the two way stop cock was also by trial and error but did not require more than 3-4 minutes. The steadiness of flow was checked by observing a constant height of moving bed for about 15 minutes. A proper bed hold up was adjusted by adding or removing from the bed a desired amount of the feed material after stopping the air flow momentarily.

4.2.7.4 Tracer Introduction

A weighed amount of clean tracer material of proper size was kept in the glass tube of the two way stop cock placed at the centre of the main column (T in Fig. 4.1). A sudden introduction of tracer to the top of the bed in the main column was possible by completely opening the valve without disturbing the bed conditions. However, some runs were also carried out in which the tracer had been introduced in the middle of a moving bed. This was done with some difficulty in the following manner. After adjusting the air flow, the valve (G4) was closed temporarily and the measured amount of tracer was dropped at the required height of moving bed. Then additional quantity of solids, sufficient to bring the level of static bed to the required height were dropped very quickly into the bed from top of the column. After that, the air flow was started by fully opening the valve (G4) to start the air flow at the desired rate. The total time taken for this addition was about 3 seconds which could be ignored in comparison to the mean residence times of nearly 300 seconds.

4.2.7.5 Sampling

A sampling tube with a glass joint B-14 (female) was tightly attached to the male B-14 joint of the three way stop cock (V2 in Fig. 4.1). Samples were collected at an interval of 10 to 15 seconds for a duration of 4 seconds each time. The tight fitting glass joint ensured air tight system. The samples were analyzed by weighing in a single pan balance having an accuracy of 0.0005 gram.

4.2.7.6 Feed Recovery

The ballotini beads leaving the column along with the tracer were collected in collecting bottle (CB in Fig. 4. 1). Tracer ballotini beads were separated from the other ballotini beads by keeping powerful magnets in three funnels, and passing the mixed ballotini beads through them in a series arrangement. The removal of tracer mate terial was almost complete from the feed material.

4. 3 STUDIES ON HOLD UP RATIOS

No tracer material was used in these experiments. For hold up ratio experiments the steady air flow rate and solid feed rate were maintained in the same manner as described in section 4. 2. All these experiments were conducted with mixed feeds and, therefore, while checking the steady solid feed rate it was also essential to check the compositions of the inlet and outlet solid streams. The composition of the inlet feed was known by prior adjustments of the orifice plates and the composition of outlet steam was determined by sieve analysis of the samples withdrawn from time to time.

After ensuring steady state flow and compositions all the stop cocks (V2, V3, V4 in Fig. 4. 1) were simultaneously closed and at the same instant air flow also was stopped. This arrested the condition of the bed hold up at the steady state for that run. The material remaining inside the bed was removed by suction through a glass tube, connected through a flask to a vacuum pump of the following specifications:

Make: Pressure Vac, No. 690660
Vacuum: 28" mercury
Displacement: 70 cm
Motor: 0. 25 H. P.

All the ballotini beads from the bed were collected in the flask and these were separated into different sizes by sieves and then weighed separately to an accuracy of $\pm$ 0.5 gram. The ballotini beads collected in the collecting bottle in all these experiments were separated by sieves for use as feed again. The sieves used for separation were 20 mesh and 30 mesh. Since these mesh sizes fall quite in between the sizes of feed material, -36 + 40 mesh, -22 + 25 and -16 + 18 mesh, separation was quick and complete.

4.4 STUDIES ON AXIAL AND RADIAL DISTRIBUTION

For a better understanding of the behaviour of solid particles in the bed under fluidizing conditions an attempt was made to study the axial and radial distribution of solids under different conditions. The details of experimental procedures for these runs are described in the following sections.

4.4.1 Axial Distribution

To study these distributions two methods were employed. The first method related to the study of axial distribution by a controlled withdrawal of solid particles from different layers of beds by vacuum technique. The second method involved the change of the axial position of tube for solid discharge from the column.

4.4.1.1 Controlled Withdrawl from Different Layers of Beds

After attaining the steady state and maintaining it for enough time, inlet and outlet flow of solids and the flow of air were all stopped simultaneously. This resulted in a settled bed of solids. It was assumed that the axial distribution did not change much during the settling of fluidized particles. The particles from top 20 per cent of the bed were then carefully sucked by vacuum by observing the change in bed height and analysed for their sizes. This procedure was repeated and subsequent layers were sucked to know size distribution at different axial positions of the bed.

4.4.1.2 Axial Position of Outlet for Solid Discharge

The other obvious way of studying the axial distribution was to keep the outlet for solid discharge at different heights of the bed. Only one more outlet position was studied at a height of 50 mm from the bottom of the bed. This is shown in Fig. 4.7 (b). Depending upon the air velocity in the bed, the outlet at 50 mm from bottom corresponded to different heights of fluidized bed. The study of particle size variation in the bed at steady state for different operating conditions, therefore, gave a rough measure of axial distribution.

4.4.2 Radial Distributions

The radial distribution of particles of different sizes was considered ~~very~~ important for the study of the effect of different parameters on fluidized bed composition. Since adequate photographic facilities were not available, the radial distribution was studied by changing the radial position of solid withdrawal connection. Two different arrangements were used, which have been explained in Sections 4.4.2.1 and 4.4.2.2. A second study on radial distribution was made through the use of an annular insert as explained in 4.4.2.3.

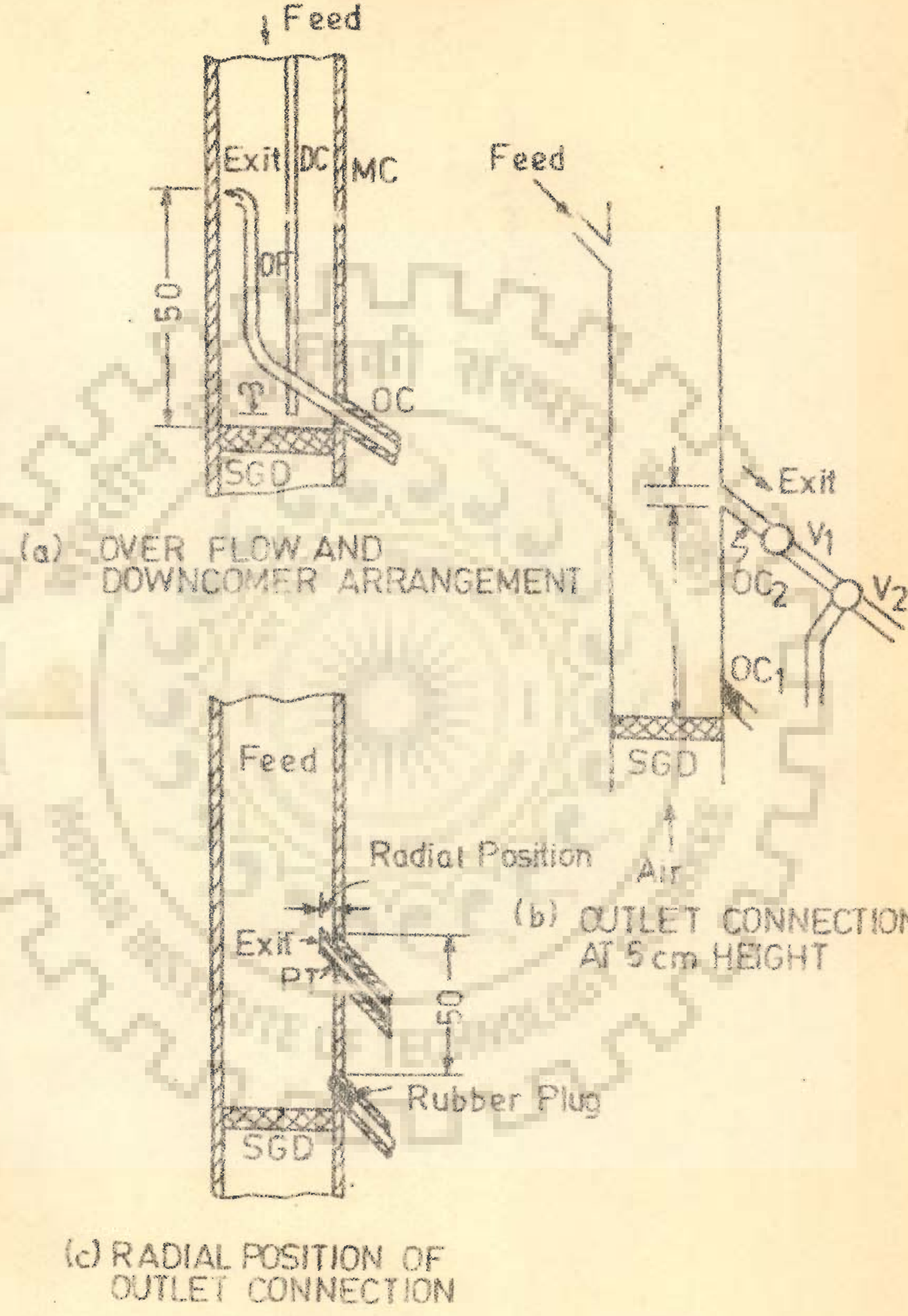


FIG.4.7 OFFTAKE SYSTEMS FOR AXIAL AND RADIAL DISTRIBUTION STUDIES

4.4.2.1 Overflow and Downcomer Arrangement

Overflow and downcomer arrangements are used very widely in the industry. The experimental set-up was slightly modified to study the overflow and downcomer arrangement by inserting an overflow tube of polythene into the bed through the outlet connection. The size of polythene tube was chosen such that its OD was same as the ID of outlet connection and, therefore, it was fitting very tightly into the outlet. The end of the polythene was cut vertically so as to permit radial entry of solids into it. The radial position was 10 mm away from the bed periphery. The radial position of the polythene tube could not bechanged because of practical limitations. The solids were fed to the main column by using a glass downcomer tube of size 6 mm. In Fig. 4.7 (a) is given the sketch of overflow and downcomer arrangement.

4.4.2.2. Overflow Connection at Bed Periphery

The radial position of the solid discharge system was an important parameter. In the overflow and downcomer arrangement the radial position of the solid discharge could not be varied. An additional solid discharge system was, therefore, provided in the experimental equipment by connecting a perspex block to the main column at a height of 50 mm and drilling through it a hole at an angle of 45°. A perspex tube of 13 mm bore was attached to the block as shown in Fig. 4.7 (b).

If the radial distributions of particles of different sizes was not uniform, then the radial position of outlet for solid discharge could be an important parameter affecting the fluidized bed composition. Experiments were, therefore, conducted by inserting a tight fitting polythene tubing into the outlet connection so as to provide outlet for solid discharge at different radial positions inside the main column. This is shown in Fig. 4.7 (c).

4.4.2.3 Use of Annular Insert-

A second way of studying radial distributions was by arresting the particles in their radial positions with the help of an insert having anulii of different diameters. This type of insert was made by first making two cylindrical tubes from galvanized iron sheet of 1 mm thickness, one tube of 25 mm diameter and the other of 42 mm diameter. They were held in concentric position using a brass rod at the centre and by soldering diagonal wires at the bottom and at the top of the cylindrical tubes. Three pins of length 3 mm each were also soldered to the outside of the outer tube to act as guides for keeping the insert concentric to the main column. The details of this insert are shown in Fig. 4.8. The radial particle size distribution runs were tried by fluidizing a batch of solid of predetermined weight and size distribution at different air flow rates. After attaining the steady state, the concentric-tubes insert was quickly dropped into the main column. The guides soldered at the outside of the outer cylinderical tube helped to keep the entire insert concentric to the main column. Simultaneous to the dropping of the insert,

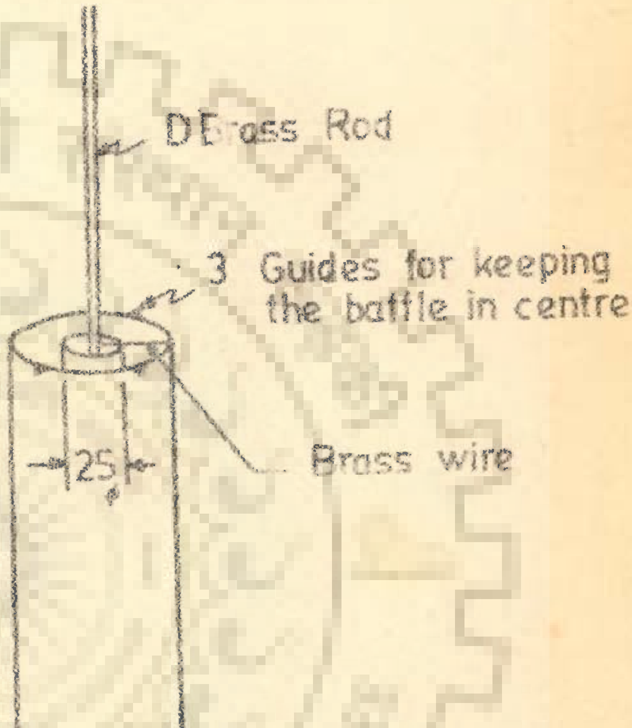


FIG. 4.8 CONCENTRIC TUBE INSERT FOR RADIAL DISTRIBUTION

the air flow was also stopped andthe particles were arrested in their radial positions. The particles in the central and the middle annulii were first sucked by vacuum separately and then the insert was removed and the particles present in the outermost annulus were sucked by vacuum. The three radial samples obtained in this way were analysed for the radial particle size distribution.

4.5 EFFECT OF BAFFLES

After analysing some of the experimental data collected so far, it was envisaged that the flow pattern of solids may get amply influenced by introduction of baffles in the fluidized bed. A few runs were taken by using baffles of two kinds which are shown in Fig. 4.9. These baffles were placed in the bed before the start of the run and apart from that the procedure of taking the runs remained exactly the same as described in Section 4.3.

The baffles were made from expanded metal wire net of diamond shape. This net had wires of 1.5 mm average diameter and were placed so as to give a rhombus pitch. The openings so formed by the wires had major axis 20 mm and minor axis 12 mm. Two kinds of baffles were mainly used in the experiments. The first type, Davisco baffle, was made by cutting two portions from expanded metal wire net of sizes 50 mm wide and 400 mm long and were soldered to a brass rod of 3 mm diameter as shown in Fig. 4.9 (a). Such baffles have been used earlier in fluidization studies by Davis et. al (1).

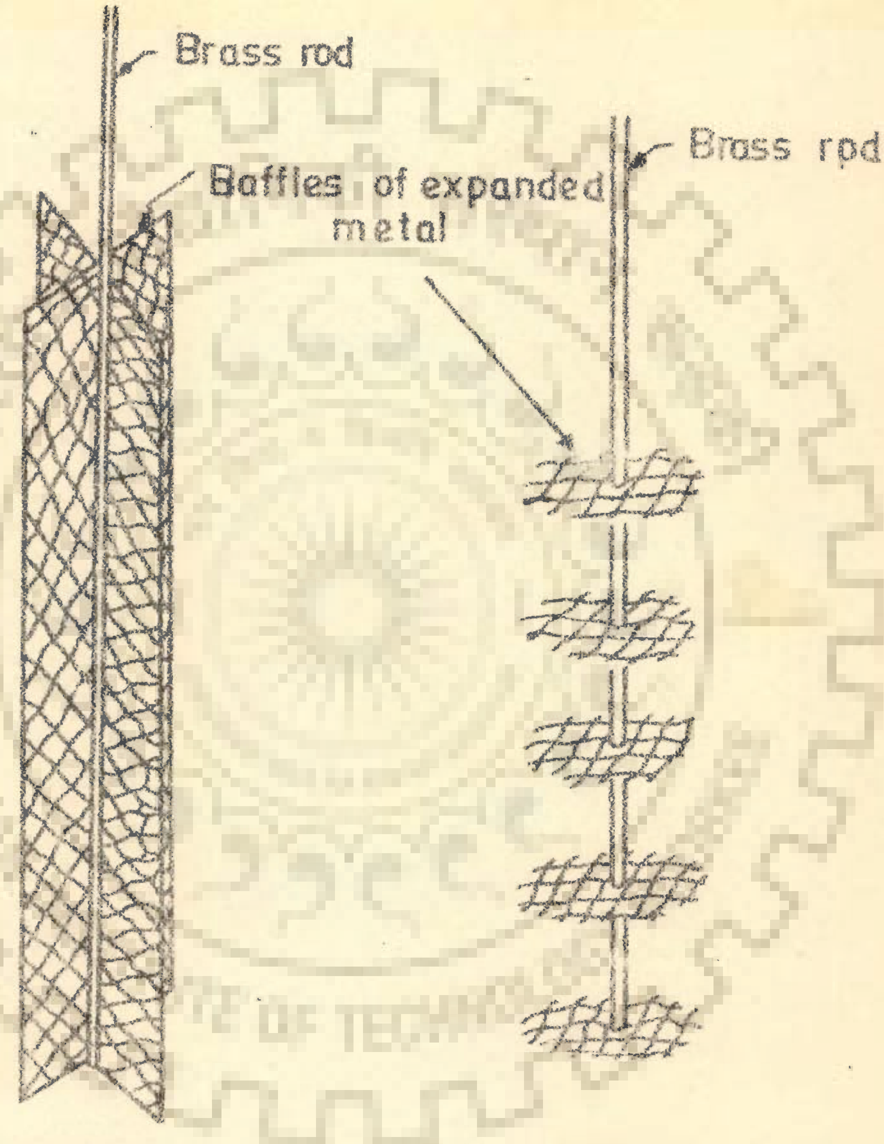


FIG.4.9 DETAILS OF BAFFLES

The second type of baffle used was also made from the same expanded metal wire net but instead of keeping the wire sheet vertical, circular discs of 50 mm diameter were cut from the wire net. These were placed at a distance of 50 mm each and soldered on a brass rod of 1.5 mm diameter. This type of baffle is shown in Fig. 4.9 (b). A third type of baffle, made from a wire screen of 6 mm holes, was also tried in some runs but was rejected because the quality of fluidization was severely affected by this type of baffle.

In the end, a few runs were taken with limestone particles of different sizes as the solid feed. These runs were carried out to ascertain the valadity of the general conclusions for non-spherical particles also.

CHAPTER 5

OBSERVATIONS AND CALCULATIONS

As discussed in Chapter 4, four kinds of experiments were conducted in the present investigation. The results of these experiments are presented in Sections 5. 3 to 5. 6 under the following main headings:

1) Stimulus-Response in moving and fluidized beds
2) Hold-up Ratios in fluidized beds
3) Axial and radial distribution of solids in fluidized beds
4) Effect of baffles on hold-up ratios in Fluidized beds.

Before start of fluidizing runs, the range of operation of different variables, equipment limitations were carefully determined and some preliminary investigations were undertaken to ensure proper functioning of the apparatus.

5. 1 EQUIPMENT LIMITATIONS

After the experimental set-up was erected and tested for leaks, some runs were taken to evaluate the limitations of the equipment. These limitations are discussed in the following subsections.

5. 1. 1 Air Flow Rate

The rated capacity of the available compressor was 20.4 cubic feet per minute i. e. 576 litres per minute. Three rotameters could cover the entire range of the air flow rate precisely. The sizes of ballotini beads for use in experimental runs was fixed on this basis.

For runs of moving beds, the highest air velocity envisaged was the minimum fluidizing velocity U_{mf}. The minimum fluidizing velocity for 2 mm diameter glass ballotini beads (density 2.4 88g/cc) was calculated from the equations available in the published literature and was estimated to correspond to 500 litres per minute of air flow at standard temperature and pressure (STP) for a column of 50 mm diameter. This flow rate was well within the range, and, therefore, beads upto a maximum size of 2 mm could be used for moving bed runs. For each run on any day, barometric pressure and temperatures were recorded and the effect of these variables was incorporated in calculating the air flow rate at the standard temperature and pressure.

5.1.2 Solid Phase

For all experiments, glass ballotini beads of density 2.488g/c.c. were used. These beads were transparent, solid and quite nearly spherical in shape. A normal industrial fluidized bed is operated at air flow rates corresponding to u/u_{mf} in the range of 5 to 10, therefore, the beads of a maximum diameter of 1 mm were chosen because for such beads the minimum fluidizing velocity corresponds to about 100 litres per minute. For these beads air flow rates were available to give velocities nearly six times the minimum fluidizing velocity. Ballotini beads of the following three sizes were mainly used in the experiments:

(i) Small particles (-36,+40 mesh) dp_{av} = 0.393 mm.

(ii) Middle size particles (-22, + 25 mesh), dp_{av} = 0.724 mm

(iii) Large particles (-16, + 18 mesh), dp_{av} = 0.96 mm

Particles smaller than 0.393 mm were not available and so they could not be used. However, except for coal gasification, most other industrial applications of fluidized beds use particle sizes above 0.1 mm, so the above sizes were considered to be quite satisfactory to give useful information.

5.1.3 Solid Feed Rate

In tracer runs it was considered desirable to have the mean residence time of particles in the bed in the range of 2 to 50 minutes. For this reason low solid feed rates were used in the experiments. The minimum solid feed rate attainable through hoppers and sliding disc system was a function of the particle size. For the small size particles a minimum rate of 25 g/min was obtainable by gravity feed without any difficulty but for the large size particles a feed rate lower than 90 g/min could not be obtained. In the study of moving beds maximum solid hold up of nearly 900 g was used, giving a mean residence time of about 10 minutes for large size particles. This was considered quite satisfactory. The apparatus could also be used to get higher mean residence time for solid but the non-availability of enough ballotini beads prevented the use of large hold-up in the bed.

5.1.4 Bed Hold-up

For moving bed runs, the first 10 cm height from the bottom of the bed was considered to be affected by the end effects. For each run of moving bed two separate runs were taken, one by introducing tracer at 10 cm height and the other by introducing tracer at higher heights. The maximum height chosen was 40 cm. giving an effective solid hold-up (corresponding to a static height of 30 cm.) of around 900 g. Hold-ups higher than this value could not be used because the total amount of desired size beads available for the experiments was exhausted before the completion of a run.

5.2 PRELIMINARY INVESTIGATIONS

The accuracy of the experimental set-up and procedure was checked by experimentally determining the minimum fluidizing velocities, u_{mf}. for ballotini beads of different sizes and comparing them from those calculated from the correlations available in literature.

The minimum fluidizing velocity was experimentally determined by plotting pressure drop ΔP as a function of air flow rate G curves. The pressure drop values at different air flow rates are tabulated for 0.393 mm size particles in Table A-1, Appendix A, for three static bed heights. Theoretical correlation (56) available to calculate the minimum fluidizing velocity is as follows:

$$u_{mf} = \left[\frac{\phi_s \, d_p}{150}\right]^2 \left[\frac{\rho_S - \rho_g}{\mu}\right] g_c \left[\frac{\epsilon_{mf}^3}{1 - \epsilon_{mf}}\right], \quad Re \leqslant 20 \qquad \ldots \text{(eq. 5.1)}$$

However, the above correlation is not very convenient to use because experimental methods for evaluating shape factor ϕ_s and voidage ϵ_{mf} are considerably involved. Fortunately the product $(\epsilon_{mf}^3 / 1 - \epsilon_{mf})\phi_s^2$ bears a simple relationship with dp and use of this relationship yields a simplified correlation for $Re_p < 5.0$ as given by Leva (59).

$$G_{mf} = 688\,(dp)^{1.82}\;\frac{(\rho_g(\rho_S - \rho_F))^{0.94}}{\mu^{0.88}} \qquad \ldots \text{(eq. 5.2)}$$

Three mixing ratios 1:0.25, 1:1 and 1:4 were tried for two combination of particles that is, small size (0.393 mm diameter) and middle size (0.724 mm diameter) particles and middle and large size (0.96 mm diameter) particles. These data are given in Tables A-2 to A-5, Appendix A. The average particle diameter dp for mixed feed was calculated from the equation 5.3.

$$dp = \frac{1}{\sum_i (x_i / dp_i)} \qquad \ldots \text{(eq. 5.3)}$$

Where x_i is the weight fraction of particles having diameter dp_i.

5.3 STIMULUS - RESPONSE STUDIES IN MOVING AND FLUIDIZED BEDS

These runs are in four different categories as described below:

5.3.1 Experiments on Moving Bed (Single size Feeds)

Before the start of the main experiments, the equipment was tested for reproducibility. In each run of moving bed, tracer was introduced at two points in the bed. One point of introduction was always kept at a height of 10 centimeters above the sintered glass disc while the other point of introduction was kept at different heights.

5.3.1 Factorial Design

A factorial design of experiments, was planned and the following variables were investigated.

(1) Air flow rate (u/u_{mf})

(2) Bed hold up (W)

(3) Solid feed rate (w)

(4) Size of solid feed (dp)

The study was first made at two extreme values of the variables and the total number of runs were reduced by planning them factorially. First set of experiments were carried out at the following extreme values of the variables:

(1) Air flow rate: $\frac{u_1}{u_{mf}} = 0.2$ and $\frac{u_2}{u_{mf}} = 0.95$

(2) Bed hold-up cm: $h_1 = 20$ and $h_2 = 40$

(3) Solid feed rate, g/min : $w_1 = 34$ and $w_2 = 300$

(4) Particle size, mm: $dp_1 = 0.393$ and $dp_2 = 0.96$.

For the test of reproducibility, mid values of all these variables were chosen and the reproducibility checked using

u/u_{mf} = 0.6

h = 30 cm

w = 100 g/min

dp = 0.724 mm

In Tables B-1 to B-3 of Appendix B, the data for these runs are given. After establishing good reproducibility, main runs were taken. Denoting the two extreme values by subscript 1 and 2, the factorially designed runs were taken according to the scheme given in Table 5.1. Experimental results for these runs are presented in Tables B-4 to B-9 of Appendix B. Run numbers 4 and 6 in Table 5.1 could not be carried out because the large size particles could not be operated at such a low flow rate of 34 g/min.

In all these runs samples were collected in the sampling tube and each sample was analyzed by taking the weight of the total sample and the weight of tracer in each sample.

5.3.2 Mixed Feed in Moving Beds

In order to determine the flow properties of one sized particles through a moving bed of mixed particle sizes, experiments were carried out by injecting pulse tracers of different sizes in the bed. It was not considered necessary to check the reproducibility of these runs again because they were similar to the earlier runs in all respects.

Three different kinds of mixed feeds were prepared by mixing appropriate sizes. Details of these experiments are given in Table 5.2. and all these experiments were carried out using u/u_{mf} = 0.95, h = 40 cm

Table 5.1 SCHEME OF FACTORIALLY DESIGNED EXPERIMENTS

Run No.	Air flow rate u/u_{mf}	Bed hold up h	Solid feed rate w	Particle size dp
1.	u_1/u_{mf_1}	h_1	w_1	dp_1
2.	u_1/u_{mf_1}	h_1	w_2	dp_2
3.	u_1/u_{mf_1}	h_2	w_2	dp_1
4.	u_1/u_{mf_1}	h_2	w_1	dp_2
5.	u_2/u_{mf_2}	h_2	w_1	dp_1
6.	u_2/u_{mf_2}	h_1	w_1	dp_2
7.	u_2/u_{mf_2}	h_2	w_1	dp_1
8.	u_2/u_{mf_2}	h_2	w_2	dp_2

and feed rate of each size particle w = 60 g/min. The total solid feed rate was 120 g/min for runs 1 to 6 and 180 g/min for runs 7 to 9. The experimental results of runs 1 to 3, 4 to 6 and 7 to 9 are tabulated in Tables C-1, C-2 and C-3 of Appendix C respectively.

5.3.3 Experiments on Fluidized Beds with Single Feeds

Experiments for fluidized beds were also designed factorially with the difference that the extreme conditions of air velocities were now different than in the moving bed experiments. The actual values of variables listed in Table 5.1 used for fluidized bed experiments are given below:

u_1/u_{mf_1}	=	1.2	u_2/u_{mf_2}	=	6.0
h_1	=	10 cm	h_2	=	30 cm
w_1	=	34 g/min.	w_2	=	300 g/min.
dp_1	=	0.393 mm	dp_2	=	0.96 mm.

While taking these runs it was observed that the injection of tracer particles at 10 cm height to study the end effects could not give the desired results because of the change in flow pattern due to a time lag between the dropping of the tracer particles and the dropping of the feed material by a beaker to obtain a bed with required amount of hold up. It was also observed that the mixing within the bed was quite intense so it was considered safe to neglect the end effects and so for fluidized bed tracer experiments, only one set of runs was taken instead of two sets as in the moving beds.

TABLE 5.2

, DETAILS OF EXPERIMENTS WITH MIXED FEEDS IN MOVING BEDS

Run no.	Mass Fractions of particles in the feed			Mass fractions of tracer particles		
	0.393 mm	0.724 mm	0.96 mm	0.393 mm	0.724 mm	0.96mm
1.	0.5	0.5	-	1.0	-	-
2.	0.5	0.5	-	-	1.0	-
3.	0.5	0.5	-	0.5	0.5	-
4.	-	0.5	0.5	-	1.0	-
5.	-	0.5	0.5	-	-	1.0
6.	-	0.5	0.5	-	0.5	0.5
7.	0.33	0.33	0.33	1.0	-	-
8.	0.33	0.33	0.33	-	1.0	-
9.	0.33	0.33	0.33	-	-	1.0
10.	0.33	0.33	0.33	-	-	1.0

Only four runs could be taken for the fluidized beds because run numbers 4 and 6 of table 5. 1 were not possible for such small solid feed rates of 0. 96 mm particles and run numbers 1 and 8 were not possible at such high velocities and such high hold-up values since the particles were thrown out of the bed. The results of experimental run numbers 1, 2, 3 and 5 in Table 5. 1 are tablulated in Appendix D, Table D-1.

5. 3. 4 Fluidized Beds with Mixed Feeds

Experimental runs similar to those mentioned in Table 5. 2 were also taken for fluidized beds using $u/u_{mf} = 3.0$ and $h = 10$ cm. The solid feed rate was kept at 90 g/min of each size particles for runs with feeds of two sized particles only and at 60 g/min of each size particles for runs with three sized particles so as to give a total feed rate of 180 g/min. The value $u/u_{mf} = 3.0$ was based on the u_{mf} value corresponding to the mean diameter of particles in the feed and not that in the bed at steady state. Experimental values of dimensionless concentration of tracer C as a function of time are tabulated in Tables D-2, D-3 and D-4 of Appendix D for run numbers 1 to 3, 4 to 6 and 7 to 10 respectively. The mass fractions of particles of different sizes in the feed for the above runs are already defined in Table 5. 2. Since the analysis of these runs showed some surprising behaviour not reported earlier in the literature, these runs were repeated 3 times to check their correctness and were observed to be completely reproducible.

While taking these runs, tracer material was introduced only after the steady state had been achieved. The steady state was checked by withdrawing samples of the outgoing stream and analysing their size distributions. The tracer was introduced only after the size distributions of the input and output streams completely matched. However, before the achievement of the steady state, it was observed that the amount of smaller particles in the samples was considerably more in comparison to the amount of larger particles. This observation suggested that the larger particles have a tendency to accumulate in the bed in comparison to the smaller particles and that the mean residence times of different size particles in the bed is likely to be different. Therefore, the hold-ups of different size particles in the bed at steady state should be different. To check this, some runs were taken for the particle size distribution in the bed after achieving the steady state. These runs were carried out under conditions similar to those recorded in Tables D-2 to D-4, Appendix D, but without the introduction of any tracer. At steady state for each run, size distribution in exit stream, which was the same as the size distribution in feed and size distribution in the bed, was recorded. The results of these observations are listed in Table D-5, Appendix D.

While adjusting feed rates of solids from hoppers through slotted discs, it was found very difficult to bring the rates exactly to the desired value. Moreover the feed rate sometimes changed by

$\pm$ 0.5% during a run and hence the ratio of particles in the feed could not be adjusted to exactly 1:1 or any other exact ratio. The ratios of particles in the feed could therefore be maintained only to within one half percent of the desired ratio. The error was taken into account by recording the size distribution in exit stream at steady state and assuming it to be the same as that of the inlet stream.

The experimental results of hold-up studies, Table D-5, Appendix D were found to be in total agreement with those obtained from the tracer experiments as explained in section 6.2.4. Since tracer runs were more time consuming, it was, therefore, decided to carry out further investigations about the fluidized bed behaviour mainly using hold-up studies for different size particles.

5.4 HOLD UP RATIOS IN FLUIDIZED BEDS

It has been pointed out in Section 5.2.4 that in a feed of mixed sizes, different sizes of particles had different mean residence times in the fluidized beds. This effect could be studied either by stimulus response technique or by a sieve analysis of bed hold-ups after the steady state is attained. The first technique is very time consuming and being an indirect method, it is not very accurate to find the hold-up ratios. Experiments were, therefore, conducted for hold-up ratios by the second technique and the following variables were studied.

i) Air flow rate

ii) Solid feed rate

iii) Bed hold-up

iv) Feed and Exit compositions

v) Particle size

5.4.1 Minimum Fluidizing Velocity

Since the size distribution in bed was not the same as that of feed or exit streams it would be wrong to define minimum fluidizing velocity on the basis of feed or exit size distribution. The distribution of particle sizes in the bed could be found only after the run was complete, so it was not possible to fix u/u_{mf} values apriori for any specified feed size distribution. For any flow rate, fixed arbitrarily, the values of u/u_{mf} were calculated after determining the hold-up ratios.

5.4.2 Effect of Air Flow Rate on Hold-up Ratio

The most important parameter for the study of hold-up ratios appeared to be the air flow rate. The range of air flow rates used in the experiments was fixed by two considerations.

(1) The minimum flow was kept higher than the minimum fluidizing flow rate for the highest sized particle in the bed. This was necessary because otherwise the higher size particles had a tendency to settle down at the bottom of the bed.

(2) The higher limit for air flow rate was kept low enough to prevent the entrainment of the lowest size particles. For the first set of runs, the following conditions were maintained.

(i) Feed composition: 50% each of 0.393 mm and 0.724 mm particles

(ii) Total solid hold-up = 90g

(iii) Bed height at static conditions = 2.5 cm.

(iv) Total solid feed rate = 120 g/min.

(v) Mean Residence Time = 90/120 = 0.75 min.

The air flow rate G for the above mentioned conditions was changed from 67.5 litres per minute to 215 litres per minute. The lower rate was fixed at 67.5 litres per minute. Since it was a little higher than the air flow rate required to fluidize 0.724 mm particles, the upper limit was fixed at 215 litres per minute because above this air flow rate the particles of 0.393 mm size were heavily entrained. In all, eleven flow rates of air were tried. For each flow rate a steady state flow of mixed solids was first established and then the bed material was arrested instantly, sucked out by vacuum and sieve analyzed to give the hold up of each size particles. If W_1 and W_2 denote the hold-ups of particle sizes 1 and 2 and w_1 and w_2 denote the feed rates of these particle sizes, then mean residence times of particles 1 and 2 were calculated by the following equations:

$$\bar{t}_1 = W_1 / w_2 \qquad \text{(eq. 5.4)}$$

$$\bar{t}_2 = W_2 / w_2 \qquad \text{(eq. 5.5)}$$

The ratio of $\frac{\bar{t}_2}{\bar{t}_1}$ has been defined as the hold up ratio $H(2,1)$ and can be found by the equation:

$$H(2,1) = \frac{W_2}{W_1} \quad \frac{w_1}{w_2} \qquad \text{(eq. 5.6)}$$

The minimum fluidizing velocity for the given bed hold up was found by first calculating average diameter of particles in the bed from Equation 5.3 and then calculating u_{mf} by the following equation:

$$(u_{mf})_{av} = (u_{mf})_1 \quad dp_{av}/dp_1 \qquad \text{(eq. 5.7)}$$

In table E-1, Appendix E, U/u_{mf} have been tabulated with $H(2,1)$. Also given in the table are three time parameters:

$$\bar{t} = \frac{W}{w} \qquad \text{(eq. 5.8)}$$

$$T_A = G\,\bar{t}/V_b \qquad \text{(eq. 5.9)}$$

$$T_B = G\,\bar{t}/V_s \qquad \text{(eq. 5.10)}$$

Where V_b and V_s are bed volume and volume of solids in the bed, and T_A and T_B are dimensionless times. The usefulness of defining and tabulating $\bar{t}$, T_A and T_B will be discussed later in the next chapter on discussion of results.

5.4.3 <u>Study of other Variables</u>

The effect of changing the total bed hold-up by changing the static bed height on the hold-up ratios was studied for three more bed hold-ups of 120 g, 320 g and 620 g corresponding to static bed heights of 5 cm, 10 cm and 20 cm respectively. In each case, runs were taken for the same eleven air flow rates and these results are summarized in Tables E-2, to E-4, Appendix E. Experiments for static bed heights of 2.5, 5.0 and 10 cm were repeated for total solid feed rates of 180 g/min. Tables E.5 to E-7, Appendix E, give the results for the above solid feed rates for static bed height of 2.5 cm. Similarly, Tables E-8 to E-10, Appendix E give the results for a static bed height of 5 cm and Tables E-11 to E-13, Appendix E, for a static bed height of 10 cm.

Keeping the total solid feed rate at a level of 300 g/min. the size distribution of particles were also varied in the feed stream. For a static bed height of 2.5 cm, two feed size distribution were tried. In .the first run, 0.393 mm particles and 0.724 mm particles were mixed in the feed in ratio of 4:1 where as in the second run, ratio was maintained at 1:4. The results of these runs are given in Tables E-14 and E-15. Tables E-16 and E-17, Appendix E, give the results of similar runs for a static bed height of 5cm.

The size combination of middle (0.724 mm) and large size (0.96 mm) particles in the feed was done in the same manner as was done for small (0.393 mm) and middle size particles. The air flow rates used here were ranging from 108 to 325 litres per minute at S.T.P. Tables E.18 to E.21, Appendix E, summarize the data for four static bed heights of 2.5 cm, 5 cm, 10 cm. and 20 cm. at a total solid feed rate of 180.0 g/min. consisting of 50 percent each of 0.724 mm and 0.96 mm particles. For the static bed heights of 2.5 and 5 cm two more solid feed rates were tried and these results are given in Tables E-22 to E-25, Appendix E. For static bed heights of 2.5 cm, feed size distribution ratios were also changed to 4:1 and 1:4 and the results for these runs are given in Tables E.26 and E.27, Appendix E.

A few runs were tried by using the sharp cut combination of small (0.393 mm) and large (0.96 mm) size particles in the feed. It was observed that the fluidization was difficult under these conditions because the larger particles had a tendency to settle at the bottom of the bed while the smaller particles were getting entrained. Discussion on these runs has been omitted because feed containing such sharp cuts are unlikely to be used in any practical situation.

5.4.4 Feeds of Three Different Sizes

More important from practical point of view were the fluidized beds with solid feeds having continuous variations in sizes. Experiments with solid feeds containing particles of many sizes were difficult to perform and hence mixed feeds prepared from three available sizes were tried. Runs were taken by feeding 0.393 mm, 0.724 mm and 0.96 mm particles in equal proportions (1:1:1) and the results are summarised in Tables F-1 to F-3, Appendix F. for three static bed heights of 2.5 cm, 5 cm and 10 cm. For all these runs the total solid feed rate was kept at 180 g/min. and the air flow rate was varied from 81 to 215 l/min.

Use of three particle sizes in solid feed can give three combinations of hold up ratios and they are designated in Tables as H(2,1), H(3,1) and H(3,2). u/u_{mf} was again calculated after determining the average diameter of particles in the bed.

Table F-4 and F-5, Appendix F, give experimental results for a total solid feed rates of 300 and 500 g/min at static bed heights of 2.5 cm and Tables F-6 and F-7, Appendix F, give similar data for a static bed height of 5 cm.

Four more sets of runs were carried out by changing the size distribution in feed. Tables F-8 to F-11, Appendix F, summarize the results obtained for a total solid feed rate of 300 g/min with feed containing particles of three sizes in 4:1:1, 1:1:4, 1:4:1 and 4:1:4 proportions.

5.5 AXIAL AND RADIAL DISTRIBUTION OF SOLIDS IN FLUIDIZED BEDS

In an attempt to understand the phenomena affecting the solid flow inside the bed as to give different residence times for particles of different sizes, some experiments were carried out to measure the radial and axial distributions of particle sizes at different conditions. Although these experiments were not designed to give exact distribution of particle sizes inside the bed, yet they were useful to provide a qualitative picture of what was happening inside the bed.

5.5.1 EFFECT OF SIZE OF OUTLET FOR SOLIDS

It was considered reasonable to assume that the particles of different sizes experience different residence times in the fluidized bed because the particles were not distributed uniformly throughout the bed. More important was the distribution of solids at the periphery of the bed because the outlet connection for solid discharge from the bed was located at the periphery. The first set of runs was taken to study the effect of the size of the outlet for solid discharge. This was done very simply by placing inside the outlet opening of 13 mm diameter a tight fit polythene tube with smaller inner diameter. The end of the polythene tube touching the bed periphery was carefully cut to the shape of an arc to match very closely with the inner surface of the main fluidizing column. For two more outlet sizes the runs were taken at solid feed rate of 120g/min using 0.393 and 0.724 mm diameter particles in 1:1 ratio. The results of these experiments are given in Table G-1, Appendix G.

5.5.2 Down-Comer and Over Flow System for Solids

Most industrial fluidized beds with continuous solid feed use an overflow for solid discharge. The solid feed is sent through a down-comer and is taken out through an overflow pipe. Such system has the advantage of an automatically controlled bed height. The experimental set up was slightly modified as discussed in Section 4.2.2.1 to provide a downcomer and overflow system. Measurements for hold up ratios were made for mixed feeds of three sizes of 1:1:1 proportion.

Data was collected for two solid feed rates of 180 and 500 g/min. and the results are given in Table G-2, Appendix G.

5.5.3 Position of Outlet for Solid Discharge

The influence of axial position of outlet connection was studied by providing another outlet connection at a height of 5 cm above the sintered glass disc.

This outlet connection was operated in the following two ways:

(i) Total solid hold up in the bed was controlled automatically by using the outlet connection as an overflow system.

(ii) Total solid hold up in the bed was regulated by regulating exit solid feed rate with the help of two way stop cock.

The arrangement for sampling was kept the same as in the earlier outlet connection.

Table G-3, Appendix G gives the results for these runs for a feed size distribution of 1:1 for particle sizes of 0.393 and 0.724 mm.

The effect of the axial distribution of particle sizes on hold-up ratios was determined experimentally by changing the bed hold-ups. For a solid feed rate of 120 g/min consisting of 0.393 and 0.724 mm particles in 1:1 ratio, experiments were conducted to determine the hold-up ratios for different bed heights and u/u_{mf} values. These results are presented in Table G-4, Appendix G. The effect of radial distribution of particles near the bed wall on the hold-up ratios was investigated by changing the radial position of the outlet for solids as discussed in

Sec. 4. 4. 2. 2 The inward radial distance for solids outlet from the outer periphery of the bed, that is, fluidizing column wall, could be easily varied by inserting a tightly fitting polythene tube of proper length inside the bed through outlets provided in the column. Using such an arrangement the data was taken for total bed hold-up of 320 g and a total solid feed rate of 120 g/min consisting of 0.393 and 0.724 mm diameter particles in 1:1 proportion. The results are presented in Table G-5, Appendix G.

5. 5. 4 Radial and Axial Distribution by Concentric Tubes Insert

In order to obtain a better understanding of the radial and axial distribution of sizes in the fluidized bed operating at steady state, a concentric tubes insert as shown in Fig. 4. 8 was used to arrest particles in their radial and axial positions. In these experiments a batch of solids was used to arrest particles in their radial and axial positions. In all these experiments a batch of solids was fluidized and the concentric tubes insert was dropped into the bed and simultaneously the air supply was completely cut-off. The particles in the bed were entrapped in their radial positions inside the compartments formed by concentric tubes of the insert. The particles were withdrawn by suction from top in layers from each compartment. Assuming that the particles did not change their relative axial position in the short time available for settling after the air supply was cut off, the hold-up ratios obtained by this technique provided a good picture of the radial and axial distribution of particle sizes in the bed. The hold up ratios for these runs are tabulated in Table G-6, G-7 and G-8, Appendix G, for particle sizes 0.393 and 0.724 mm in 1:1 proportion, 0.724 and 0.96 mm in 1:1 proportion, and 0.393, 0.724 and 0.96 mm in 1:1:1 proportion respectively.

5.6 EFFECT OF BAFFLES ON HOLD UP RATIOS IN FLUIDIZED BEDS

As discussed in the next chapter on the discussion of results, the study of axial and radial distributions helped in the understanding of the phenomena of the particle size distribution within the fluidized beds. It was, therefore, reasonable to expect that the introduction of baffles in the bed may change hold-up ratios by making the bed probably more uniform. This was verified by using basically two different types of baffles which have been described in Section 4.5.

The results of observations with the sieve plate type baffles with low spacing in between the baffles are given in Table H-1, Appendix H. The results obtained on the same type of baffles with a larger spacing in between the baffles are tabulated in Table H-2, Appendix H. These experiments were carried out with a total solid feed rate of 180 g/ min consisting of equal amounts of all the three sizes for two bed hold-ups of 90 and 160 grams. The use of Davis type of baffles (the transvers baffles) gave better results. Experimental results with Davis baffles for a total solid flow rate of 180 g/min consisting of all three sizes in equal proportion for three bed hold-ups of 90, 160 and 300 g. are given in Tables H-3 to H-5, Appendix H. For all these bed hold-ups, two more solid feed-rates were used. In Tables H-6 and H-7, Appendix H, data for feed rates of 300 and 500 g/min is given for a bed hold-up of 90g. Similar data for bed hold-up of 160 and 320 g are given in Tables H-8 and H-9, and Tables H-10 and H-11, Appendix H, respectively.

CHAPTER 6

DISCUSSION OF RESULTS

It has been indicated in Chapters 1 and 3 that this investigation was undertaken to study the flow behaviour of solids in continuous moving and fluidized beds with particular reference to mixed sizes of solids in the feed. These investigations are divided basically in four parts as mentioned in Chapter 5. The results of experimental investigation presented in Chapter 5 and Appendices A to H are discussed in this Chapter.

6.1 EQUIPMENT STANDARDIZATION

The minimum fluidizing velocity was experimentally determined by plotting pressure drop across the bed as a function of superficial air flow rate. The data given in Tables A.1 to A.3 Appendix A are plotted in Fig. 6.1. The minimum fluidizing velocity was also determined experimentally by plotting, Figs. 6.2 and 6.3, from the data given in Tables A.4 and A.5 for mixed feeds prepared from ballotini beads of three sizes in different proportions. The air flow rates corresponding to the minimum fluidizing conditions as determined from Figs. 6.1 to 6.3 are given in Table 6.1.

For 0.393 mm size particles, the minimum fluidizing velocity as calculated from the equation given by Leva (59) comes to 16.3 cm/sec. Details of these calculations are given in Section A-1, Appendix A. This value corresponds to an air flow rate of 19.8 litres

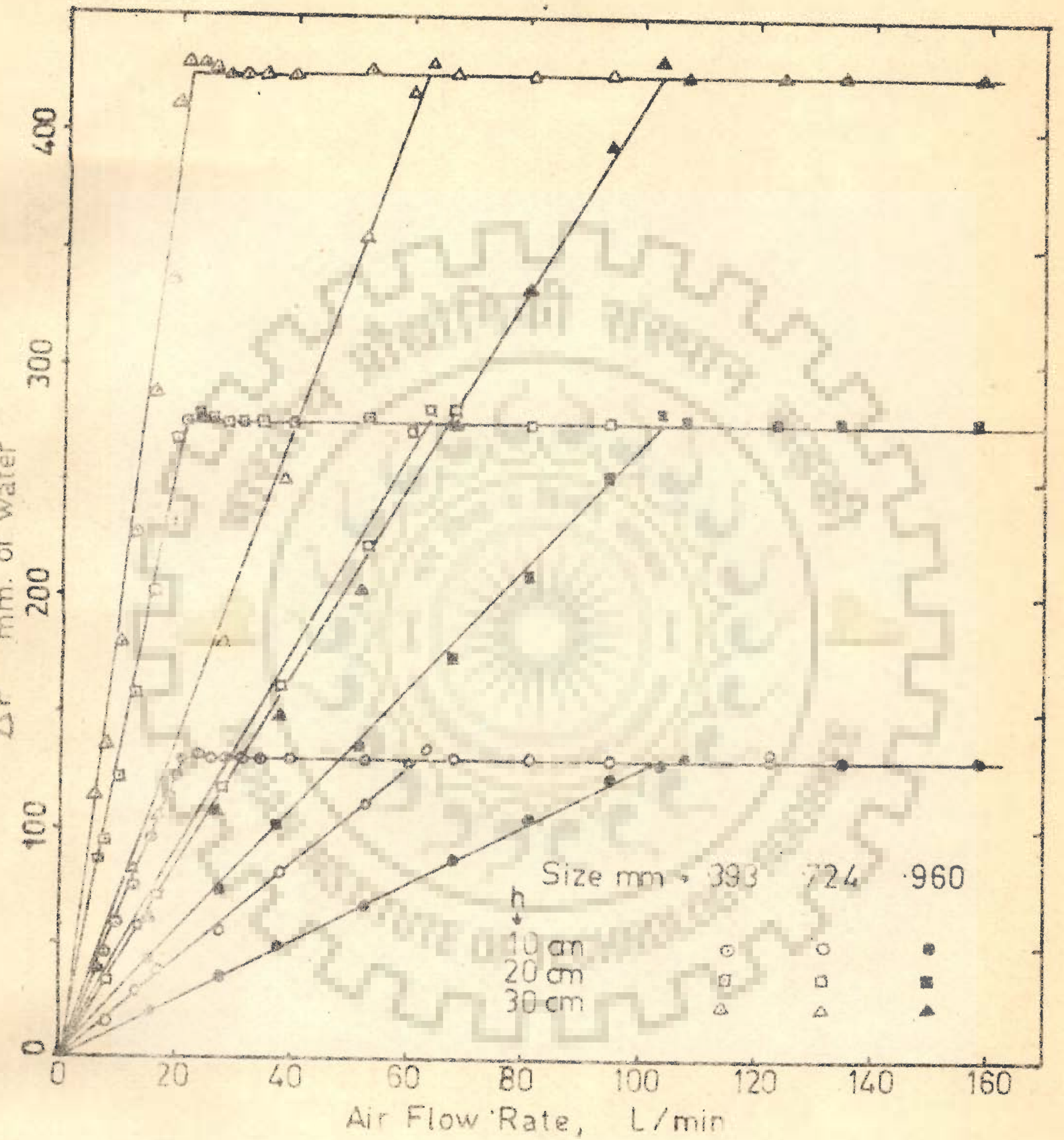


FIGURE 6·1 DETERMINATION OF G_{mf} FOR BALLOTINI BEADS

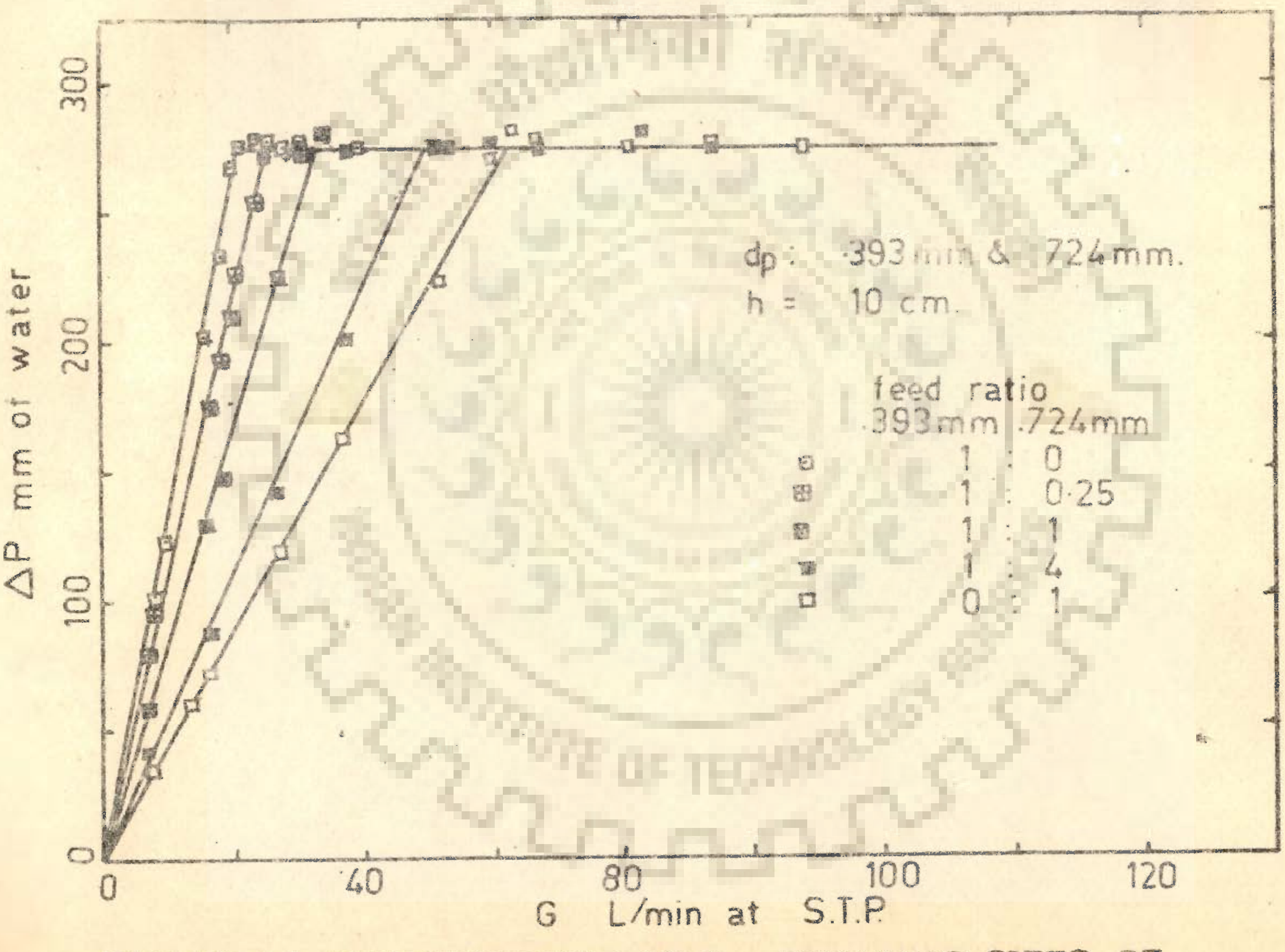


FIGURE 6.2 DETERMINATION OF G_{mf} FOR MIXED SIZES OF BALLOTINI BEADS

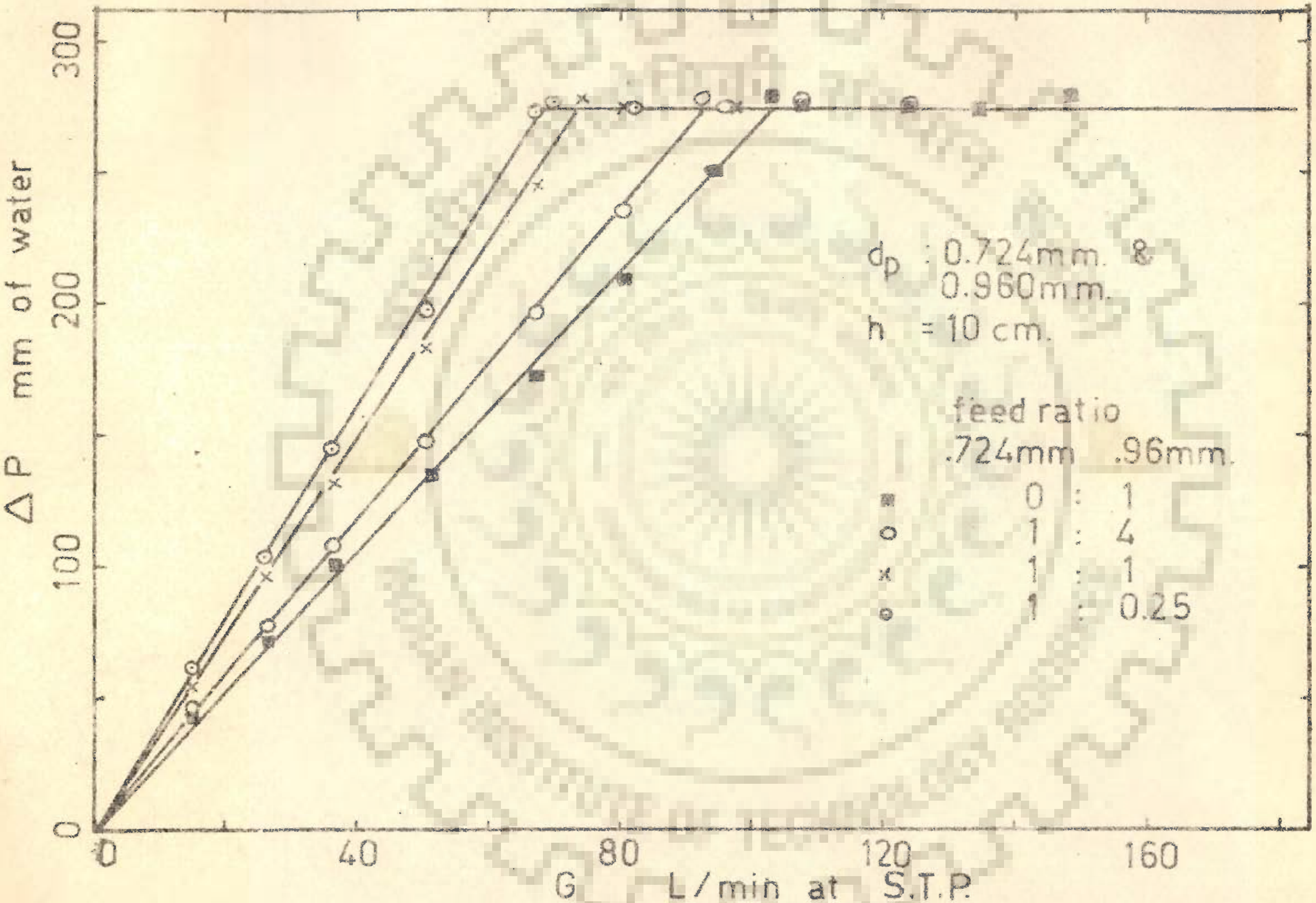


FIGURE 6.3 DETERMINATION OF G_{mf} FOR MIXED SIZES OF BALLOTINI BEADS

Table 6.1

MINIMUM FLUIDIZING AIR FLOW RATES

(From Fig. 6-1, Fig. 6-2 and Fig. 6-3)

S. No.	Mass fraction in bed of 0.393 mm	0.724 mm	0.96mm	Average dp mm	G_{mf} in L.P.M.
1.	1.0	-	-	0.393	20.4
2.	0.80	0.20	-	0.431	24.2
3.	0.50	0.50	-	0.508	32.8
4.	0.2	0.80	-	0.700	59.0
5.	-	1.0	-	0.724	62.4
6.	-	0.8	0.2	0.755	68.3
7.	-	0.5	0.5	0.886	92.7
8.	-	0.2	0.8	0.886	92.7
9.	-	-	1.0	0.96	130.0

per minute at S. T. P. and it compares very well with the experimental values of 20.4 litres per minute. In order to obtain a relationship between dp and u_{mf} the values given in Table 6.1 are plotted in Fig. 6.4 on a log-log graph paper. All the points fall in a straight line with a slope of 1.82. Correlation suggested by Leva (59) also uses the value of exponent on particle diameter as 1.82. It is, therefore, concluded that the experimental set up and procedure is quite accurate and the results can be used with confidence.

6.2 RESIDENCE TIMES OF SOLIDS IN MOVING AND FLUIDIZED BEDS

The work done by Morris et. al (78), Yagi and Kunii(112) and Kamiya (48) on the residence times of solids in moving and fluidized beds has been discussed in Chapter 2. These investigations have used solid feed of single size only. In the present investigations solid feeds of mixed sizes were also used in addition to single size feeds used by earlier workers. The results on residence times of solids in moving and fluidized beds using stimulus response technique are discussed in sections 6.2.1 to 6.2.4.

6.2.1 Moving Bed with Single Size Feed

The experimental runs were planned factorially and the test of reproducibility was done at the centre of the factorial cube, that is, the conditions of variables were set at the average value of the two extreme values for the variables. At u/u_{mf} of 0.5 and w of 100 g/min, the results of three replicate runs taken for 0.724 mm

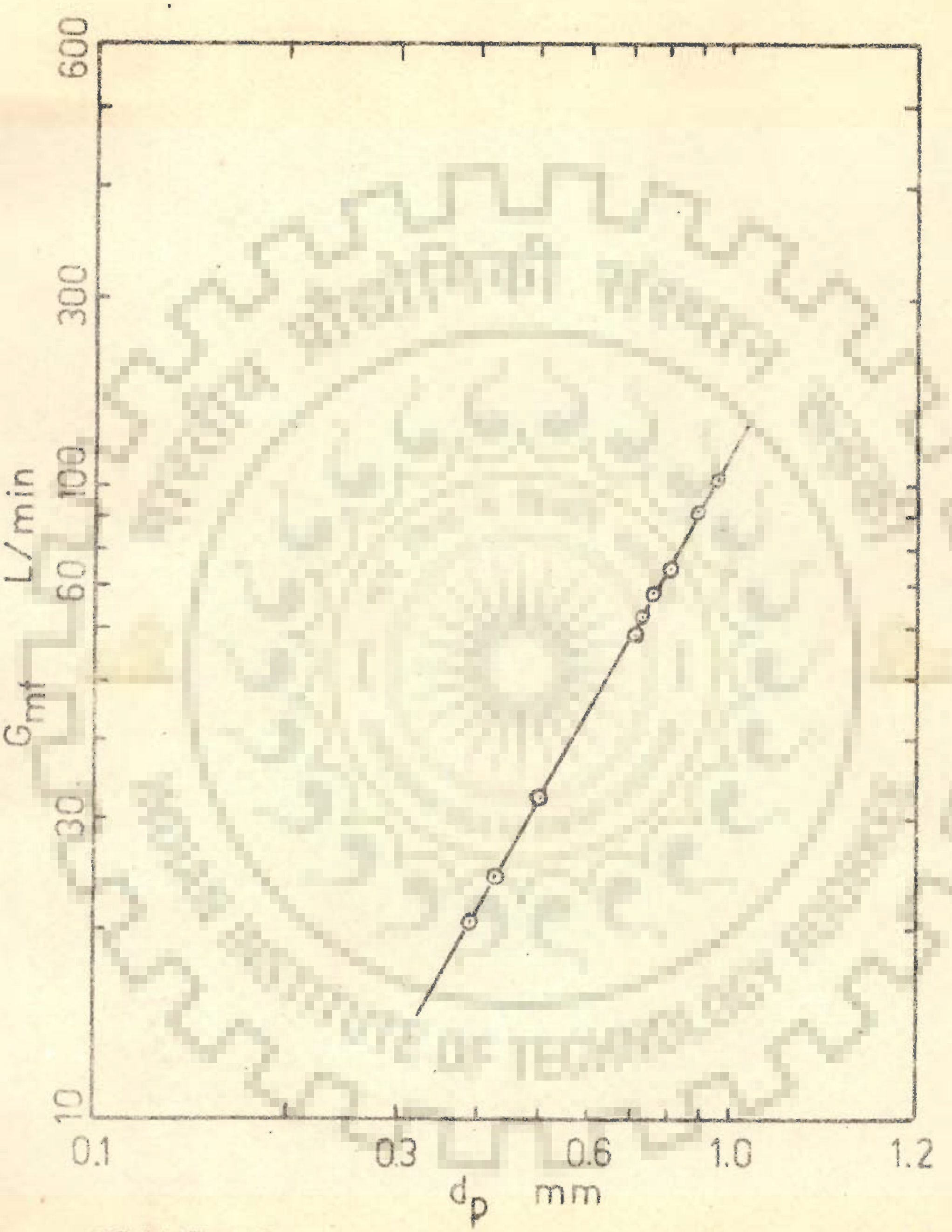


FIGURE 6.4 CORRELATION OF G_{mf} WITH d_p

particles are plotted in Figure 6.5. From the figure it is clear that most of the points fall on the two curves. Considering the fact that the response of tracer-inputs is not always smooth because of the probability error, the reproducibility of these runs can be taken as extremely satisfactory. For a quantitative estimate of the reproducibility the dispersion number as defined by Danckwerts (22) was calculated using IBM 1620 computer by the method of weighted moments. Theoretical analysis of the tracer runs is given in literature (3,4,80), the programme used for calculation is given in Appendix I. The computed values of dispersion numbers are -0.00010, 0.00012 and + 0.00009. These values of dispersion number are so low that they indicate completely plug flow behaviour. The negative dispersion numbers indicate that the output curve displays a closer plug flow behaviour as compared to the input curve. Even by visual observation it appears that the output curve is closer to plug flow behaviour than the input curve. This may be attributed to the slight disturbance of the bed while adjusting the signal tracer at 10 cm height. From the replicate runs it can be easily inferred that the flow pattern and solids in a moving bed is like a plug flow and the tracer runs are very reliable.

The data obtained on tracer experiments using the extreme values of air flow rates corresponding to u/u_{mf} values of 0.2 and 0.95 and different bed hold-ups and solid feed rates, Tables B-4 to B-9, Appendix B, is presented in Figs. 6.6 to 6.11.

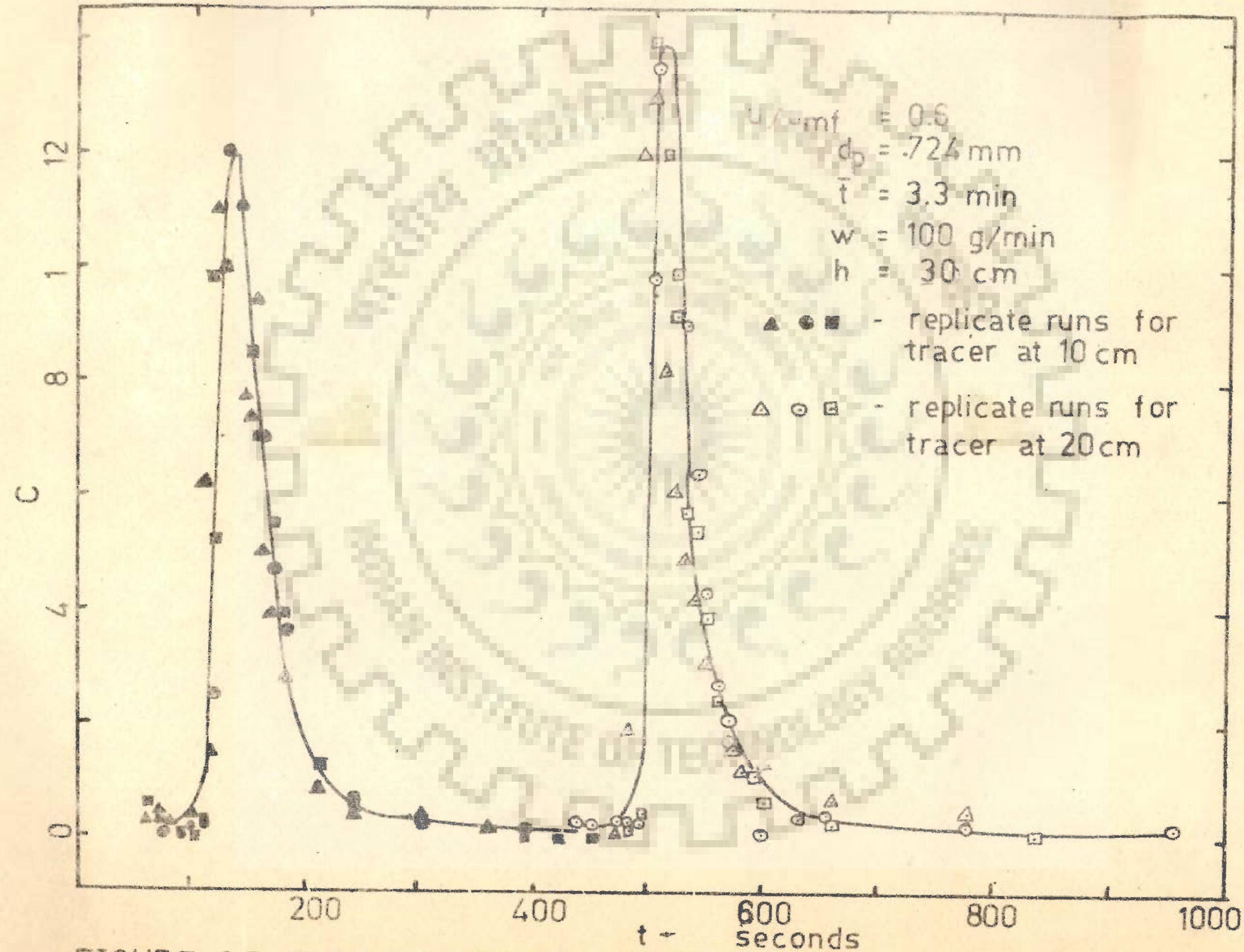


FIGURE 6.5 REPRODUCIBILITY TEST IN MOVING BED RUNS

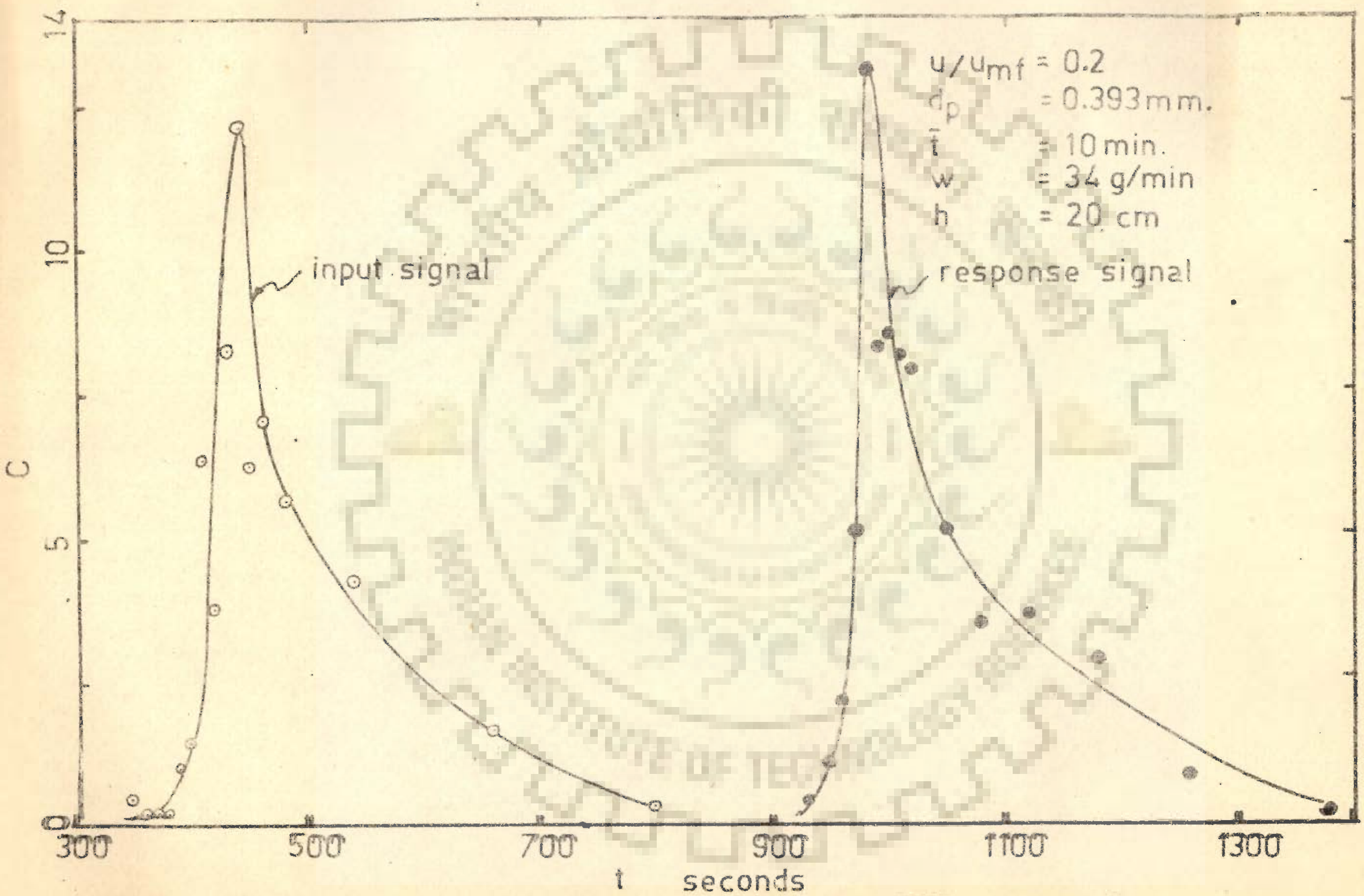


FIGURE 6.6 STIMULUS RESPONSE IN MOVING BED. RUN 1

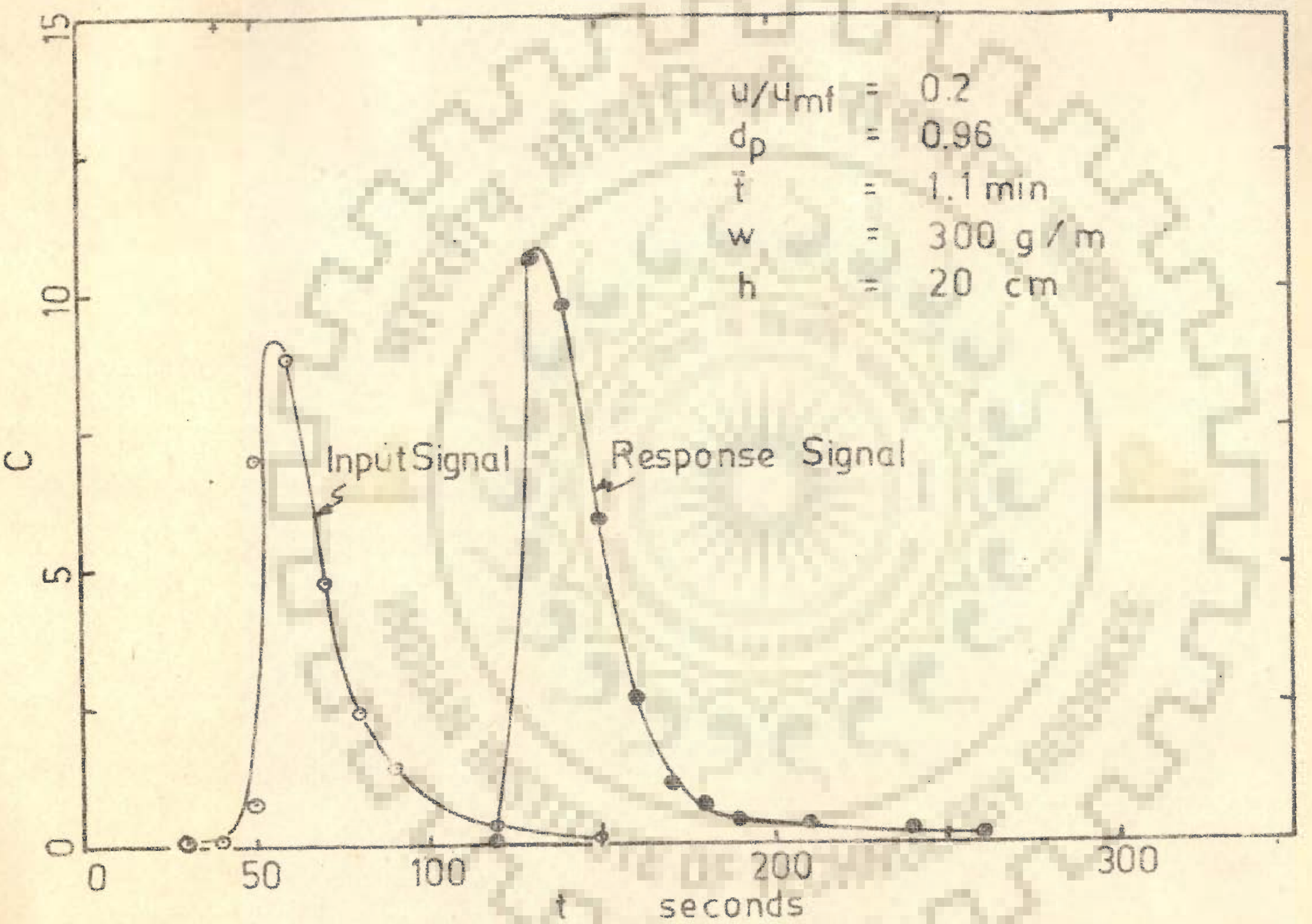


FIGURE 6.7 STIMULUS RESPONSE IN MOVING BED, RUN 2

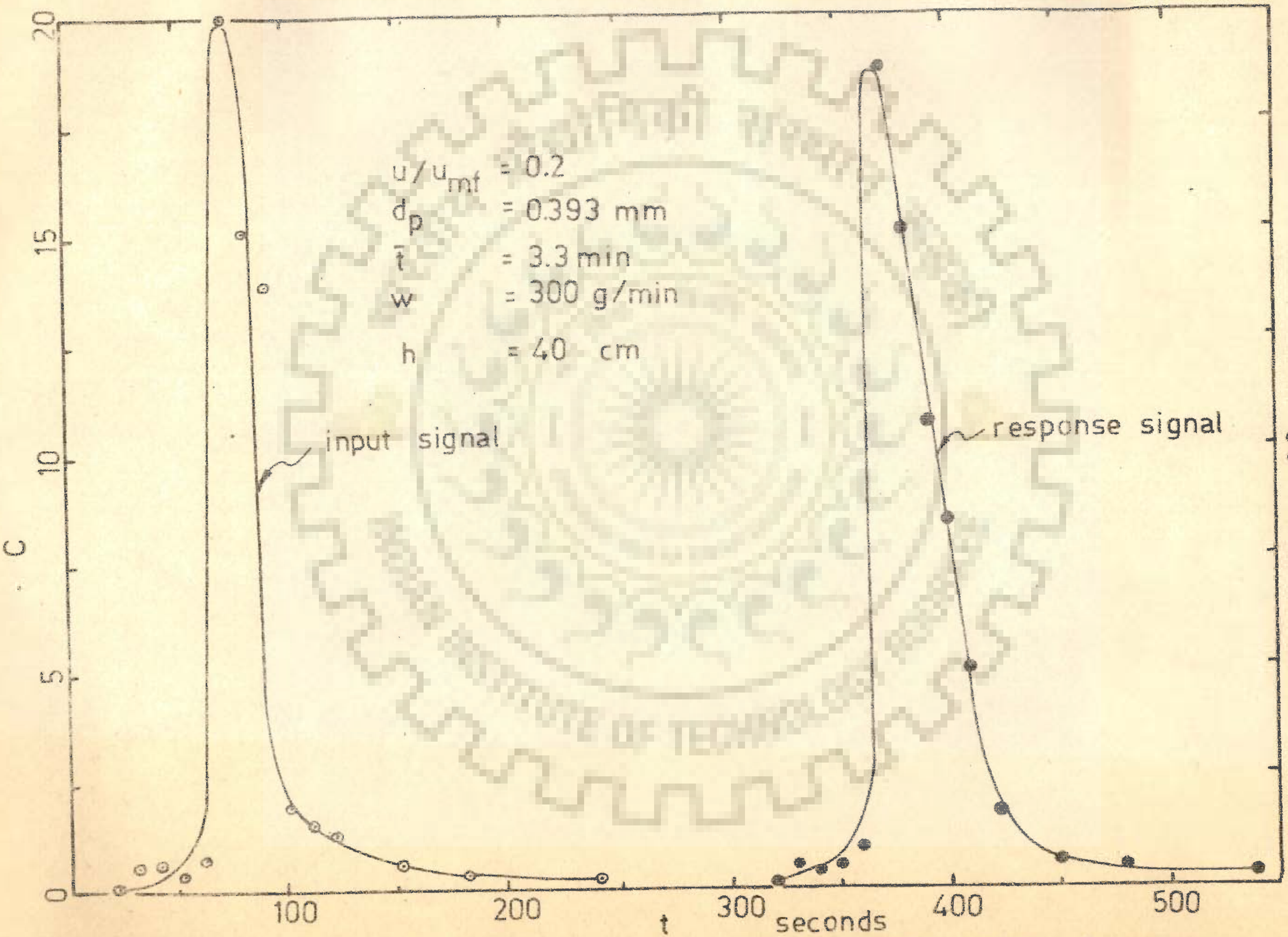


FIGURE 6.8 STIMULUS RESPONSE IN MOVING BED, RUN 3

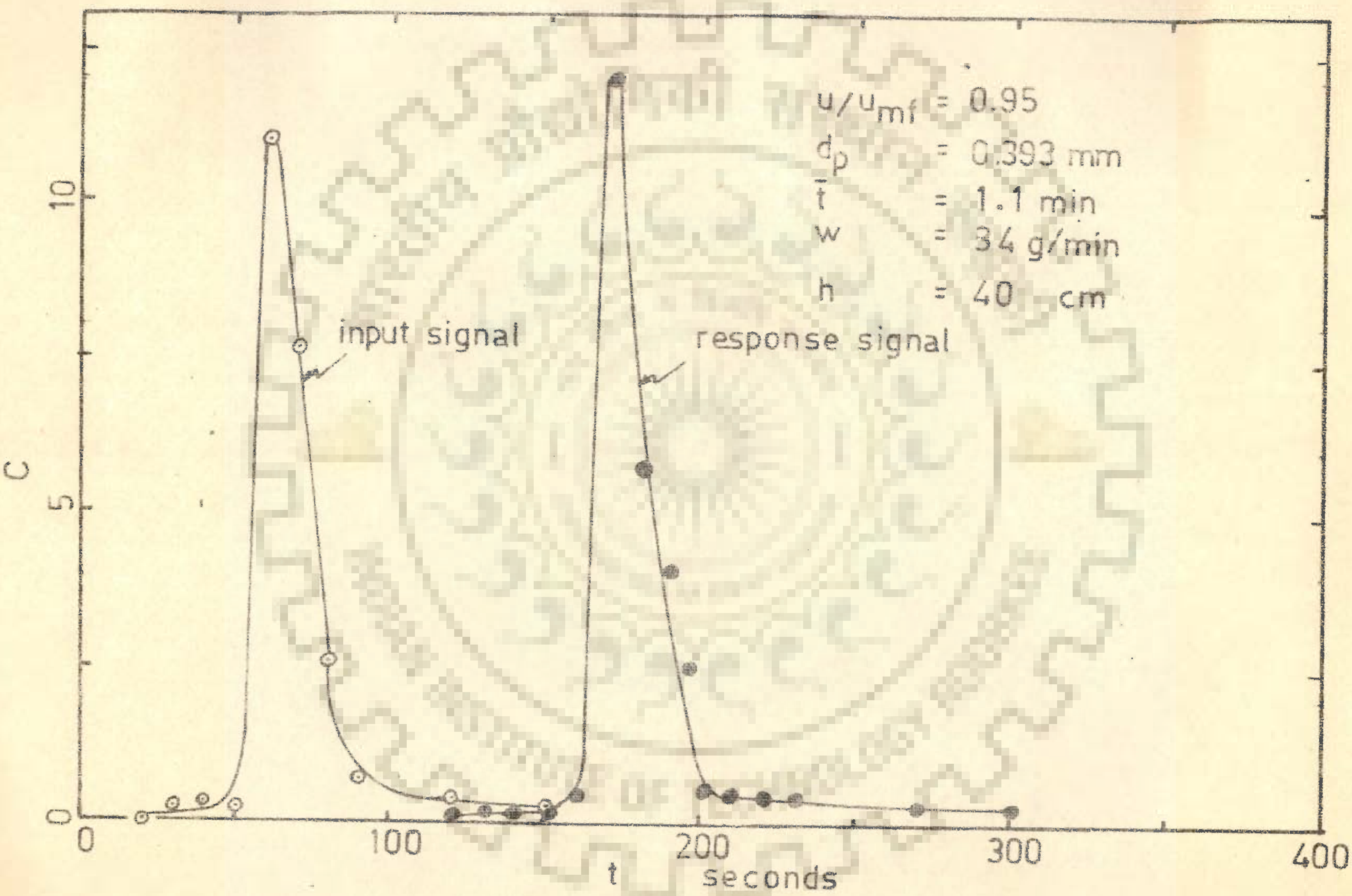


FIGURE 6.9 STIMULUS RESPONSE IN MOVING BED, RUN 5

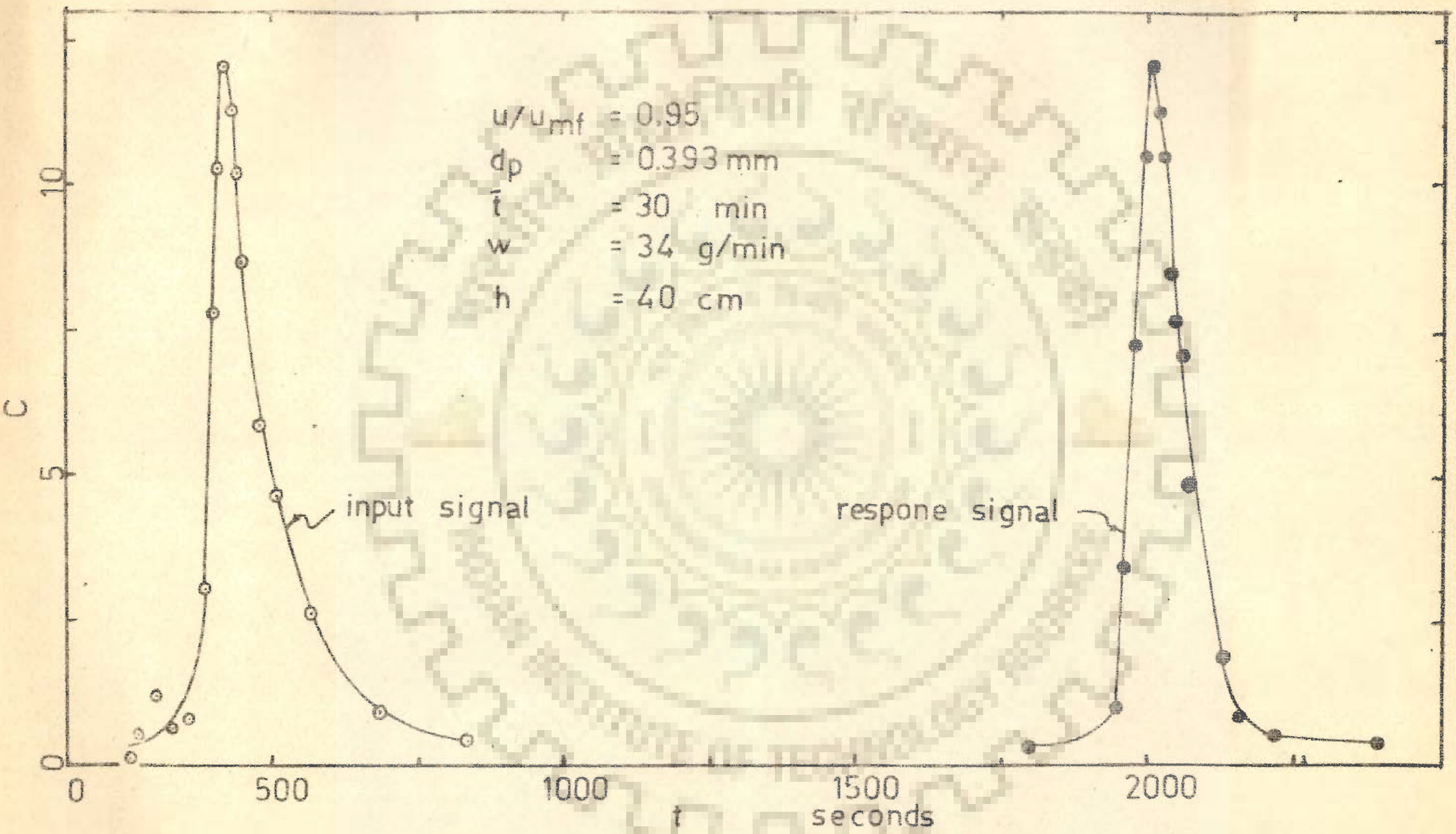


FIGURE 6.10 STIMULUS RESPONSE IN MOVING BED, RUN 7

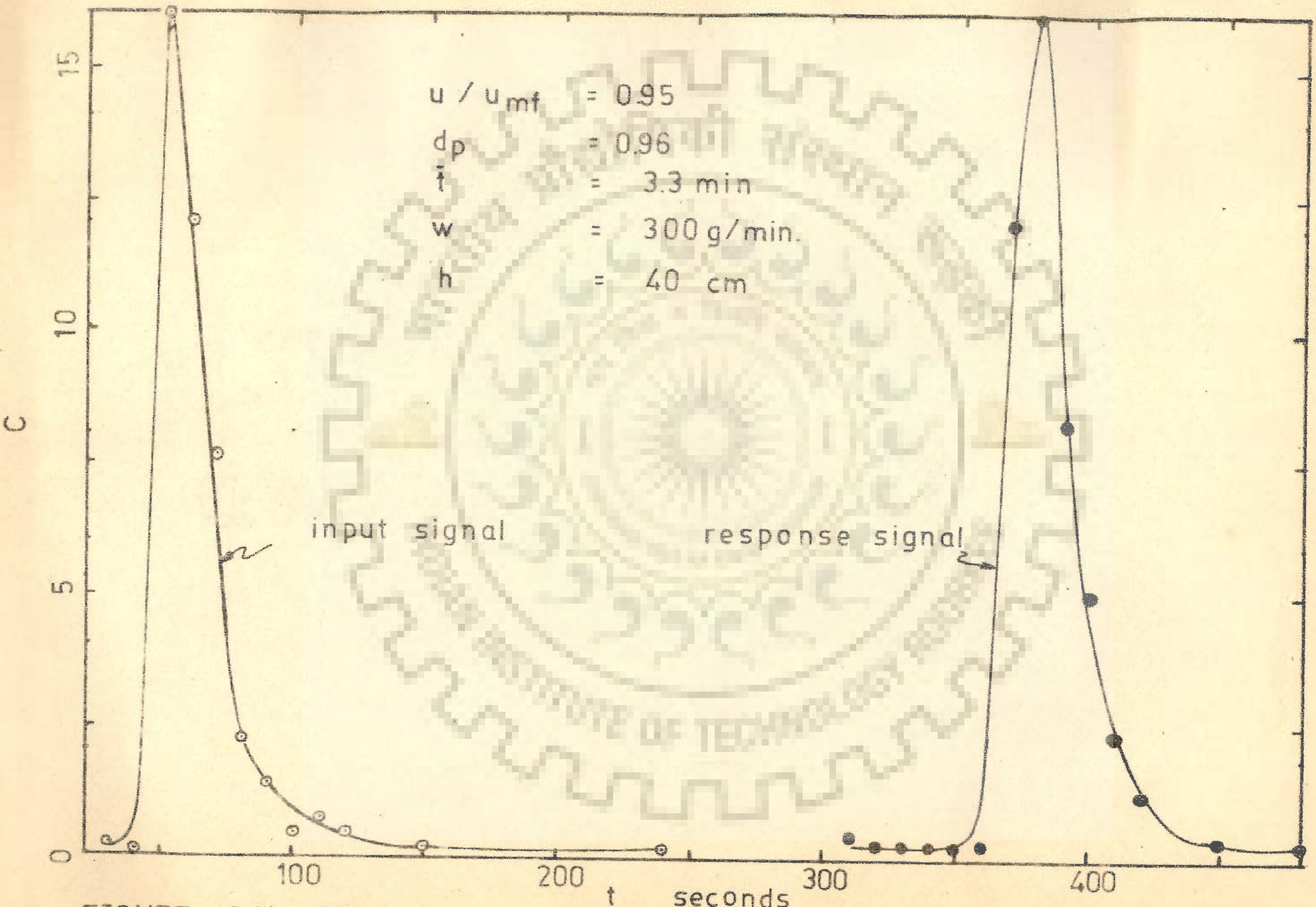


FIGURE 6.11 STIMULUS RESPONSE IN MOVING BEDS, RUN 8

For each experimental conditions used in the tracer experiments, the response curve was also obtained for the pulse injection of the tracer at 10 cm height from the base of the bed to eliminate the end effects by the method of weighted moments. In this method the response of tracer input at 10 cm height was considered as the input signal and the response curve of tracer introduced at a higher height from the base was considered as the response signal. Thus the error due to the end effects stand common in both the input and the output signals and cancel out. In Figs. 6.6 to 6.11, the input and the output signals are plotted as dimensionless tracer concentration C and real time. A visual observation of these figures clearly indicates that both the input and the response curves match very closely in shape. This clearly suggests that the flow pattern of solids in the moving beds for all these conditions was extremely close to plug flow. The deviations from the ideal pulse tracer signal in the input and the output curves are, therefore, totally attributable to the end effects and not to the mixing phenomena in the bed. It is also noteworthy that the input tracer curves can give some idea about the extent of dead zone. For instance in Fig. 6.6, the input tracer peak is reached approximately at 440 sec instead at mean residence time of 600 sec corresponding to a solid flow rate of 34 /min and the bed hold-up of 340 g for a 10 cm bed height.

This gives a dead zone equivalent to approximately 90 g of solids in the bed. Similarly, in Fig. 6.7, the input tracer peak is reached at approximately 56 sec instead of at mean residence time of 68 sec. corresponding to a solid flow rate of 300 g/min and the bed hold up of 340 g for a 10 cm bed height. This gives a dead zone equivalent to approximately 60 g solids in the bed. It was also visually observed that the lower portion of the bed diameterically opposite the outlet for solids normally behaves as a dead zone and the volume of this zone decreased with increasing solids flow rate in conformity with the above calculations.

The tracer data obtained for moving beds was analysed for dispersion number by the method of weighted moments and the dispersion numbers are found to be always lower than 0.001. Such low values of dispersion number for all these runs confirmed that the flow pattern of solids in moving beds with u/u_{mf} as high as 0.95 is extremely close to plug flow. In Fig. 6.12, the response (output tracer) curves for u/u_{mf} of 0.2 and 0.95 from Figs. 6.6 and 6.10 are replotted on dimensionless corrdinates. The mean residence time was calculated for both these curves by subtracting the dead zone from the total bed hold up. Both the curves match quite well indicating that the slight dispersion caused due to the end effects is independent of u/u_{mf} upto a value of 0.95. Experiments were not carried out beyond 0.95 u/u_{mf} because some movement of bed particles was observed which resulted in extremely poor reproducibility. Behaviour of the bed in the gas flow-rate range corresponding to u/u_{mf} between 0.95 and 1.2

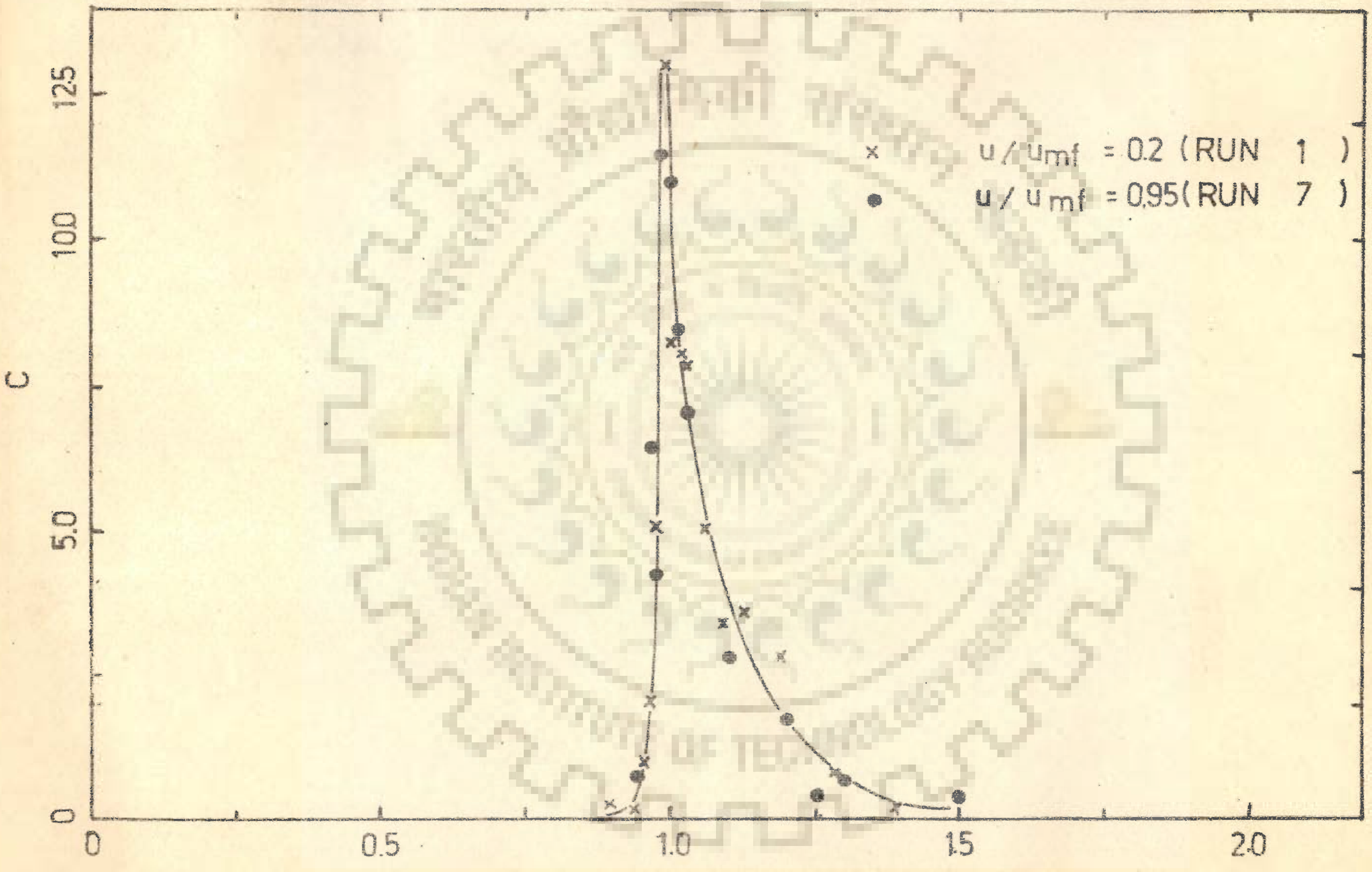


FIG. 6.12 EFFECT OF u/u_{mf} ON RESPONSE CURVES

was not investigated because the behaviour of the bed is in the transition zone and such operation has no practical significance.

Morris et. al (78) had conducted similar experiments on 9 inches bed diameter and their results for $u/u_{mf} = 0.5$ and 9 in. bed height show a considerable departure from the plug flow. For a bed height of 48 inches, the output response matched very closely to the plug flow. The departure from piston flow for 9 inches bed height observed by Morris et. al can be explained by the fact that the solid outlet arrangement used by these workers was quite tortuous. Solids had to pass through a number of bends and valves and as such the end effects in experiments are likely to be much more pronounced than in the present investigation. The results of the present study has clearly established that for all bed heights and for all air velocities upto $0.95\ u_{mf}$, the flow pattern of the solids in the bed can be accurately described by plug flow. The only deviation from the plug flow behaviour is caused mainly due to the end effects in the outlet solid discharge from the bed. Morris et. al have plotted minimum residence time as a fraction of the mean residence time as a function of the bed height and used it as a measure of deviation from plug flow. The suitability of this measure to indicate deviation; from plug flow is open to question since the minimum residence time is dependent on ldead zone and tracer experiments are likely to be influenced by the end-effects depending on the design of outlet for solid discharge . The deviation from plug flow observed by these workers was not very large and they also concluded that the solids movement in moving beds very closely resemble plug flow.

The main conclusions drawn from the present study on moving beds with single particle sizes are summarized below:

(i) For air velocities as high as $0.95u_{mf}$, the flow behaviour of solids in moving beds quite closely resemble the plug flow.

(ii) The tracer experiments on moving beds are likely to be influenced by the end effects depending on the design of outlet for solid discharge.

(iii) There is practically no effect of bed height or mean residence time or particle size on the flow pattern of solids in the moving beds upto u/u_{mf} values of 0.95.

(iv) In the experimental range investigated a dead zone of particles exists in the moving beds and this dead zone is not much influenced by variables like particle size and bed height but increase in solid feed rates tend to decrease the dead zone.

6.2.2 Moving Bed with Mixed Sizes of Feed

After establishing the existence of plug flow for solid movement in the moving bed, the behaviour of solid feed with mixed sizes was investigated. using a feed containing particles of 0.393, 0.724 and 0.96 mm diameter in different proportions. Input and output tracer signals were obtained with tracers of different sizes. Experimental results are given in Tables C-1 to C-3, Appendix C. Of greater interest are the runs taken at the air velocity of $0.95\ u_{mf}$, calculated using average particle diameter for mixed tracer particles and since the velocity was higher than the u_{mf} for small size particles, some movement of these particles was observed. The data of Table C-3, Appendix C, is plotted in Fig. 6.13. From the figure it is clear that the input and output tracer curves for 0.724, 0.96 mm diameter and mixed tracer particle curves get almost super imposed and can be represented by a single curve. For tracer particle size 0.393 mm, however, the curves are drawn separately because of a slightly longer tail. Since the superficial gas velocity used in this experiments is more than u_{mf} of small size particles, it appears that these particles have some tendency to move upwards resulting in slightly longer tail. However, this effect is only marginal and it can be concluded from these experiments that the moving beds exhibit plug flow behaviour even for solid feeds of mixed sizes. The solid particles of all sizes travel in the moving bed in a plug flow even though the superficial gas velocity may be little higher than the minimum fluidization velocity for the smaller particles. The assumption of plug flow for the design of moving bed systems is, therefore, quite adequate even for feeds of mixed sizes. No work has been reported in the literature as yet on the movement of different size particles of a mixed feed in moving bed.

6.2.3 Fluidized Bed with Single Size Feeds

The stimulus-response technique was used to study the solid mixing in fluidized bed with continuous solid flow in the air velocity range of 1.2 u_{mf} to 6 u_{mf}. Similar studies have been made earlier by many workers as ldiscussed in Chapter 2. and the present

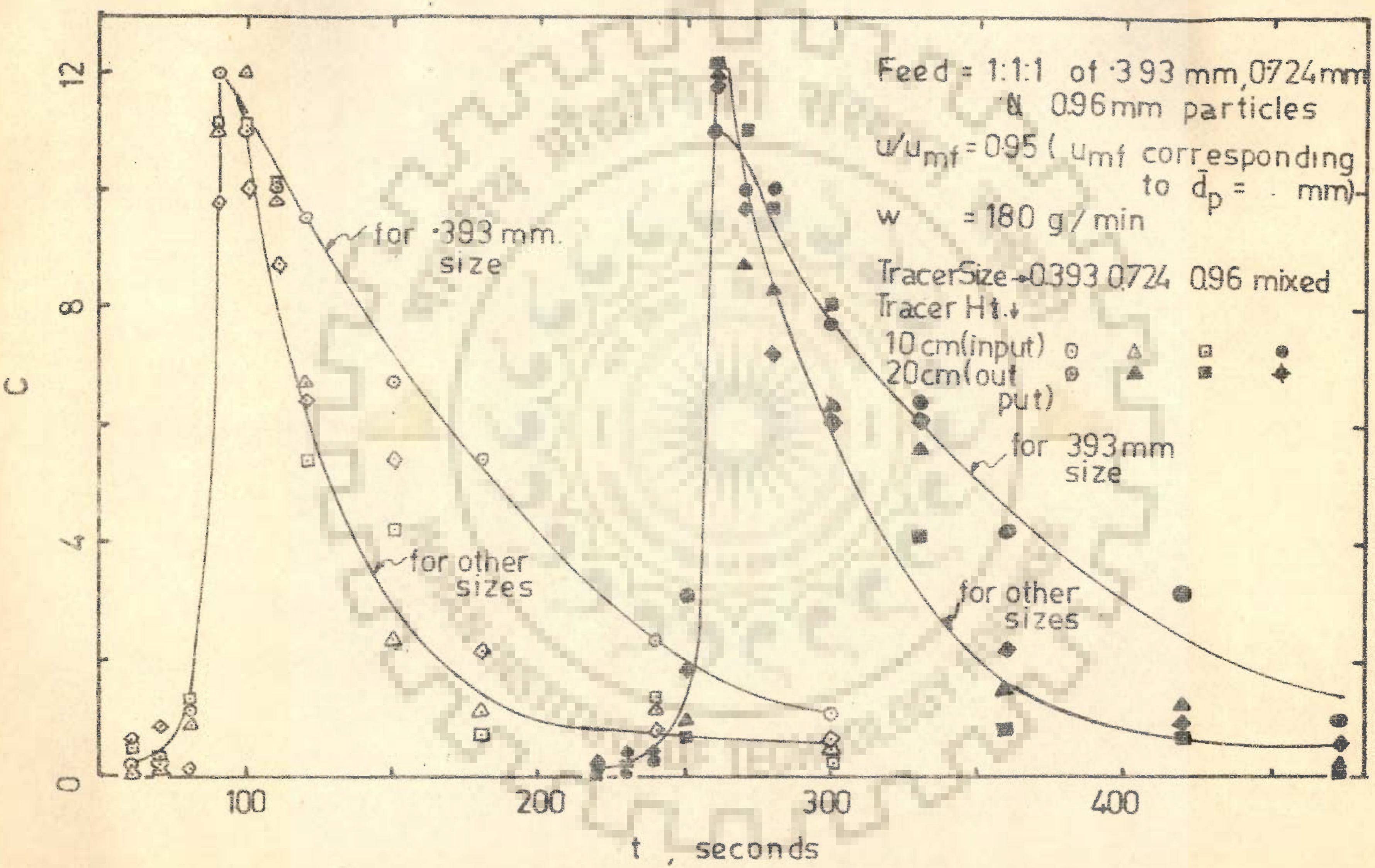


FIG. 6.13 RESPONSE CURVES FOR MOVING BED WITH MIXED FEEDS

investigation is an extension of earlier studies.

Unlike the moving bed experiments, tracer pulse was injected in the fluidized bed at one bed height only because mixing phenomena in a fluidized bed is very rapid and the injection of tracer pulse at two bed heights in some preliminary runs resulted in more or less two superimposed curves. Fig. 6-14 shows the results of one such typical experiment for u/u_{mf} of 1.5. From this data it is clear that the mixing was quite intense in a fluidized bed even at velocities as low as 1.5 u_{mf} and the influence of end effects on tracer experiments in a fluidized bed is not as significant as in the case of moving bed. It was, therefore, decided to use only a single point pulse tracer injection technique for the stimulus - response studies in the fluidized beds. The results of four runs, designed factorially are given in Tables D-1 to D-4, Appendix D, and shown in Figs. 6.15 to 6.17. Figs. 6-15 and 6-17 give three tracer response curves along with their corresponding ideal back mix flow curves at a velocity of 1.2 u_{mf} for different values of bed height, solid feed rate, particle size and the mean residence time. A careful study of these curves reveals that for t values of greater than $\bar{t}$, experimental response curves match the curves for ideal back-mix flow. For t values of less than $\bar{t}$, deviation from backmix behaviour is significant because the mixing of tracer particles at low air velocity of 1.2 u_{mf} is not very rapid to show perfect backmix behaviour immediately after the tracer-injection. For higher air velocity of 6 u_{mf} the response of pulse tracer input is shown in Fig. 6.17 along

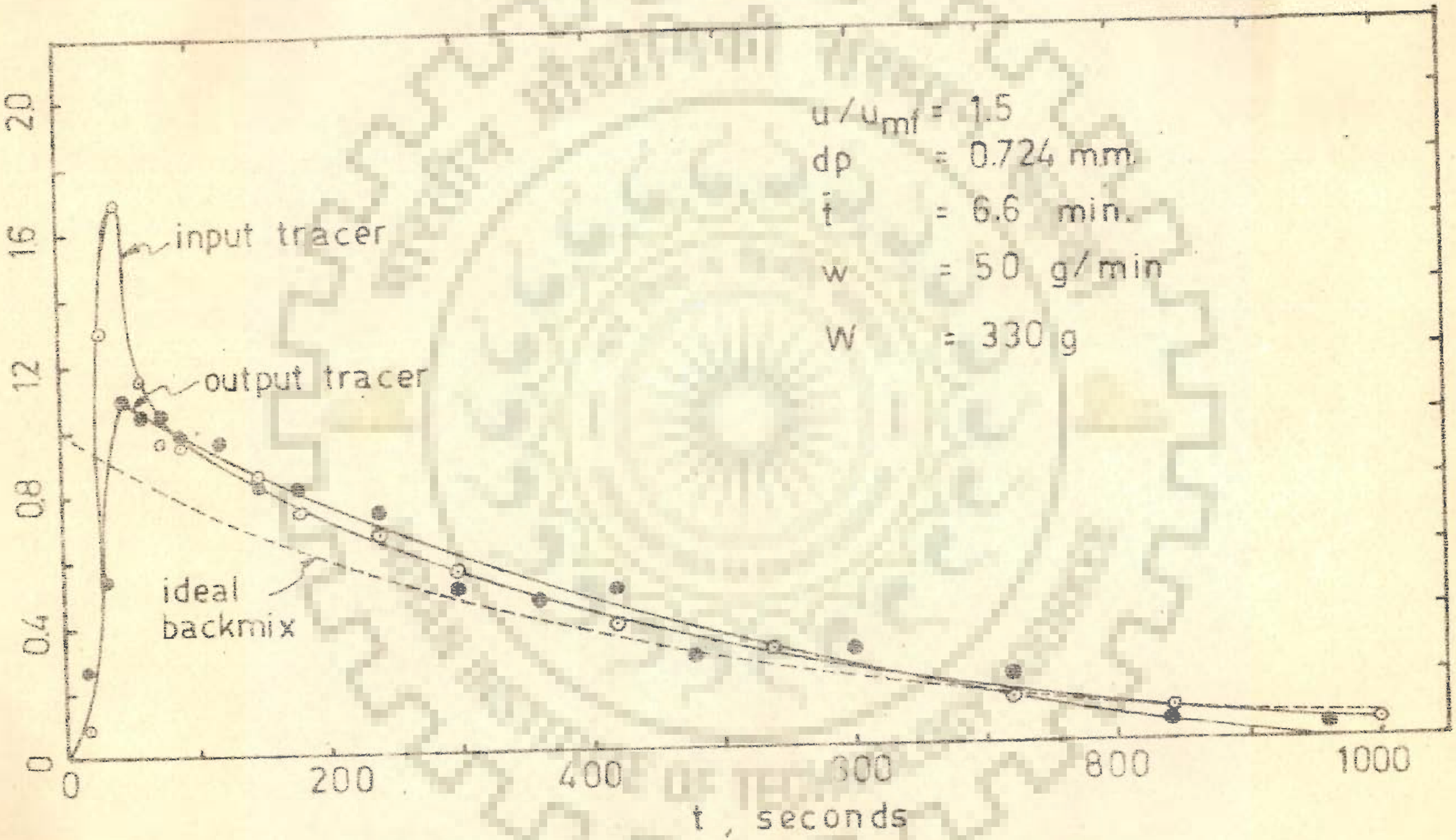


FIG. 6.14 STIMULUS RESPONSE IN FLUIDIZED BED
(use of double tracer injection)

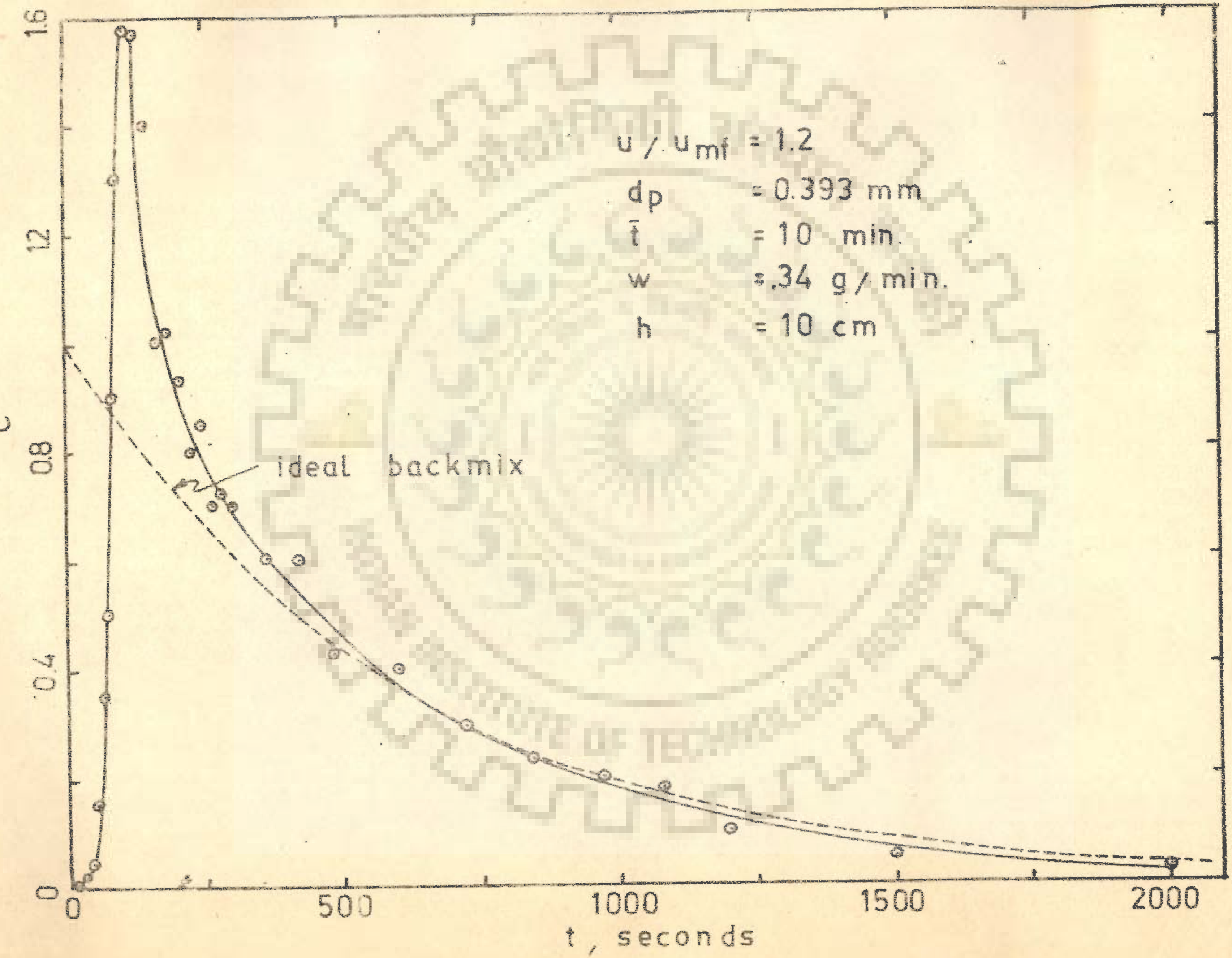


FIG. 6.15 STIMULUS RESPONSE IN FLUIDIZED BED, RUN 1

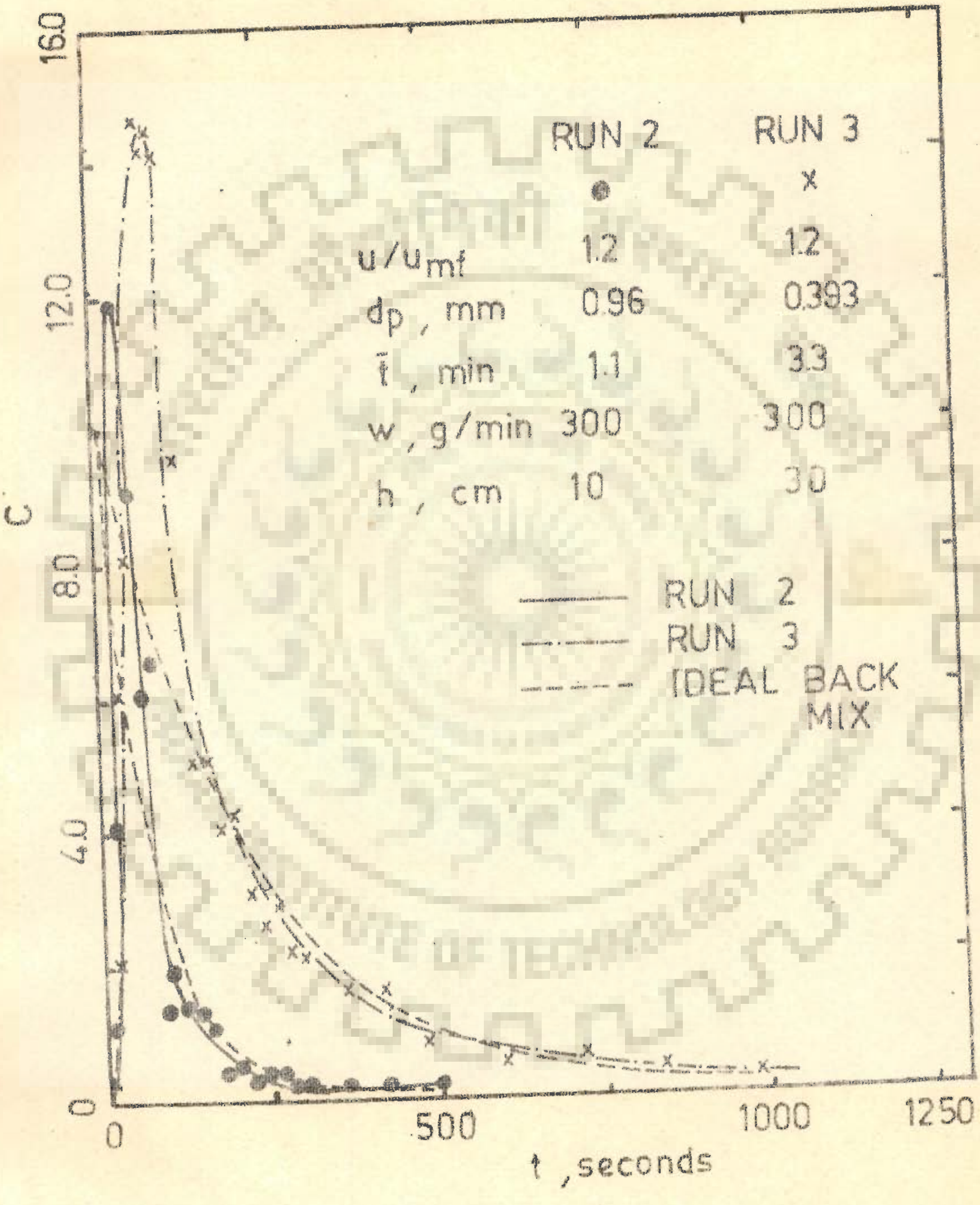


FIG. 6.16 STIMULUS RESPONSE IN FLUIDIZED BED RUNS 2 & 3

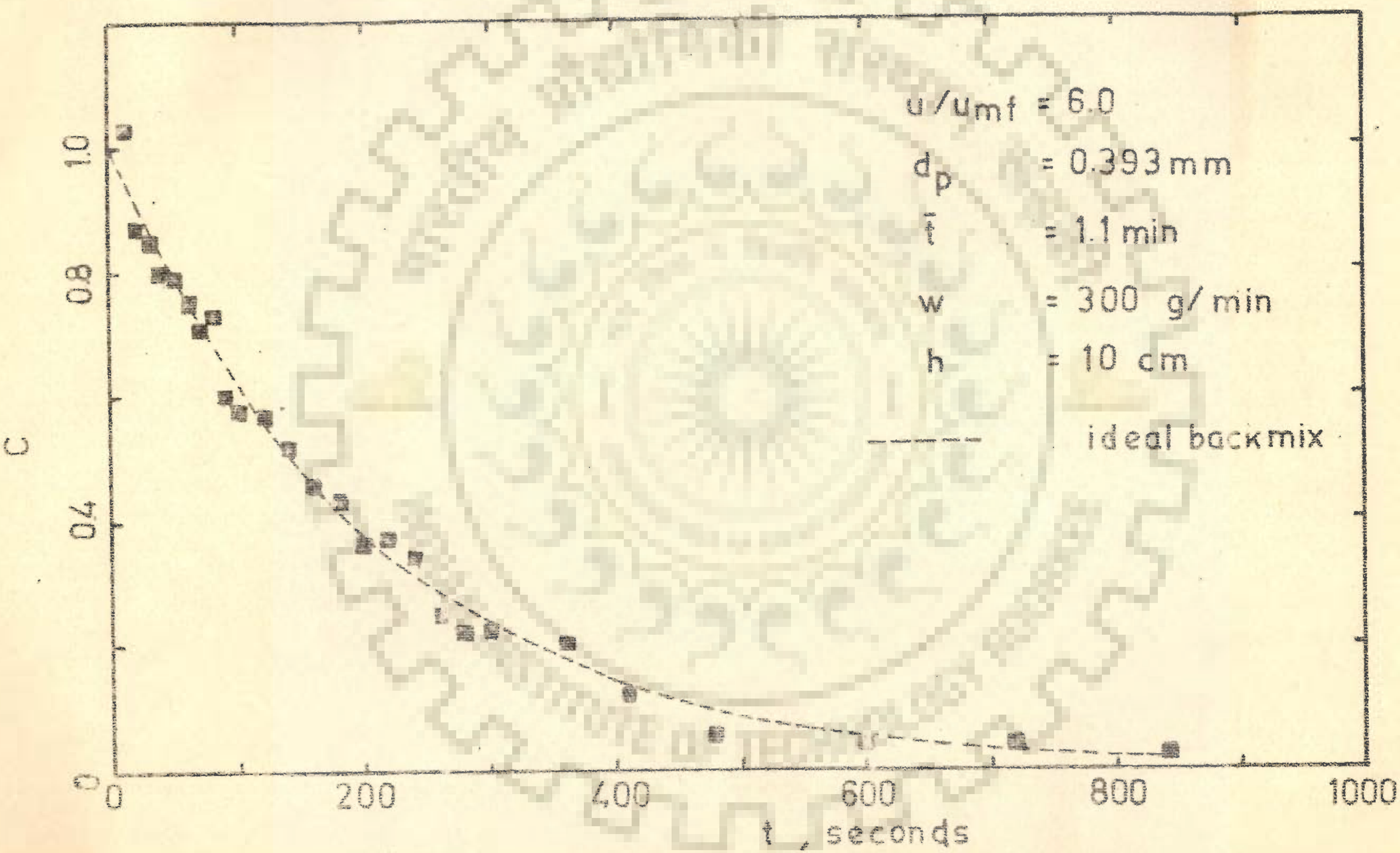


FIG 6.17 STIMULUS RESPONSE IN FLUIDIZED BED , RUN 5

with the curves for ideal backmix flow at same value of mean residence time ($\bar{t}$ = 210 sec.). The points fall very closely on the ideal back mix flow curve. The results of these four runs are also plotted between dimensionless concentration C on log scale and time in seconds on linear scale. For $u = 6.0\ u_{mf}$ all the points fall very closely on a straight line intersecting C axis at unity ($t = 0$) except for the last few points of low concentration. The deviation of these points from the straight line may be attributed to measurement error for discreet tracer particles. For $u = 1.2\ u_{mf}$, the points show a greater scatter and th best straight line connecting these points reach a value of C equal to unity at positive values of t. The dotted lines in Fig. 6.18 represent ideal backmix flow for the corresponding $\bar{t}$. Moo-Young et. al (76) and Cholette and Cloutier(20) and Zaloudik (118) have given non-ideal flow parameters to represent practical situations.

The findings of these workers also confirm that the experimental results presented in Fig. 6.18 indicate close to backmix flow behaviour of fluidized beds even at air velocity as low as $1.2\ u_{mf}$. For a bed height 10 cm, t 70 sec and $u = 1.2\ u_{mf}$, the best straight line indicate slight short circuiting, but for the same conditions of u/u_{mf} and bed height decrease in feed rate to give higher mean residence time, $\bar{t}$ = 600 sec, the behaviour of the bed lcan be represented by slight plug flow in series with complete backmix flow. The response line for bed height of 30 cm was again close to backmix flow with little bit of dead zone associated with it. For velocities greater than $1.5\ u_{mf}$, these deviations

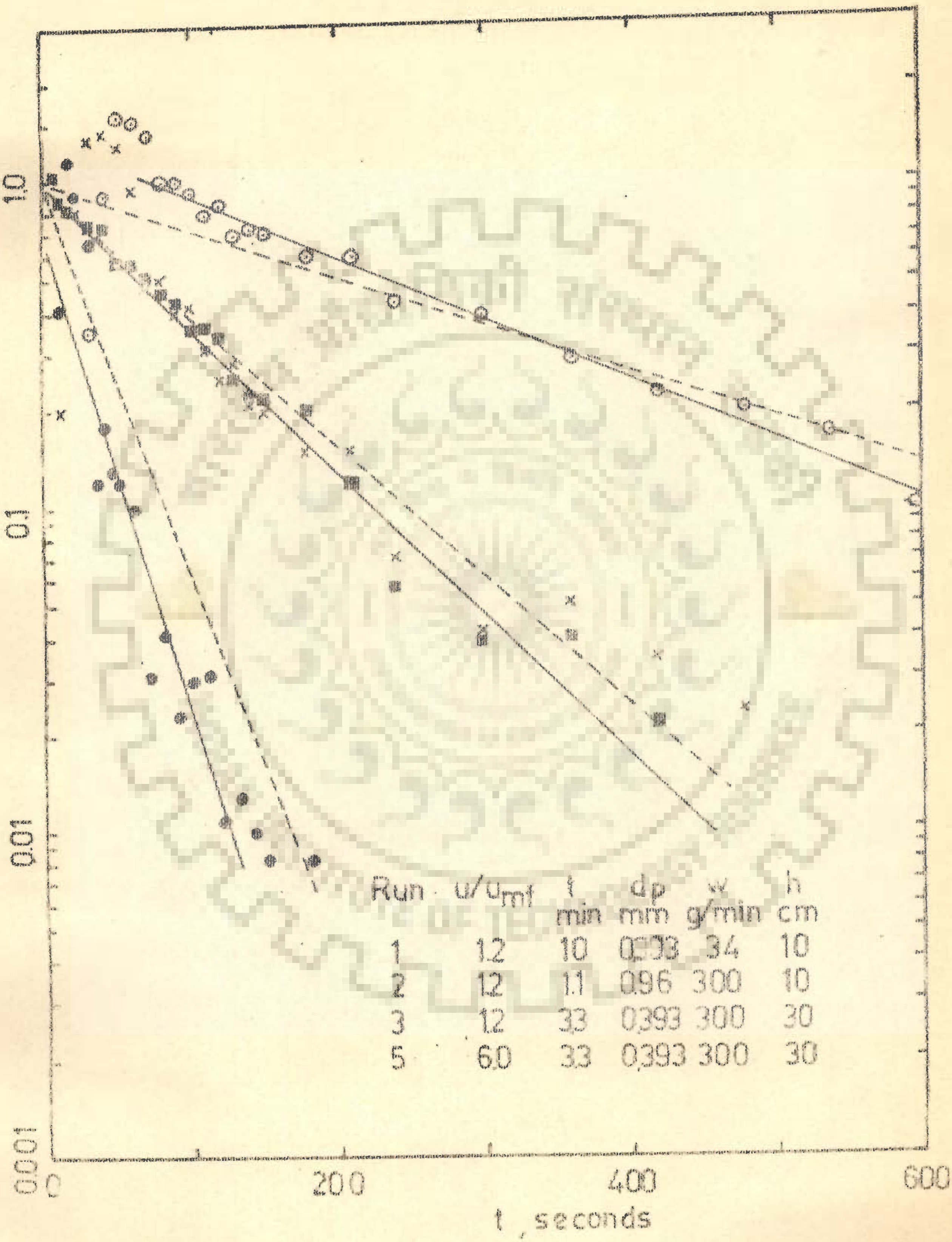


FIG. 6.18 RESPONSE LINES IN FLUIDIZED BEDS

disappear and the bed behaviour tended to approach complete backmixing.

As would be expected, at low air velocity ($u = 1.2\ u_{mf}$) short circuiting is noticed only at high solid flow rates (300 g/min) and low bed heights of 10 cm. Slight plug flow tendency is quite likely at low solid flow rates (34 g/min) corresponding to a mean residence time of 600 seconds and high bed height of 30 cm at low air velocity. Morris et. al (78) have found similar results for fluidizing velocities between $1.3\ u_{mf}$ to $1.5\ u_{mf}$ for similar beds with solid feed from top and solid removal from bottom. They used a bed 9 inches in diameter with six outlets for solids removal. As indicated earlier, the tortuous path for solid removal resulted in significant end effects leading to somewhat misleading interpretations of solid flow behaviour in moving beds by these workers. The tortuous path for solid removal in the case of fluidized bed is likely to cause less error as compared to moving bed because mixing conditions in fluidized beds are close to complete backmixing. The results of Morris and co workers also show that for air velocity between $1.3\ u_{mf}$ to $1.5\ u_{mf}$ the flow of solids is not completely backmixed but for air velocity greater than $1.5\ u_{mf}$ the flow tends to become completely backmixed. These results are in good agreement with the results obtained in the present investigation. Similar conclusions are also drawn by Rowe and Sutherland (85). Yagi and Kunii(112) and Kamiya (48) have also investigated the flow behaviour of solids in fluidized beds and reported that the assumption of backmixing for solid flow in fluidized beds is quite adequate for air velocities higher than $1.5\ u_{mf}$. Flow behaviour of solids at air velocities lower than 1.5 u_{mf} was not investigated by these workers because such operation is not generally used in industry. Sutherland (96) has extended the study of solid axial mixing to tapered beds and annular beds and observed that the effect of tapering is to reduce the vertical mixing rate, specially for very deep beds of dense materials at flow rates little greater than minimum fluidization velocity only. Similar effect is also observed by him for annular beds.

Many investigations have analysed the flow behaviour of solids at low velocities close to u_{mf} values in terms of hold-back and segregation The concept of hold-back and segregation was first introduced by Danckwerts (22) and it has found extensive use for interpreting the flow behaviour of solids in fluidized beds. Tailby and Cocquerel(100) have shown that air velocity, solid feed rate and bed aspect ratio affect the solid flow behaviour in fluidized bed but the effect of these variables even under extreme conditions does not change the value of hold back and solid flow conditions significantly from ideal backmix behaviour. Heertjes et. al (42) have studied the non-ideality of solid flow in rectangular cross-section bed of fluidized solids.

Another way of expressing the non-ideality of solid flow is to use a diffusional mixing model as presented by Danckwerts (22) However, Morris et. al have calculated from their experimental observations that solid mixing in moving and fluidized beds cannot be described adequately by a simple diffusional model. Morris et. al have also concluded that the residence time distribution observed at the outlet for moving and fluidized beds operated close to u_{mf} appears largely due to the variation in solids velocity across the diameter of the column. In the present investigation attempt has not been made to describe the behaviour of a fluidized bed in non-ideal region, that is, u less than 1.5 u_{mf}, mathematically in terms of any of the available models or by proposing a new model because this region is not important industrially. Furthermore, large number of equationsare available in the literature to describe this non-ideality. The behaviour of feeds with mixed particle sizes not investigated earlier, is emphasized in the present work and attempt is made to study the flow behaviour of each size particles for a mixed feed system. Results on mixed feeds in a fluidized bed are discussed in Section 6.2.4.

The important conclusions about the behaviour of single size particles in the fluidized beds are given below:

(1) Solid flow pattern resembles ideal backmix flow for air velocity of greater than 1.5 u_{mf}. For air velocity values 1.2 u_{mf} to 1.5 u_{mf}, the flow behaviour is somewhat non-ideal in nature and exhibits the presence of some short-circuiting, dead zone and plug flow alongwith the major backmix flow.

(2) The results obtained in the present investigation are in general agreement with those of earlier workers on the continuous fluidized beds.

(3) For air flow less than 1.2 u_{mf}, the results show non-reproducibility.

6.2.4 Fluidized Bed with Mixed Feeds

In order to evaluate the flow pattern of particles of different sizes in a fluidized bed of mixed size particles, experiments were planned similar to those of moving beds with mixed feeds. An arbitrary value of air velocity equal to 3 u_{mf} was selected for the initial experiments. The u_{mf} for such feeds was calculated from Fig. 6.4 using average particle diameter corresponding to solid feed composition. The first set of experiments were carried out with a solid feed composition of 1:1 proportion of 0.393 mm and 0.724 mm particles, at a total solid feed rate of 180 g/min and for an average mean residence time of 113 sec. using tracer particles separately of size 0.393 and 0.724 mm and a mixture of these sizes in 1:1 ratio. The results are given in Table D-2, Appendix D and are shown in Fig. 6-19. In these and subsequent experiments with mixed feeds, it was interesting to observe that the initial samples withdrawn from the bed were found to contain more of 0.393 mm size particles than 0.724 mm particles and only after about 4 to 5 minutes, the exit stream was found to contain the particles of two sizes in 1:1 ratio. The pulse tracer was injected in the fluidized only after the steady state of the main solid feed was attained. In Fig. 6-19 the dimensionless concentration C is plotted as a function of time for these runs. The curves obtained for each tracer type resembled ideal backmix flow curves but the area under each curve is different.

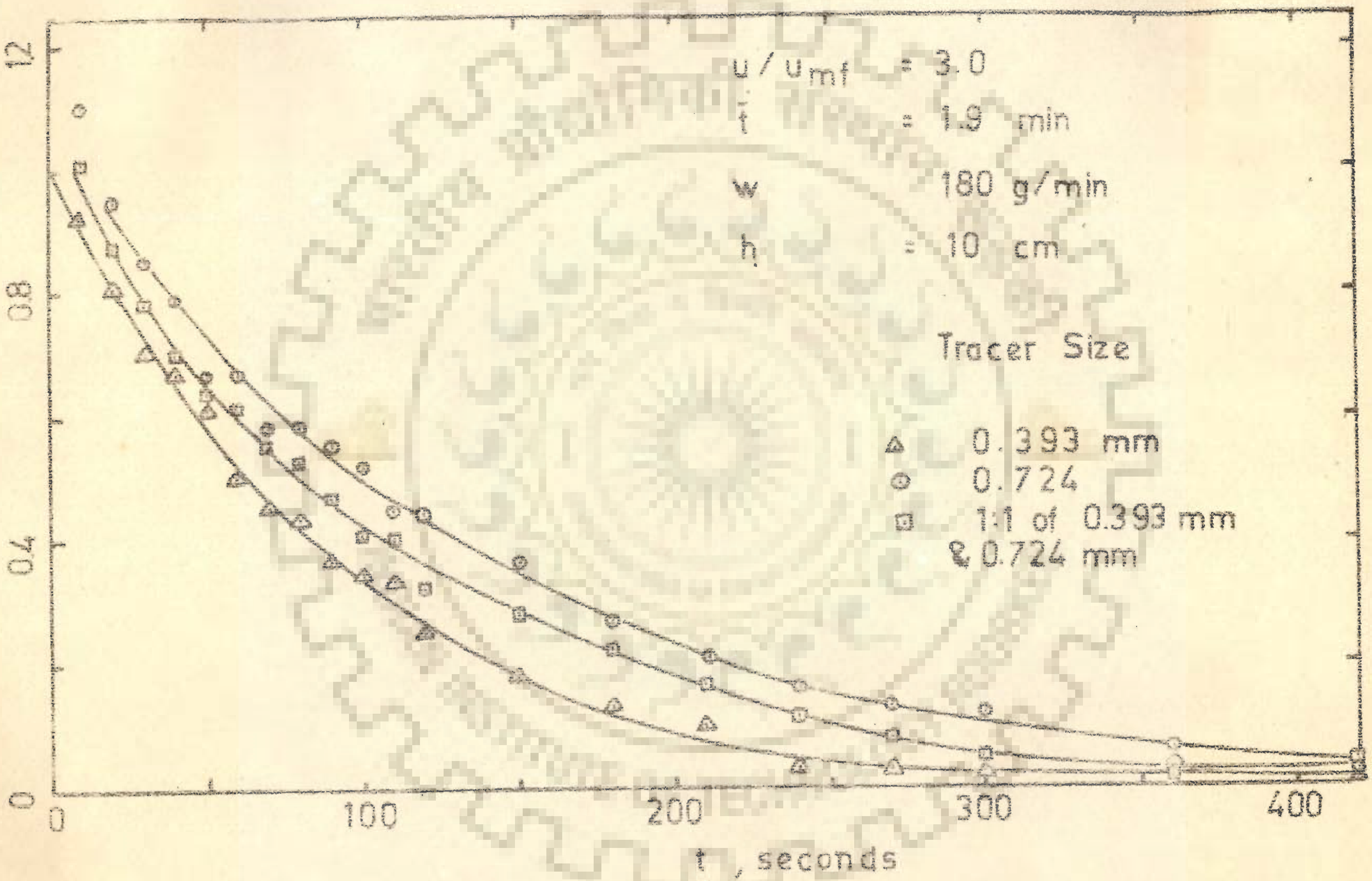


FIG. 6.19 RESPONSE CURVE IN FLUIDIZED BEDS WITH MIXED FEEDS
(Feed size combination 1:1 of 0.393 mm & 0.724 mm)

This suggests that the mean residence time of particles of different sizes in the bed is not identicle. These points are plotted on a semi-log graph paper in Fig. 6. 20 and three straight lines, one for each tracer pulse, with different slopes are obtained. Differences in the slopes indicate that the two particle sizes have different mean residence times in the bed. It is also concluded from the straight lines in Fig. 6. 20 that both these particle sizes were individually backmixed in the bed and yet they are found to have different mean residence times in the bed. The response curve for mixed size tracer is also ideally backmixed. From the slope of the three lines in Fig. 6. 20, the mean residence times of 0.393 and 0.724 mm particles are determined as 87.5 and 138 seconds respectively. This suggests that the hold up of 0.724 mm particles in the fluidized bed is considerably more than the hold up of 0.393 size particles. This result is in conformity with the observation that initially more of small size particles are found in the solids leaving the bed before the attainment of the steady state. In the initial samples the initial accumulation of large size particles in the bed before the steady state of solid flow is achieved results in the larger mean residence time for the larger particle sizes.

At the time of writing this thesis, a short paper by Chechetkin et. al (19) presents the results of a similar study on mixed feed of two particle sizes. The details of the experimental set-up are not given but it appears that they have used an overflow type arrangement for the removal of solids from the fluidized bed.

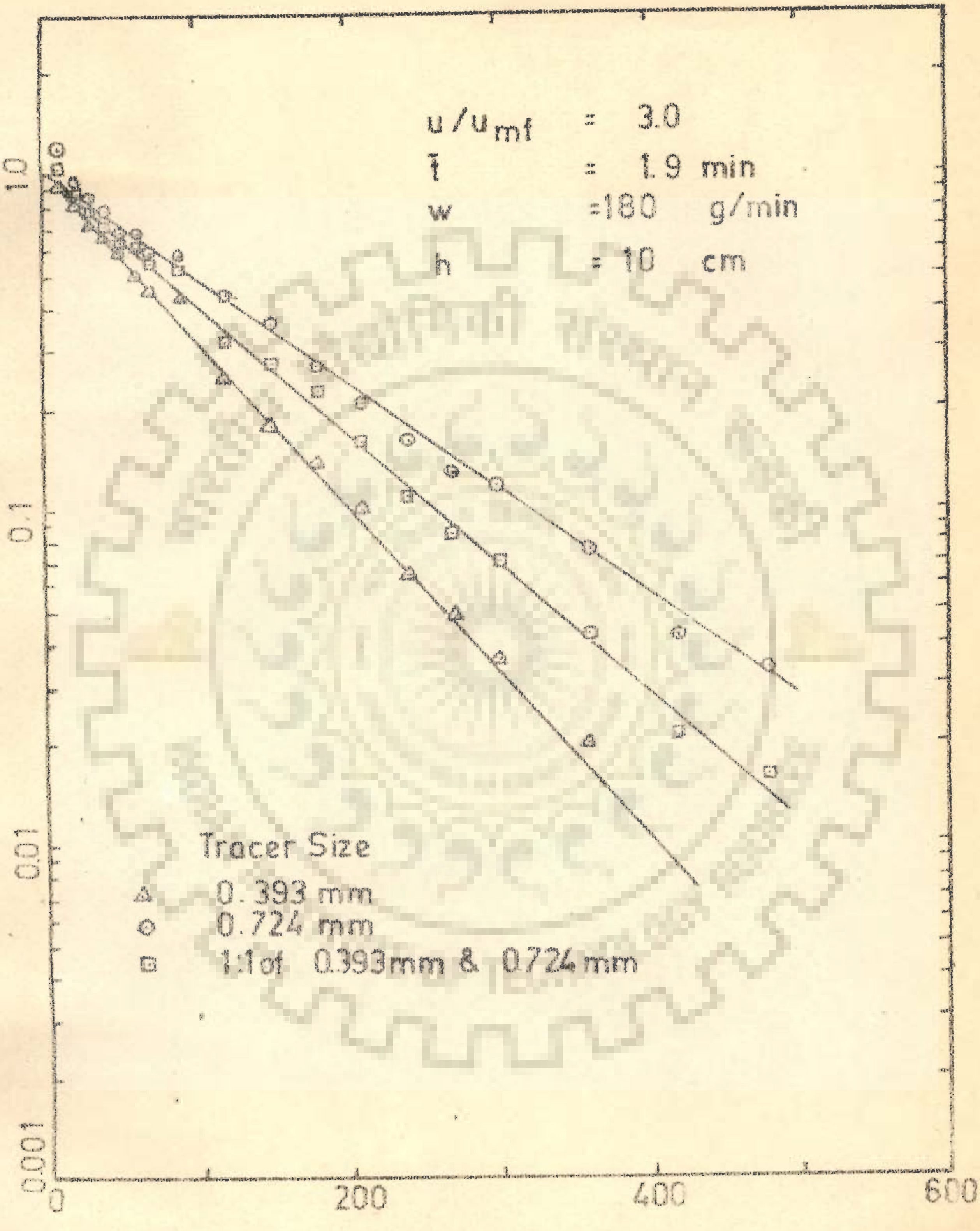


FIG. 6.20 RESPONSE LINES IN FLUIDIZED BEDS WITH MIXED FEEDS
(Feed combination 1:1 of 0.393 mm & 0.724 mm)

These workers investigated the behaviour of polydisperse size quartz sand of size range 0.10 to 1.0 mm, fluidized in a flat model of a flow apparatus of 25 x 250 x 600 mm size. Narrow fraction of 0.1 - 0.2 and 0.63 - 1.00 mm of marble, whose density was close to that of the quartz sand, was used as the pulse tracer. The results obtained by them also show that the mean residence time of larger particles is more than that of smaller particles. The results of Chechelkin and co-workers are in agreement with the present investigation.

This difference in the mean residence times of different size particles in the fluidized beds is very useful because for solid-fluid reactions in a fluidized bed, application of Shrinking Core Model (56) indicates that the time needed for complete reaction for smaller particles is less than the time required for larger size particles. Often the fluidized beds are operated by adjusting the operation to give a desired value for the average mean residence time. In such an operation, the smaller particles tend to become overburnt because of longer residence time in the bed than actually required. Further, the larger particles are only partly converted because residence time for these particles is not sufficient for the completion of reaction. This results in a decrease in overall conversion and a poor product quality. A careful design of fluidized bed reactor so as to give longer residence time for particles of larger sizes, shorter residence time for smaller particles, as observed in the present work, will give higher overall yields and sometimes better product quality.

The experiments of stimulus-response in fluidized bed were also conducted for mixed feeds of compositions 1:1 ratio of 0.724 and 0.96 mm and 1:1:1 proportion of 0.393, 0.724 and 0.96 mm particles. The results of these runs are given in tables D-3 and D-4, Appendix D, and plotted in Figs. 6.21 and 6.22. In both these figures the effect observed earlier in Fig. 6.19 is present again. From the slopes of the lines in Fig. 6.21, the calculated values of the mean residence times are 80 sec. for 0.724 mm particles, 145 sec for 0.96 mm particles and 113 sec. for the mixed tracer particles. The mean residence time for the mixed feed calculated from bed hold up and solid feed rate also comes to 113 sec. Similarly, the mean residence times of 0.393, 0.724 and 0.96 mm size particles is calculated from Figure 6.22 and the values are 62, 92 and 152 seconds respectively.

To verify the above results, some preliminary runs were taken to qualitatively determine the bed composition at steady state. The air flow to the fluidized bed at steady state was suddenly cut-off along with the solid feed and discharge. The bed material was removed and sieve analyzed. For the conditions of Fig. 6.19, the bed was found to contain 132 grams of 0.393 mm particles and 208 grams of 0.724 mm particles. If it is assumed that the particles of both these sizes were individually backmixed in the bed, then the mean residence time for the solid feed rates of 90 g/min of each size particles shall correspond to 88 and 138 sec. for sizes 0.393 and 0.724 mm respectively. This compares very well with the values obtained for

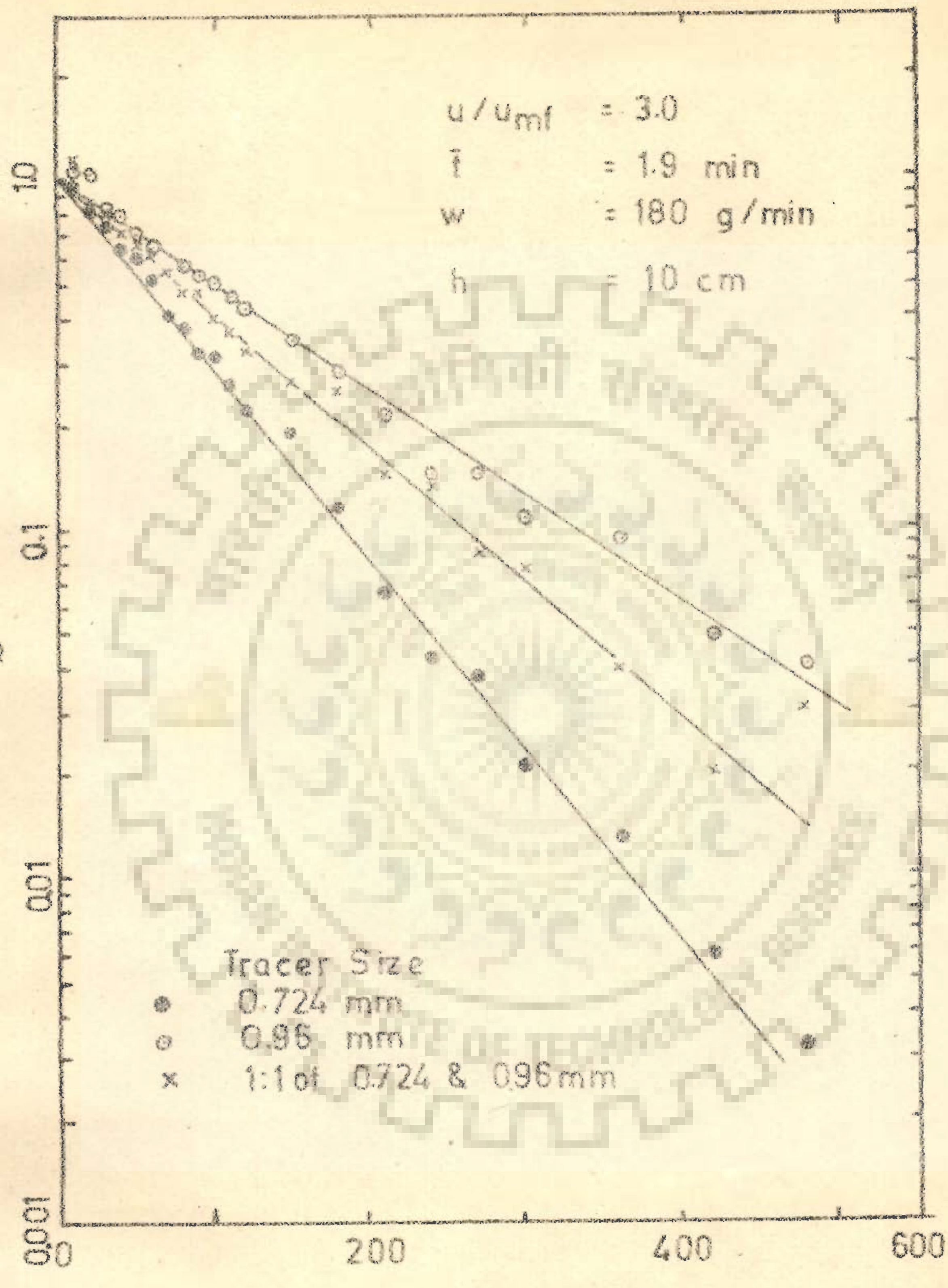


FIG. 6.21 RESPONSE LINES IN FLUIDIZED BEDS WITH MIXED FEEDS
(Particle size combination 1:1 of 0.724 mm & 0.96 mm

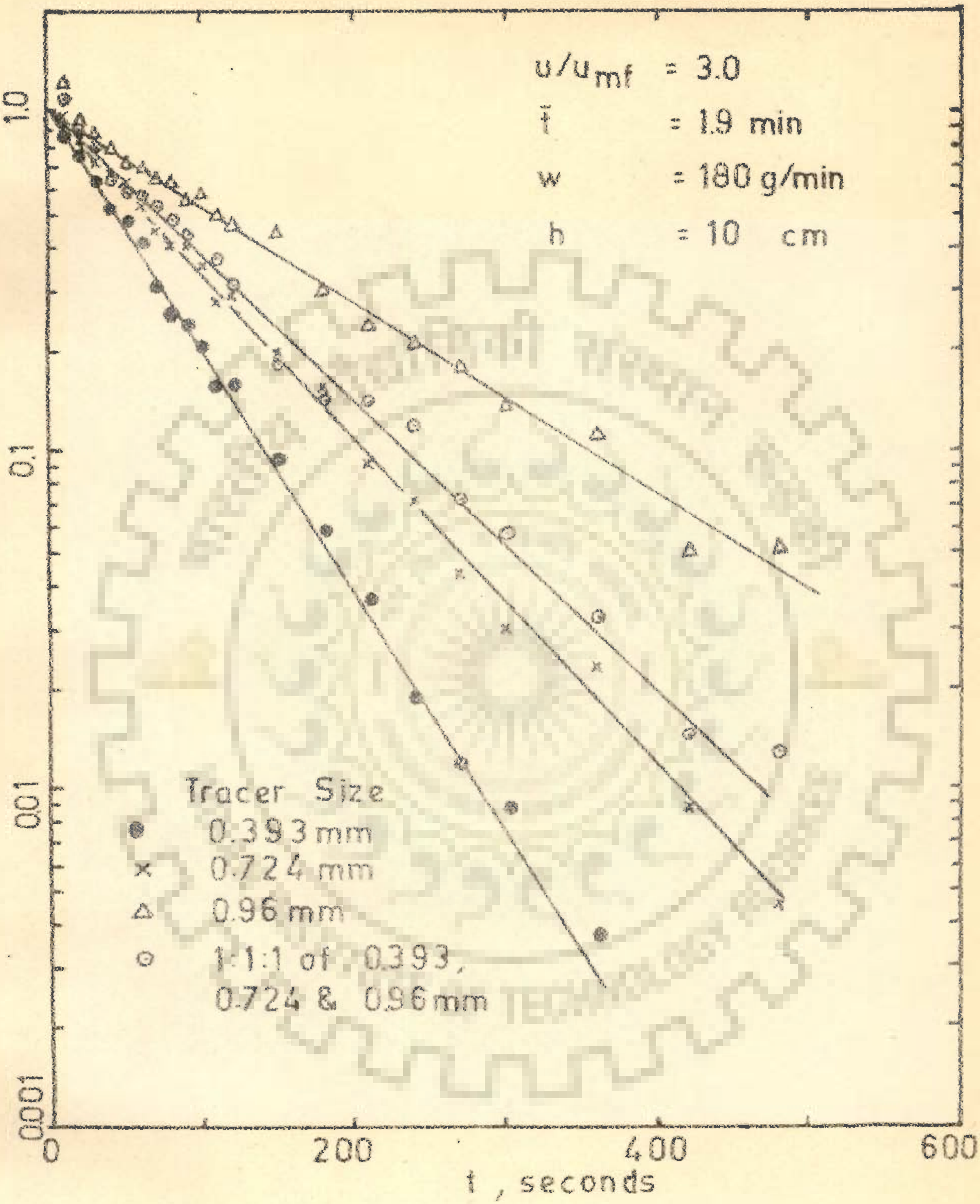


FIG. 6.22 RESPONSE LINES IN FLUIDIZED BEDS WITH MIXED FEEDS
(Feed size combination 1:1:1 of 0.393, 0.724 & 0.96 mm)

mean residence times 87.5 sec. and 138 sec. by stimulus response techniques. Similarly, the mean residence time of 113 sec. obtained through stimulus-response experiments for mixed tracers is exactly the same as obtained by dividing the total hold-up by the solid feed rate. Similar experiments for analysing bed hold-ups for larger size combinations and mixed feeds of all the three sizes also gave results identical to those obtained by stimulus-response technique. The important conclusions about the behaviour of the mixed size feeds in the fluidized beds are given below:

(i) In a fluidized bed of mixed sizes with continuous flow of solids, particles of each size experience a completely backmix flow.

(ii) The mean residence time of each size particle is different in the fluidized bed.

(iii) The mean residence time of larger size particles is found to be more than the mean residence time of small size particles.

(iv) The bed hold-up for particles of different sizes in a fluidized bed at steady state to predict the mean residence time for each size particle is as accurate as stimulus-response technique.

Since the particles are individually backmixed, the mean residence time could be directly calculated from the bed composition at steady state rather than resorting to a time consuming stimulus-response - technique Further investigations to study the flow behaviour of solids in a fluidized bed were carried out by measuring steady state bed compositions directly.

6.3 STUDIES ON HOLD UP RATIOS IN FLUIDIZED BEDS

Once it is established that different size particles have different mean residence times in the bed, the ratio of the mean residence times are defined as hold up ratio

$$H(2,1) = \bar{t}_2 / \bar{t}_1 \qquad \text{(eq. 6.1)}$$

$$H(3,2) = \bar{t}_3 / \bar{t}_2 \qquad \text{(eq. 6.2)}$$

where suffix 1, 2 or 3 denote particles of small, middle and large sizes respectively. The hold up ratio H(2,1) provides a measure of the extent of variation in mean residence time of particles of different sizes.

For each particle size combination obtained from particles of three available sizes, four variables were studied. These variables are air flow rate G, bed hold up W, solid feed rate w and solid feed composition. After fixing the values of bed hold up, solid feed rate and solid feed composition the air flow rate was varied in the range of about 1.5 G_{mf} to 6 G_{mf} for each run.

6.3.1 Hold up Ratios for Size Combinations of 0.393 mm and 0.724 mm Particles

The values of hold up ratio H(2,1) given in Tables E-1 to E-4, Appendix E are plotted in Fig. 6.23 as a function of air flow rate G for different values of solid hold up in the bed. From the figure it is clear that the value of H(2,1) first increases with increase in G for low values of G, reaches a maximum and a further increase in the value of air flow rate, G, results in a decrease in the value

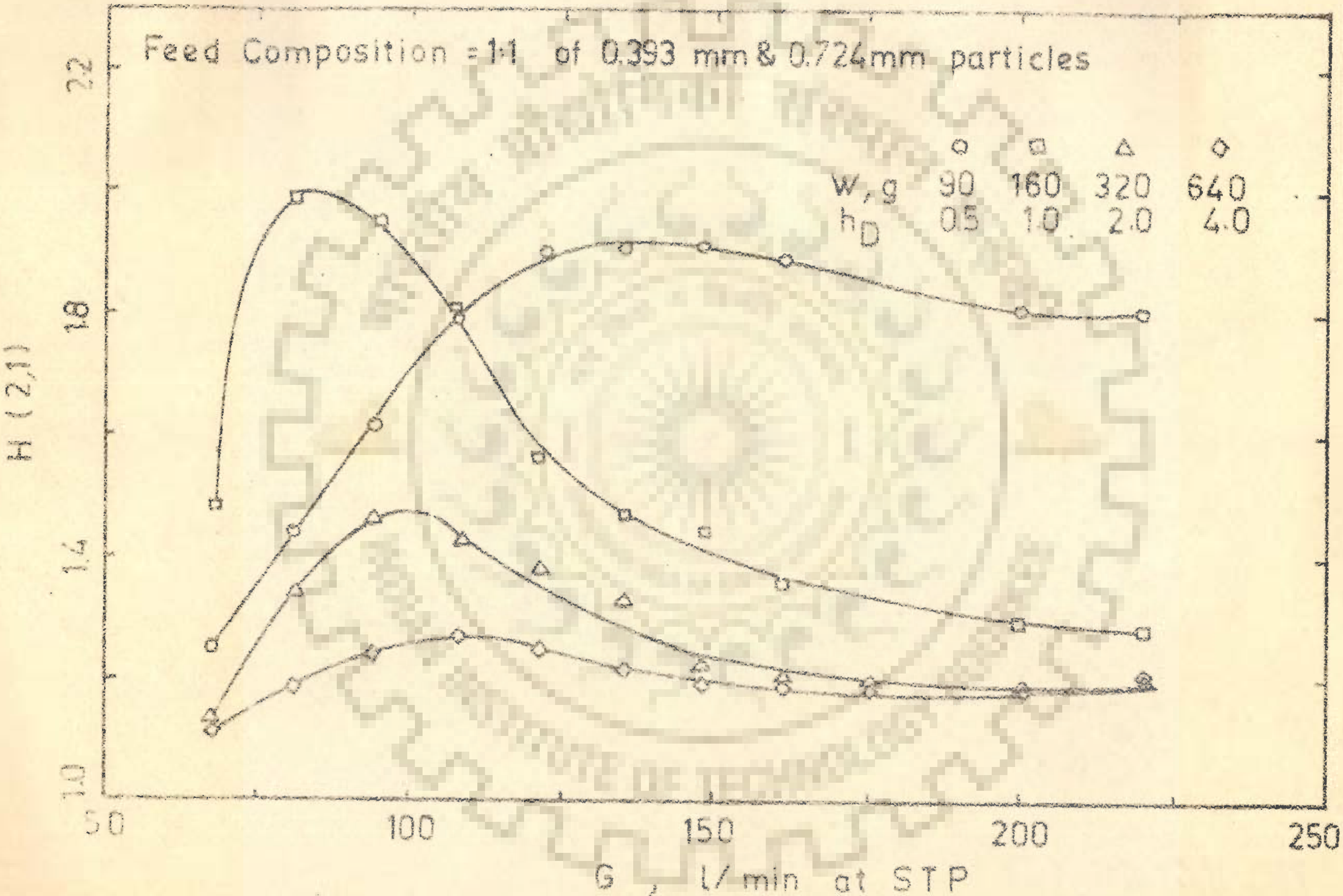


FIG. 6.23 EFFECT OF AIR FLOW RATE G AND BED HOLD UP ON HOLD UP RATIO, H (2,1)

of H(2, 1). Chechetkin et. al (19) have also reported a decrease in the value of H(2, 1) with increase in G for high gas flow rates but they did not report the initial increase of H(2, 1) with G because the three air flow rates used by them were very high as compared to G_{mf}. The bed hold up has a significant effect on the shape of the curves. For a bed hold up corresponding to a static bed height to diameter ratio h_D of unity (90 g solid hold up in the bed), H(2, 1) increases sharply to about 2.1 as the value of G increases to 150 l/min and then it decreases slowly reaching a value of 2.0 at 220 l/min. For h_D of 2, H (2, 1) increases rapidly to a value of 2.2 at 80 l/min and then it decreases sharply to 1.5 at 220 l/min. The nature of curves for h_D greater than 2 is similar to that at h_D value of 2 but the maximum value of H(2, 1) tends to decrease and this maximum is achieved at higher air flow rates with increasing solid hold up in the bed.

The value of hold up ratio H(2, 1) for different solid feed rates and three bed hold ups given in Tables E-5 to E-13, Appendix E are shown in Fig. 6.24 only for a bed hold up corresponding to h_D value of unity for different solid feed rates. The nature of curves remains the same, that is, H(2, 1) first increases, reaches a maximum and then decreases with increasing air flow rates. However, the maximum value of H(2, 1) tends to decrease and this maximum is achieved at lower air flow rates with increasing solid feed rate. Further the decreasing portions of the curves for different solid feed rates are approximately

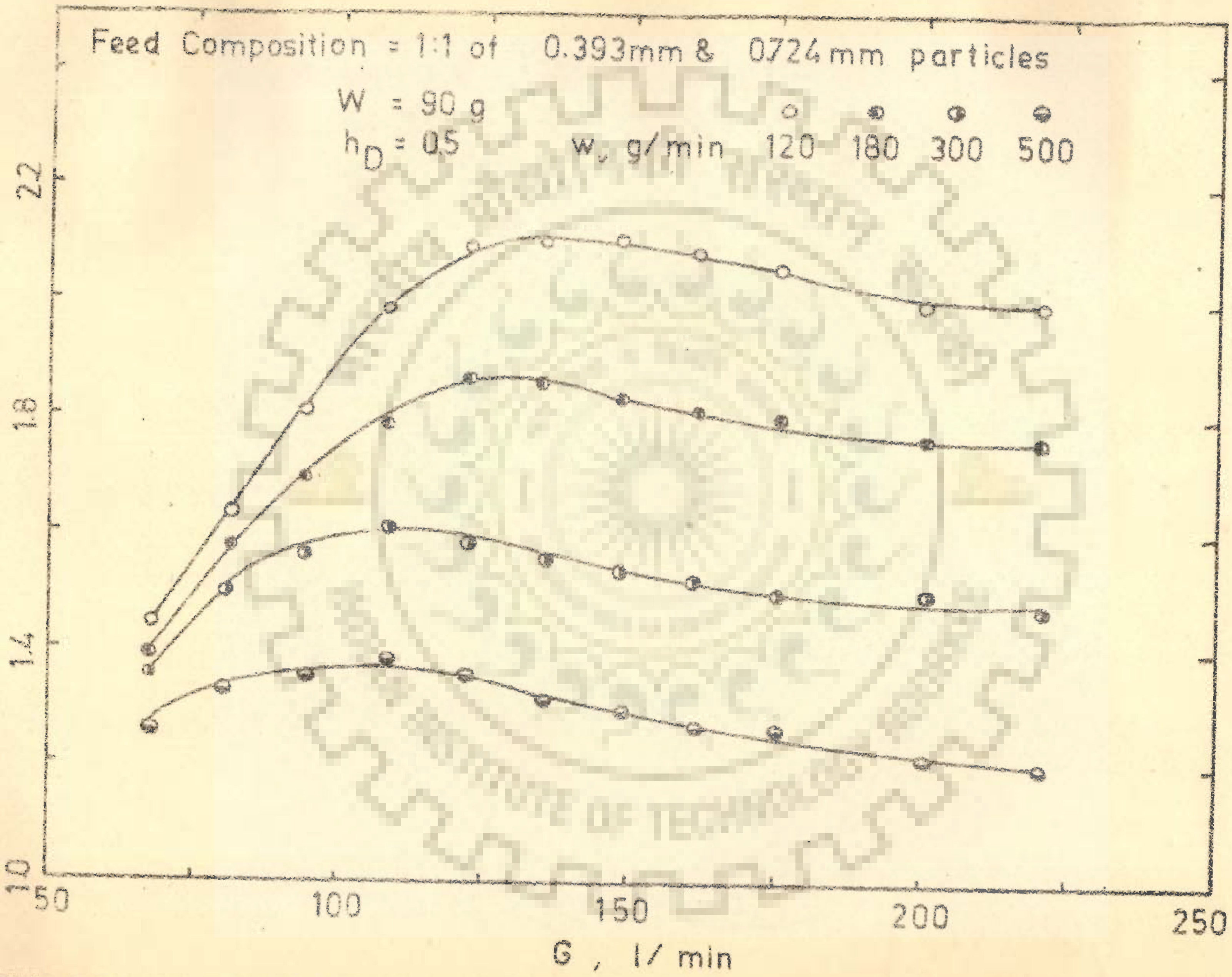


FIG. 6.24 EFFECT OF SOLID FEED RATE w ON HOLD UP RATIO H(2,1)

parallel. The values of hold up ratio H(2,1) for different feed compositions and two bed hold ups given in Tables E-14 to E-17, Appendix E are plotted in Fig. 6.25 only for a bed hold up corresponding to h_D value of unity for different feed compositions. Interestingly there appears to be a lateral shift of the two curves for 1:4 and 4:1 compositions as compared to the curves for 1:1 composition. This observation suggested that the air flow could be represented more appropriately in dimensionless form as u/u_{mf} but for such runs two values of u_{mf} are possible, one corresponding to the feed and the other corresponding to the bed composition. Theoretically, it is the bed composition at steady state which shall determine u_{mf}, and, therefore, dimensionless air flow rate was based on the bed composition at the steady state. For each run of Tables E-1 to E-17, u/u_{mf} was calculated by first calculating average dp for the bed composition and then using this average diameter in Leva's equation to calculate the u_{mf} as discussed in Section 6.1. When the data of Fig. 6.25 is replotted as a function of u/u_{mf}, Fig. 6.26, it is observed that the lateral shift of the curves for different feed composition is eliminated and almost all the points fall on the same curve. Therefore, the use of u/u_{mf} eliminates the effect of changes in the feed composition.

When the data of Fig. 6.24 is replotted as a function of u/u_{mf}, Fig. 6.27, the parallel nature of the decreasing portions of the curves is still maintained. This suggests that the effect of solid feed rate w can be correlated by using a relationship

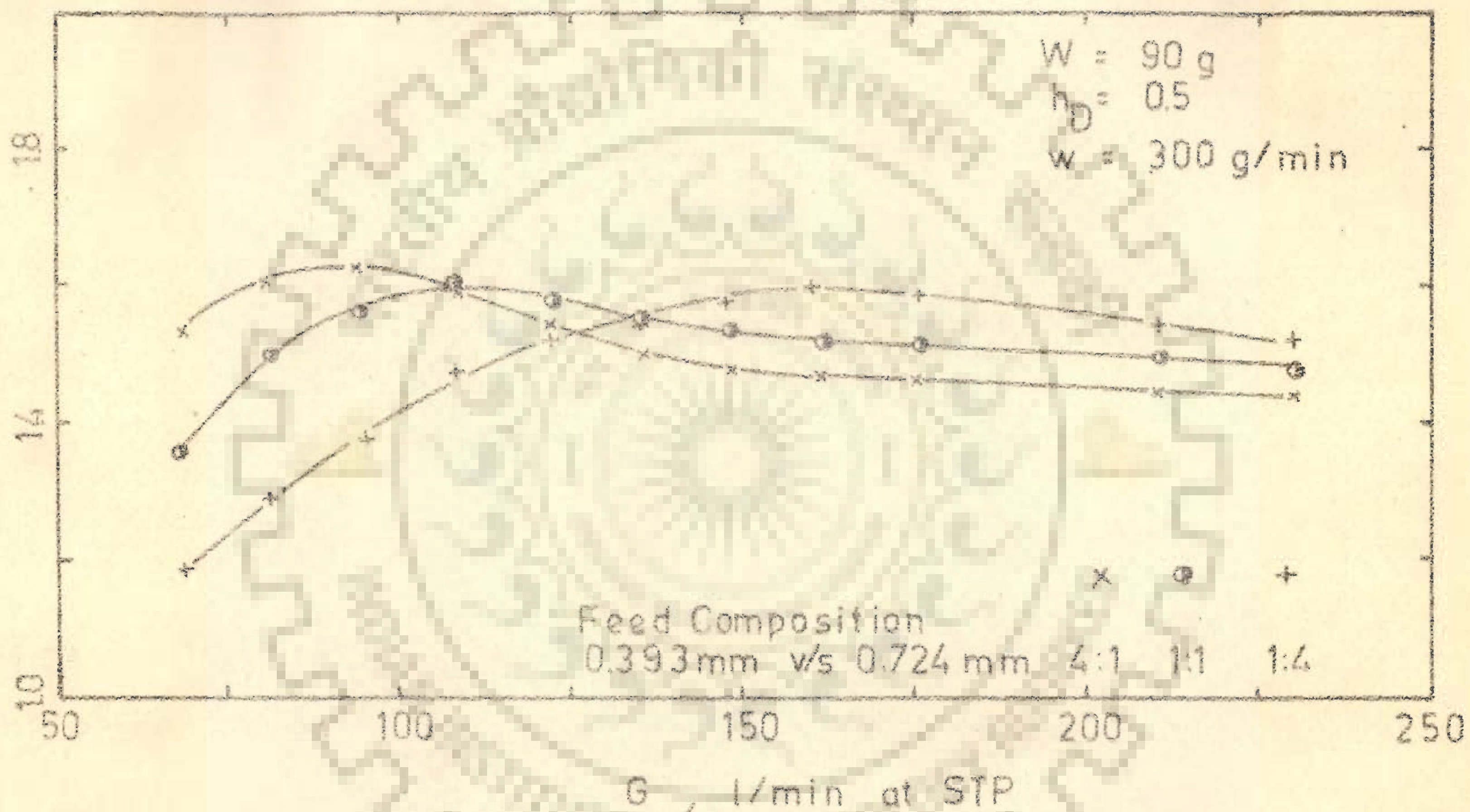


FIG. 6.25 EFFECT OF SOLID FEED COMPOSITION ON HOLD UP RATIO, H(2,1)

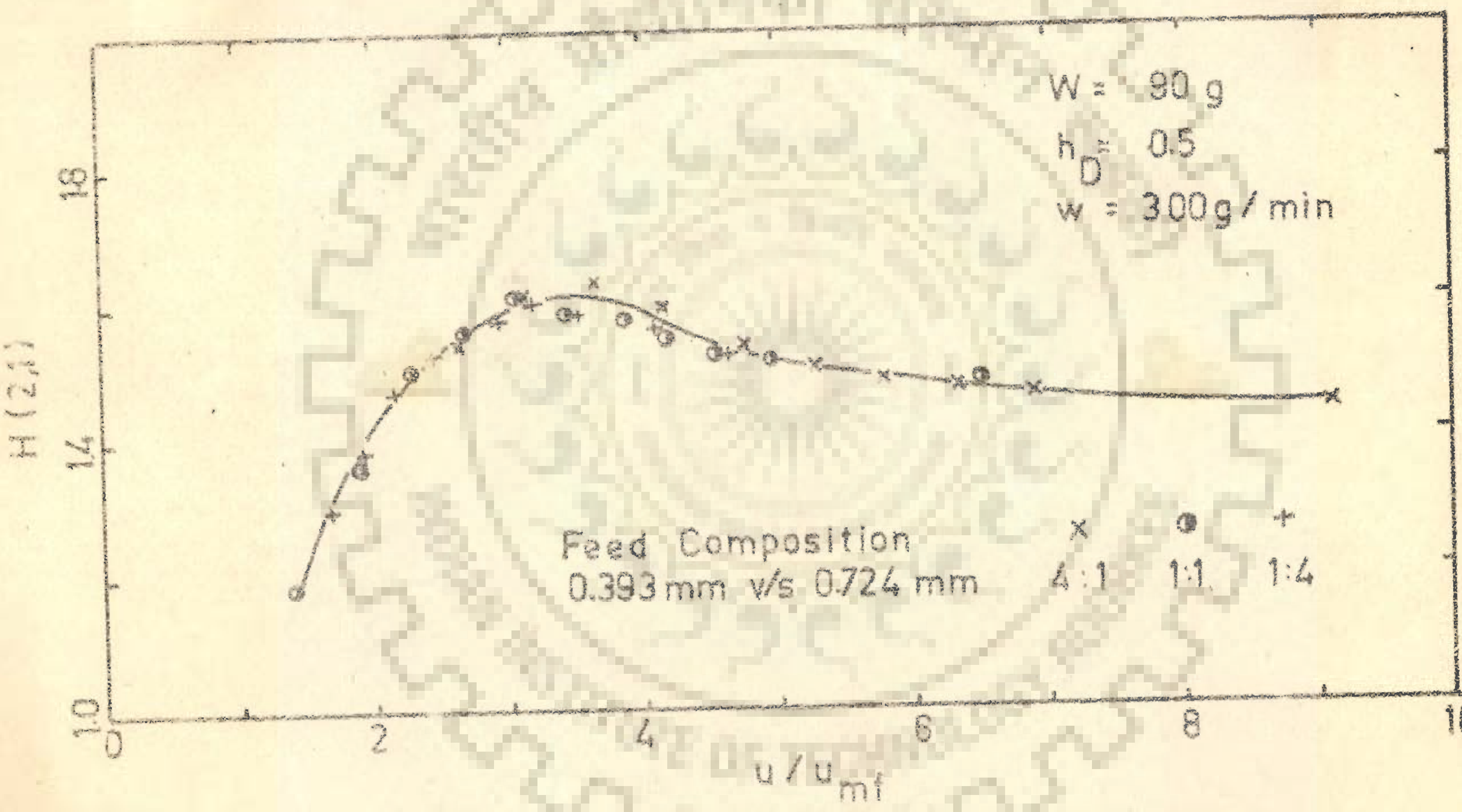


FIG. 6.26 EFFECT OF REPLACING AIR FLOW RATE, G BY u/u_{mf}

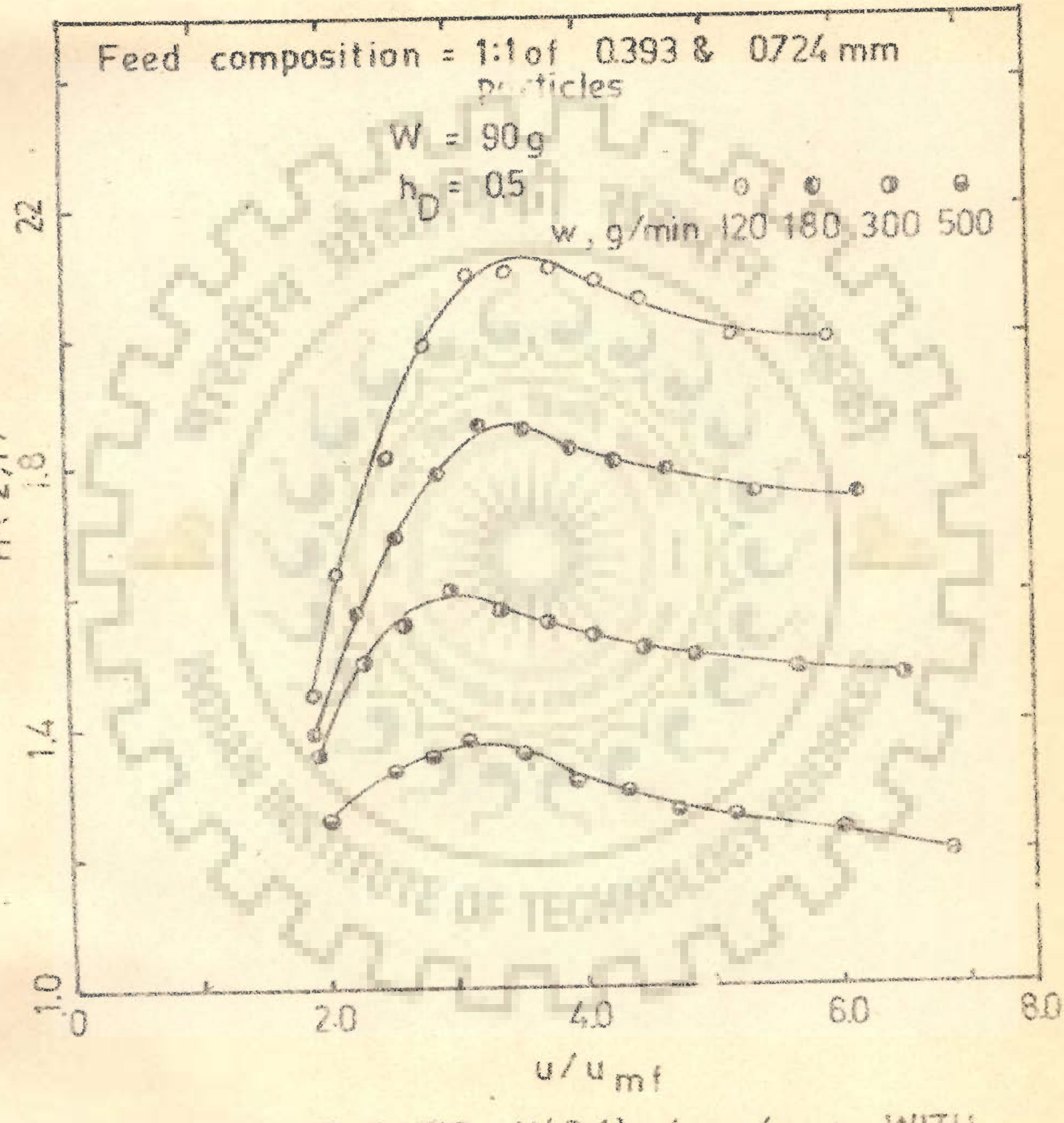


FIG.6.27 HOLD UP RATIO H(2,1) v/s u/u_{mf} WITH w VARYING

$$H(2,1) = f(u/u_{mf})\,(w)^{n} \quad \ldots \text{(eq. 6.3)}$$

where n is a constant. However, the use of w as such in the correlation is not very convenient and its replacement by a suitable dimensionless parameter is desirable.

The gas contact time in fluidized bed can be defined in one of the following ways:

$$T_{GA} = V_f / G \quad \ldots \text{(eq. 6.4)}$$

$$T_{GB} = V_s / G \quad \ldots \text{(eq. 6.5)}$$

Where V_f is the fluidized bed volume and V_s is the volume of solids in the bed. T_{GA} is less precise as compared to T_{GB} because the bed heights for fluidized beds at high u/u_{mf} is ill defined and correct estimation of V_f is not possible. T_{GB} is actually the space time and finds frequent use in solid catalyzed gas reactions. Levenspiel (63) has discussed the use of space time T_{GB} and also of the weight time defined as

$$T_w = W/G \quad \ldots \text{(eq. 6.6)}$$

in the design of fluidized bed reactors. The solid feed rate can be defined better in terms of the average time spent by the solids in the bed. Three ways of defining this time are indicated below:

(i) $$\tau_s = W/w \quad \ldots \text{(eq. 6.7)}$$

(ii) $\tau_A = \tau_s / T_{GA}$ ....(eq. 6.8)

(iii) $\tau_B = \tau_s / T_{GB}$ ......(eq. 6.9)

The first definition, eq. 6.7, gives the average time for which the solids remain in the bed, τ_s. Since τ_s has a dimension of time, it is better to use the dimensionless parameter τ_A or τ_B as defined by eqs. 6.8 and 6.9. As pointed out earlier that the correct estimate of T_{GA}, specially at high air velocities, is not possible, hence τ_A has some measure of uncertainty. Therefore, τ_B was chosen as the most appropriate parameter to account for the effect of solid feed rate and eq. 6.3 can be written as

$$\frac{H(2,1)}{(\tau_B)^n} = f\,(u/u_{mf}) \qquad \text{....(eq. 6.10)}$$

In Fig. 6.27 the data points shown for different solid feed rates are not for similar values of u/u_{mf}, even though the values of gas flow rates were similar but the values of u_{mf} changed because of changes in the bed composition. It was, therefore, not possible to use regression analysis without graphical interpolation to estimate the value of constant n in eq. 6.10 and only graphical procedure was followed. H(2,1) and τ_B were obtained from curves shown in Fig. 6.27 for different values of u/u_{mf}. Values of H(2,1) are plotted as a function of τ_B in log-log graph paper, Fig. 6.28, nearly parallel lines with average slope of 0.3 are obtained. Using 0.3 as the value for constant n, Fig. 6.29 is a

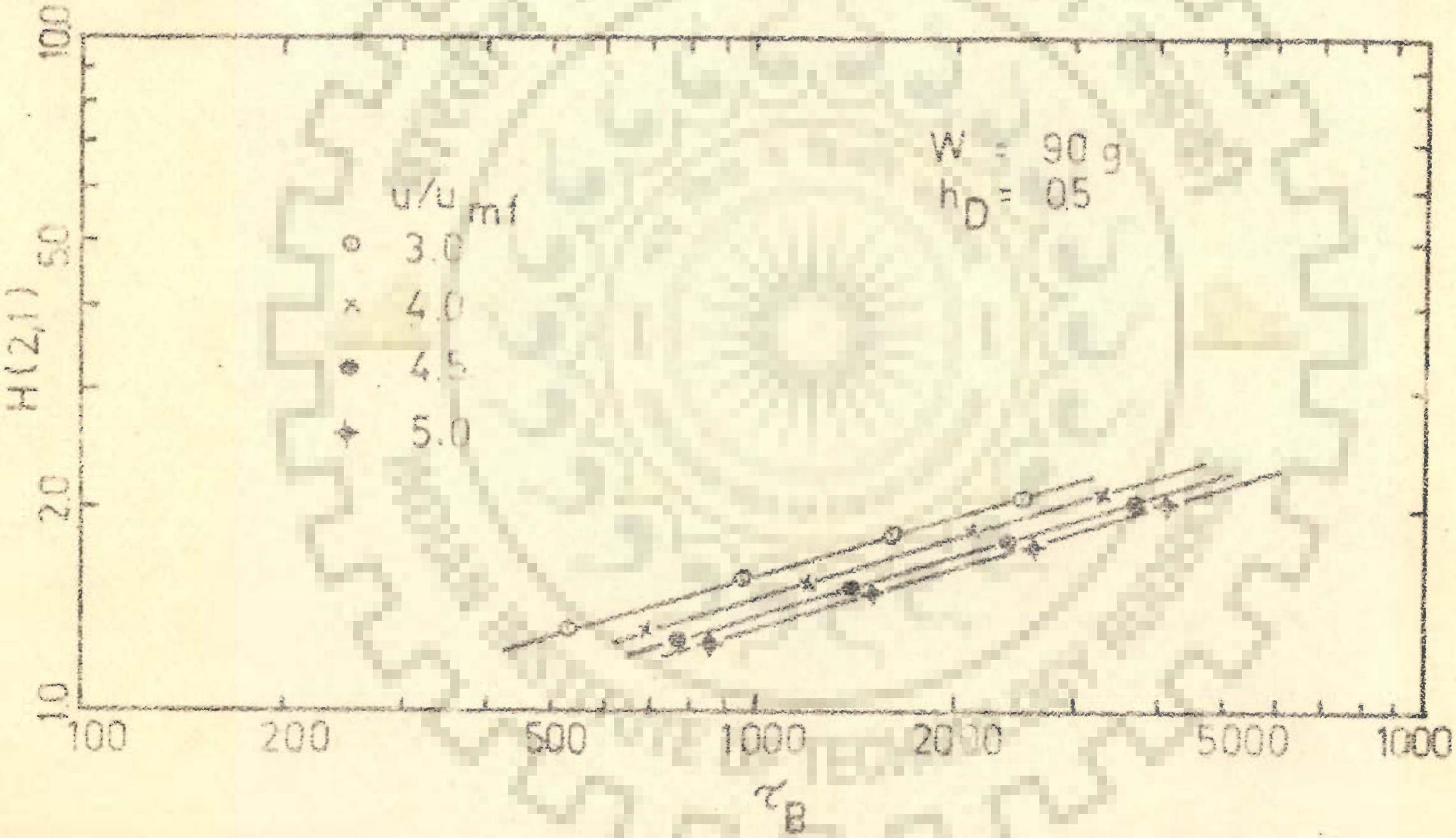


FIG. 6.28 CORRELATING H(2,1) AND τ_B

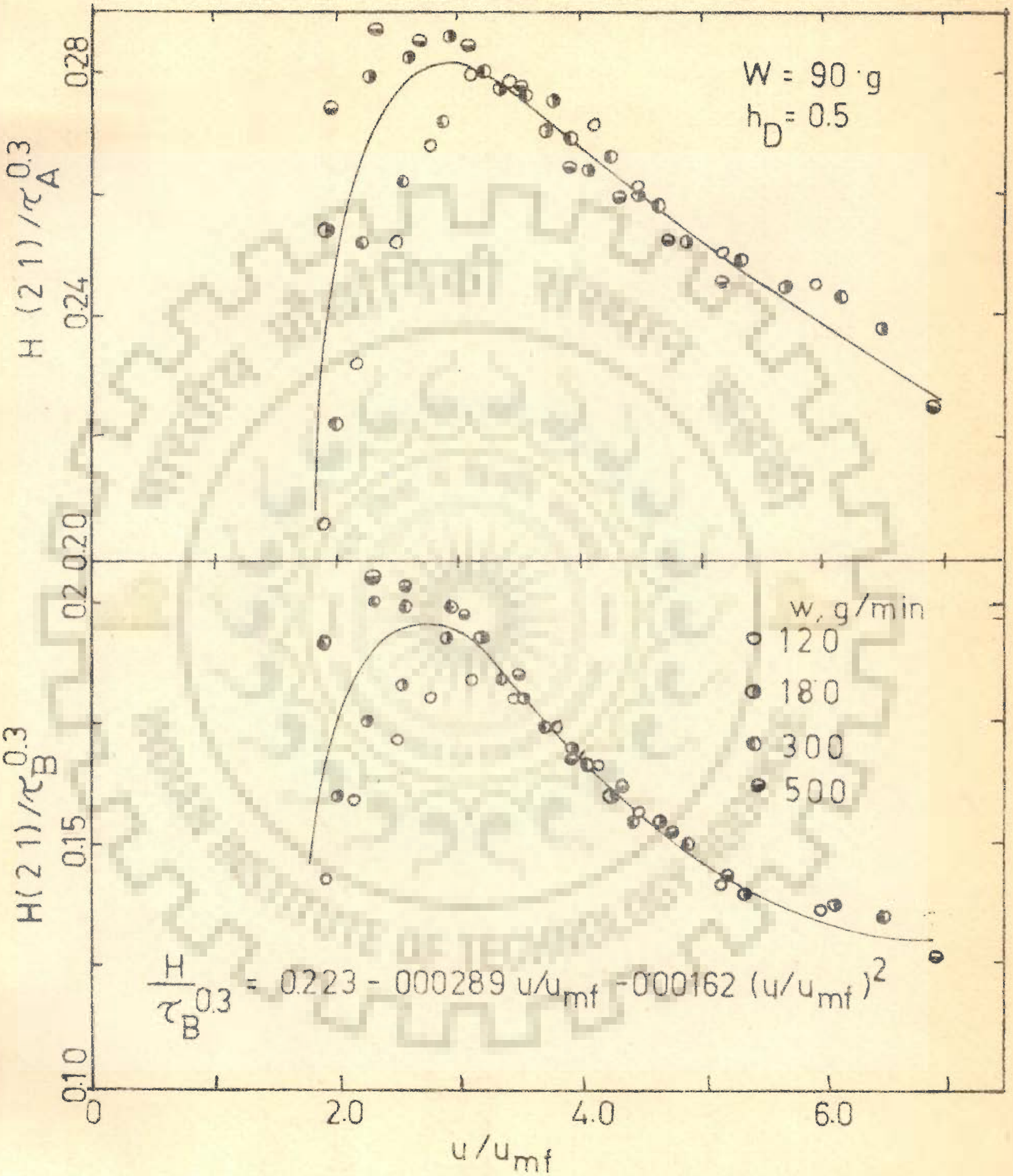


FIG. 6.29 CORRELATION OF (H 2,1) AND τ_B OR τ_A FOR h_D = 0.5

plot of $H(2,1)/(\tau_B)^{0.3}$ as a function of u/u_{mf} for the data points shown in Fig. 6.27. Even though τ_A has some measure of uncertainty, for comparison sake, $H(2,1)/(\tau_A)^{0.3}$ is also plotted in Fig. 6.29. It is clear from Fig. 6.29 that the four curves of Fig. 6.27 get superimposed into one curve with little scatter especially for u/u_{mf} values larger than 2.5. The curve has two distinct regions, one for u/u_{mf} less than 2.5 where $H/(\tau_B)^{0.3}$ increases sharply with u/u_{mf} and the other for u/u_{mf} greater than 2.5 where $H(2,1)/(\tau_B)^{0.3}$ falls slowly with further increase in u/u_{mf}. The region u/u_{mf} less than 2.5, where the scatter is more is generally not important because most industrial fluidized beds operate only at the fluidizing velocities greater than 2.5 u_{mf}.

For fluidizing velocities greater than 2.5 u_{mf}, a second order polynomial of the following form was tried to correlate the values of $H(2,1)/(\tau_B)^{0.3}$ as a function of u/u_{mf} using regression.

$$\frac{H(2.1)}{(\tau_B)^{0.3}} = b + c\,(u/u_{mf}) + d\,(u/u_{mf})^2 \qquad \ldots \text{(eq. 6.11)}$$

The values of constants b, c and d were found by regression analysis using IBM 1620 computer and are given below:

b	=	0.223
c	=	-0.00289
d	=	-0.00162

The error in the calculated values of $H(2,1)/(\tau_B)^{0.3}$ is found to be within $\pm$ 10 per cent of the experimental values.

Similar analysis was carried out for higher solid hold up in the bed. The data given in Tables E-2, E-8 to E-10, Appendix E, for solid hold up in the bed corresponding to $h_D = 1.0$ is plotted in Fig. 6.30(a) and that given in Tables 2 E.3, E.11 to E.13 for $h_D = 2.0$ is plotted in Fig. 6.30 (b). The parallel nature of curves for the decreasing portion as observed in Fig. 6.27 is again visible in Fig. 6.30 and the values $H(2.1)/(\tau_B)^{0.3}$ as a function of u/u_{mf} are shown in Fig. 6.31 (a) and Fig. 6.31(b) for two bed hold-ups. Four curves for different solid feed rates again tend to merge into one curve only for each bed hold up. For fluidizing velocities greater than 2.5 u_{mf} correlation of the form given by eq. 6.11 was tried and the values of constants were determined by regression as earlier. The values of constants b, c and d and the error in calculated values for each bed hold up are given below:

(1) for h_D = 2.0

$b = 0.395$, $c = -0.0947$ $d = 0.00832$

Per cent error = $\pm$ 10

(2) for h_D = 2.0

$b = 0.285$, $c = 0.0484$ $d = 0.00359$

Per cent error = $\pm$ 10

The final conclusions from the experiments on 0.393 mm and 0.724 mm combination of particle sizes in the feed are summarized below:

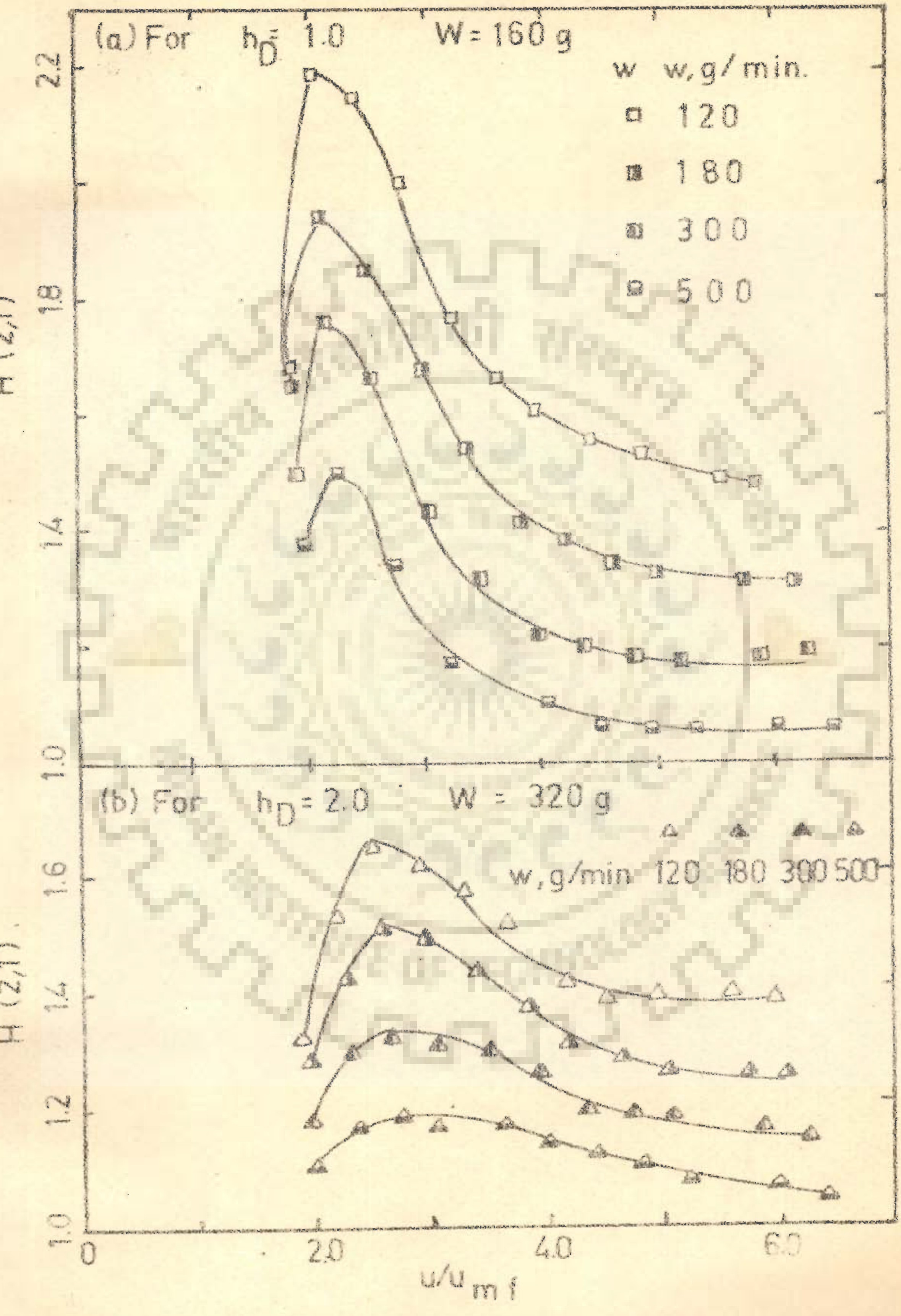


FIG. 6.30 EFFECT OF SOLID EEED RATE ON HOLD UP RATIO (For h_D = 1.0 AND 2.0)

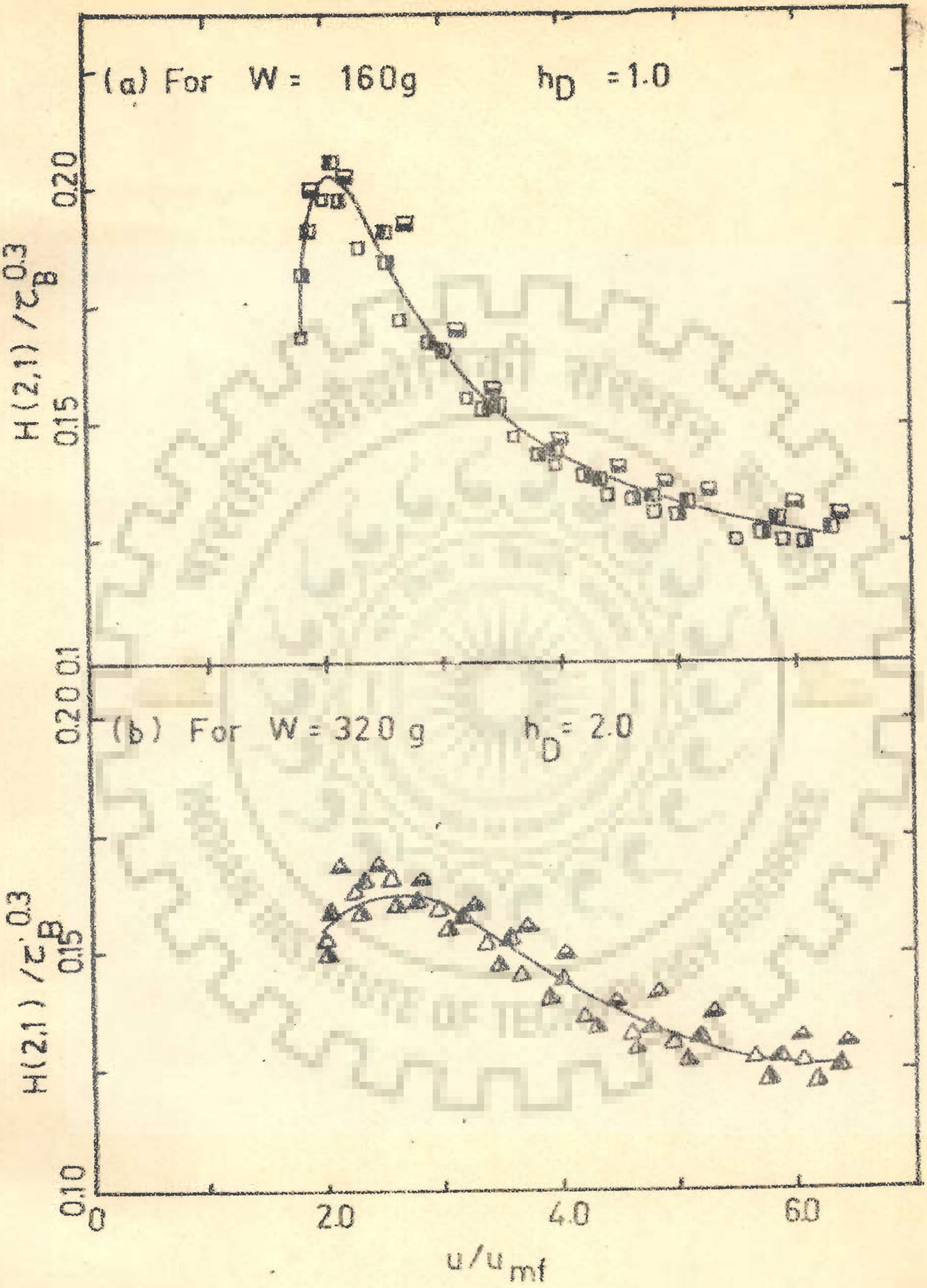


FIG. 6.31 CORRELATION OF H(2,1) AND τ_B FOR h_D = 1.0 AND 2.0

(1) The hold up ratios increase with increase of u/u_{mf} to about 2.5. With further increase in the value of u/u_{mf}, the hold up ratios experience a sharp fall.

(2) There is no effect of feed composition on hold up ratio provided the dimensionless form u/u_{mf} is used to represent air flow rate.

(3) The effect of solid feed rate on hold up ratio can be represented by a second order polynomial, Equation 6.11, for u/u_{mf} values higher than 2.5, and the constants of the polynomial obtained by regression analysis are given below:

(i) for h_D = 0.5

$b = 0.223, \; C = -0.00289, \; d = 0.00162.$

Percent error = ± 10

(ii) for h_D = 1.0

$b = 0.395, \; c = -0.0947, \; d = 0.00832$

Percent error = ± 10

(iii) for h_D = 2.0

$b = 0.285, \; c = -0.0484, \; d = 0.00359$

Percent error = ± 10

6.3.2 Hold up ratios for size combination of 0.724 mm and 0.96 mm particles

The results of experiments for determining hold up ratios for the size combination of 0.724 mm and 0.96 mm particles in 1:1 ratio of the two sizes, given in Tables E-18 to E-21 for bed hold ups of 90 g, 160 g, 320 g and 620 g, corresponding to h_D = 0.5, 1.0, 2.0 and 4.0 respectively are plotted in Fig. 6.32. The qualitative nature of

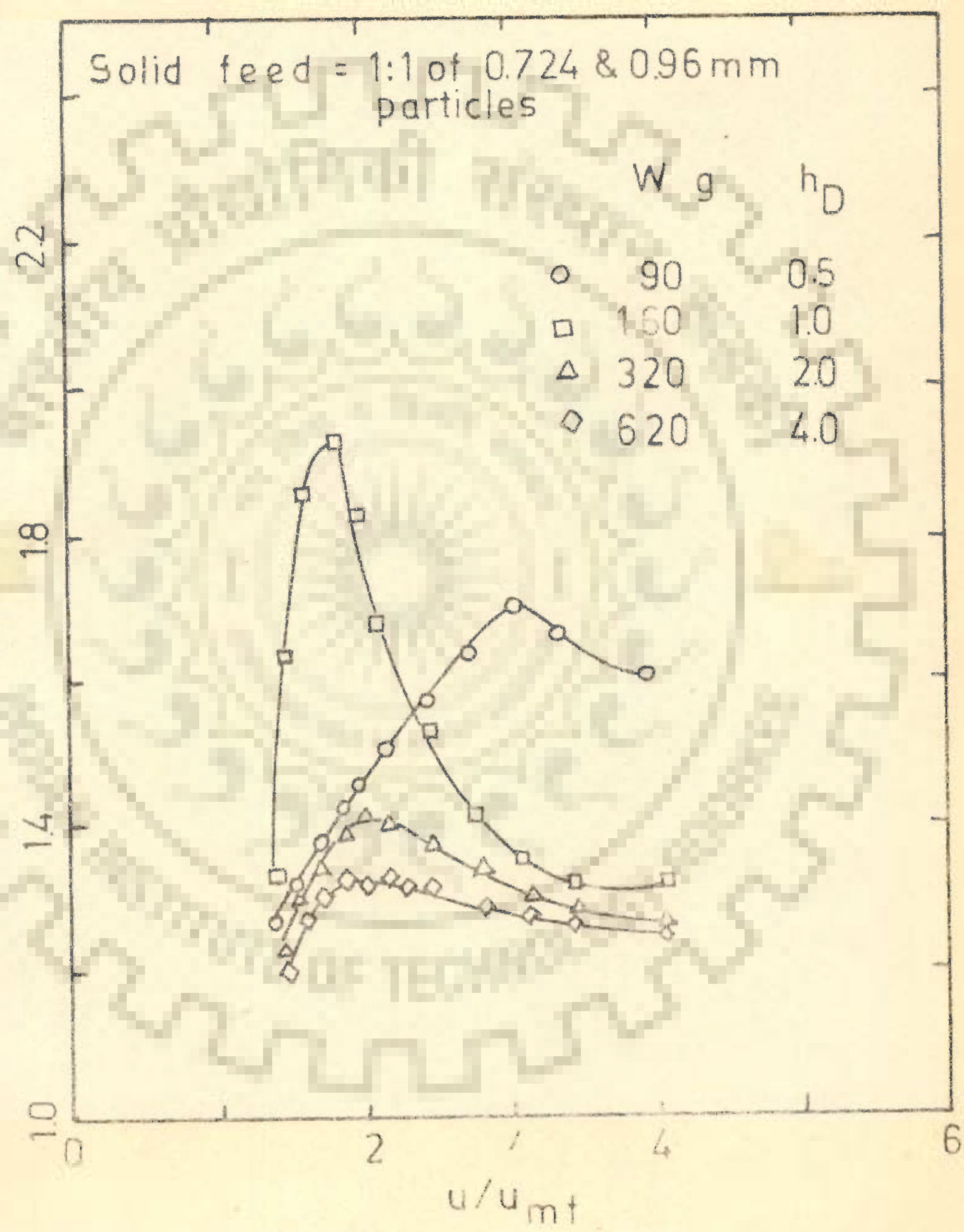


FIG.6.32 EFFECT OF BED HOLD UP ON HOLD UP RATIO H(3,2)

curves is almost the same as that for size combination of 0.393 and 0.724 mm particles as observed earlier in Fig. 6.23. At a bed hold up corresponding to $h_D = 0.5$, curve first increases till about 3.0 u_{mf} and then drops slightly. For a bed hold up corresponding to $h_D = 1.0$ the curve starts over the first curve and H(3, 2) increases very sharply till 2.0 u_{mf}, but then it drops again sharply. For higher hold ups, the curves remain below the curves for $h_D = 0.5$ and 1.0. The H(3, 2) values are in general found to be lower than H(2, 1) values. Figs. 6.33 (a) and 6.33 (b) show the data given in Tables E.22 and E.23 and Tables E.24 and E.25 respectively for different bed hold ups and solid feed rates. For the decreasing portion of the curves, the parallel nature is again observed as discussed earlier in Figs. 6.24 and 6.27. Using the same value for the exponent, n = 0.3, in Equation 6.3, the calculated values of $H(3,2)/\tau_B^{0.3}$ are plotted as a function of u/u_{mf} in Figs. 6.34(a) and 6.34(b). The three curves of Fig. 6.33 tend to come closer but they do not merge into a single curve. Further their relative positions is also found to reverse, that is, curve for highest solid feed rate is above the curve for lower solid flow rate. When the values of H(3, 2) were plotted as a function of $_B$ for different values of u/u_{mf} on log-log paper, the value of constant n was found to be 0.2. Figs. 6.34 (c) and 6.34(d) also show the calculated values $H(3,2)/\tau_B^{0.2}$ as a function of u/u_{mf}. A much better super-imposition of the decreasing portion of the theee curves is observed. A second order polynomial

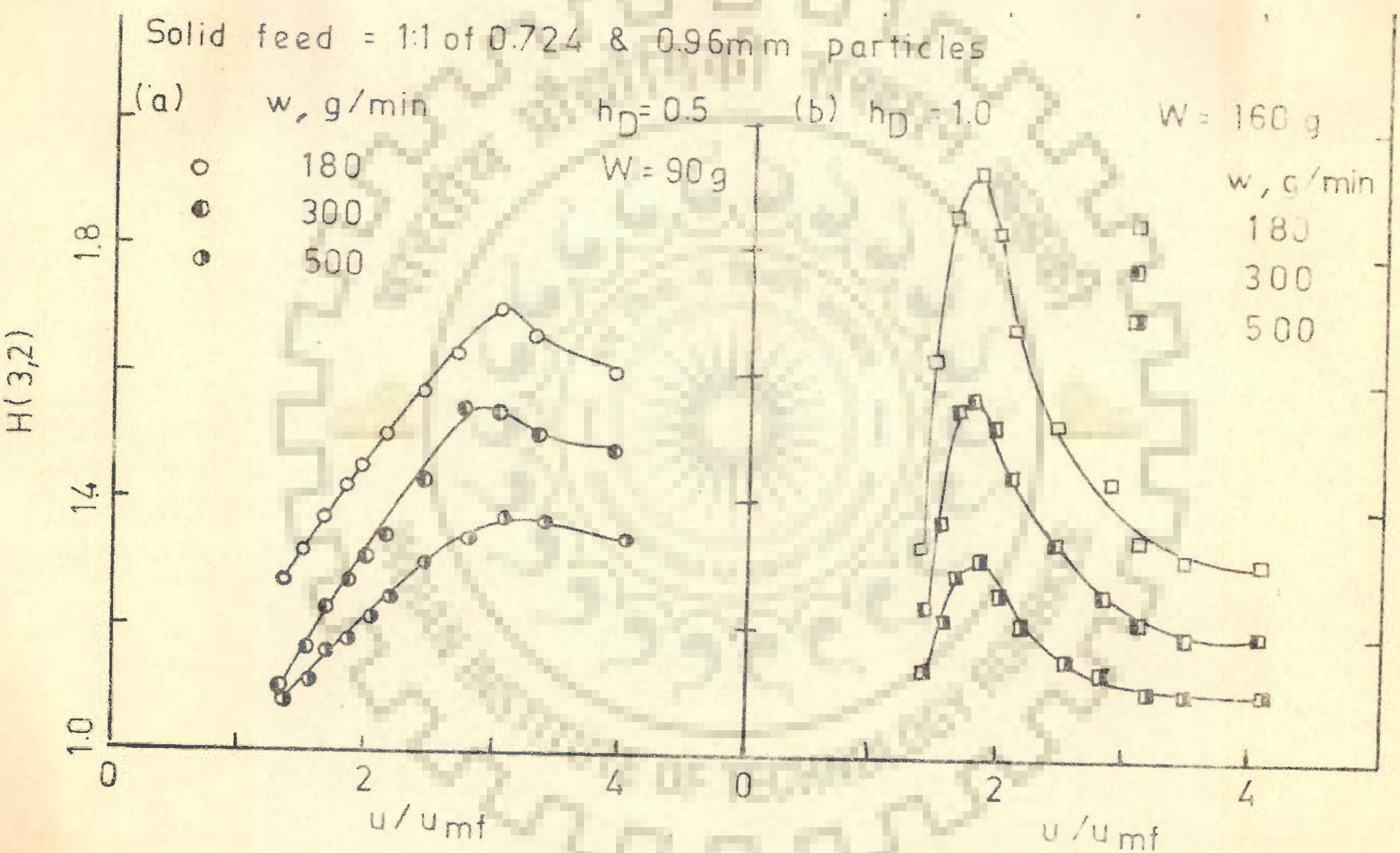


FIG. 6.33 EFFECT OF SOLID FEED RATE ON HOLD UP RATIO, H(3,2)

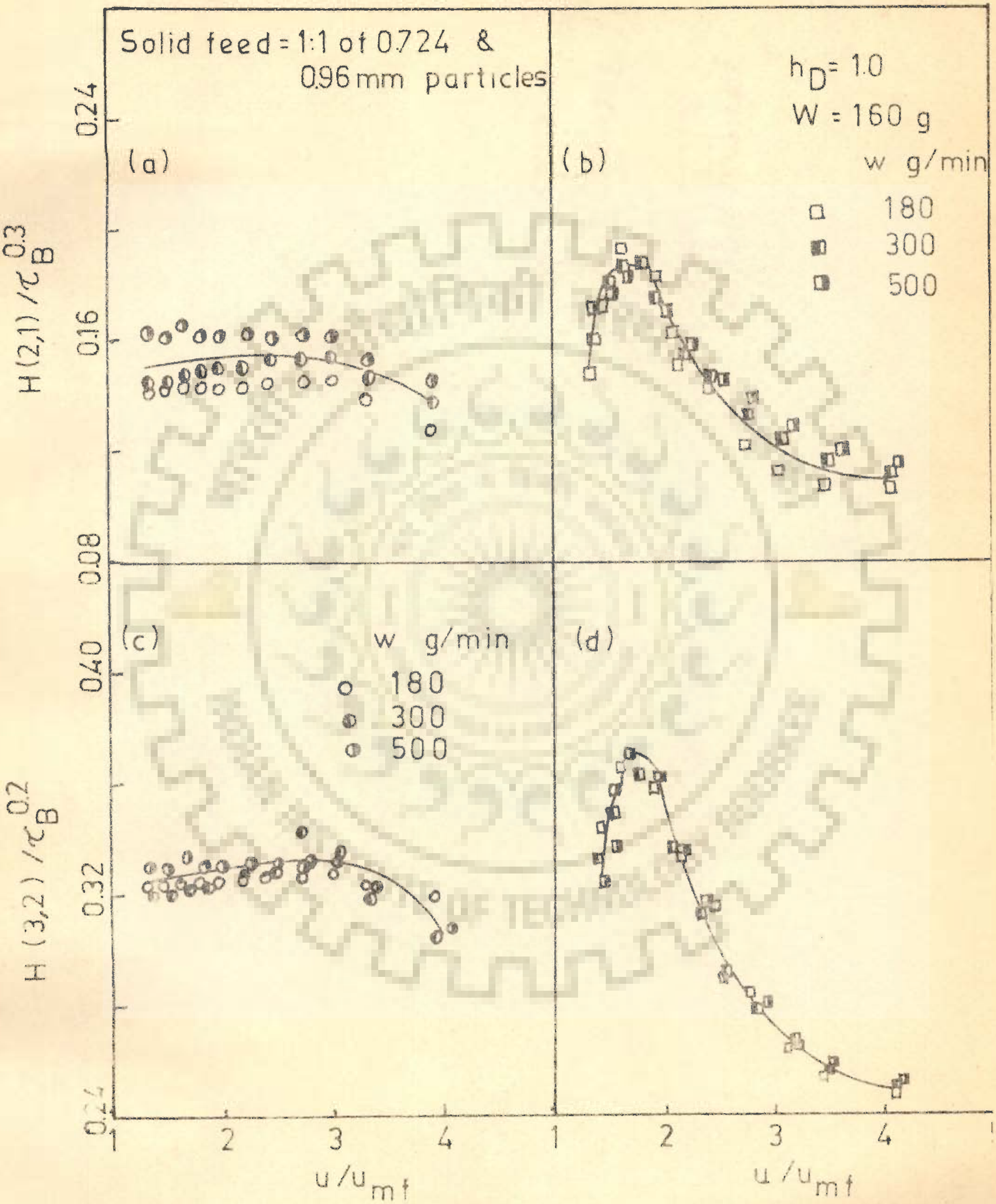


FIG. 6.34 CORRELATION OF H(3,2) WITH τ_B

$$\frac{H\,(3,2)}{(\tau_B)^{0.2}} = b + c\ u/u_{mf} + d(u/u_{mf})^2 \qquad \text{...(eq. 6.12)}$$

is used to correlate the decreasing portion of the curve using regression analysis and the computed values are given below:

(a) for h_D = 0.5

b = 0.329, c = 0.0213, d = -0.00702

Percent error = $\pm$ 3.0

(b) for h_D = 1.0

b = 0.619, c = -0.186, d = 0.0235

Present error = $\pm$ 4.0

The experimental data for different feed composition of the solid particles, tabulated in Tables E.26 and E.27 is shown in Fig. 6.35 for H(3,2) as a function of u/u_{mf} for three feed compositions. Again it is observed that the feed composition has insignificant influence on H(3,2) provided u/u_{mf} is used as the correlating parameter instead of air flow rate.

The important conclusions from the experiments on 0.724 mm and 0.96 mm combination of particle sizes are:

(1) The trend of the variation of H(3,2) as a function of u/u_{mf}, w, h_D and solid feed composition is qualitatively very much similar to the one obtained for the combination of 0.393 mm and 0.724 mm particles.

(2) A second order polynomial, Equation 6.12, is suitable to represent the experimental results for u/u_{mf} values higher than 2.5 and the constants of the polynomial are as given below:

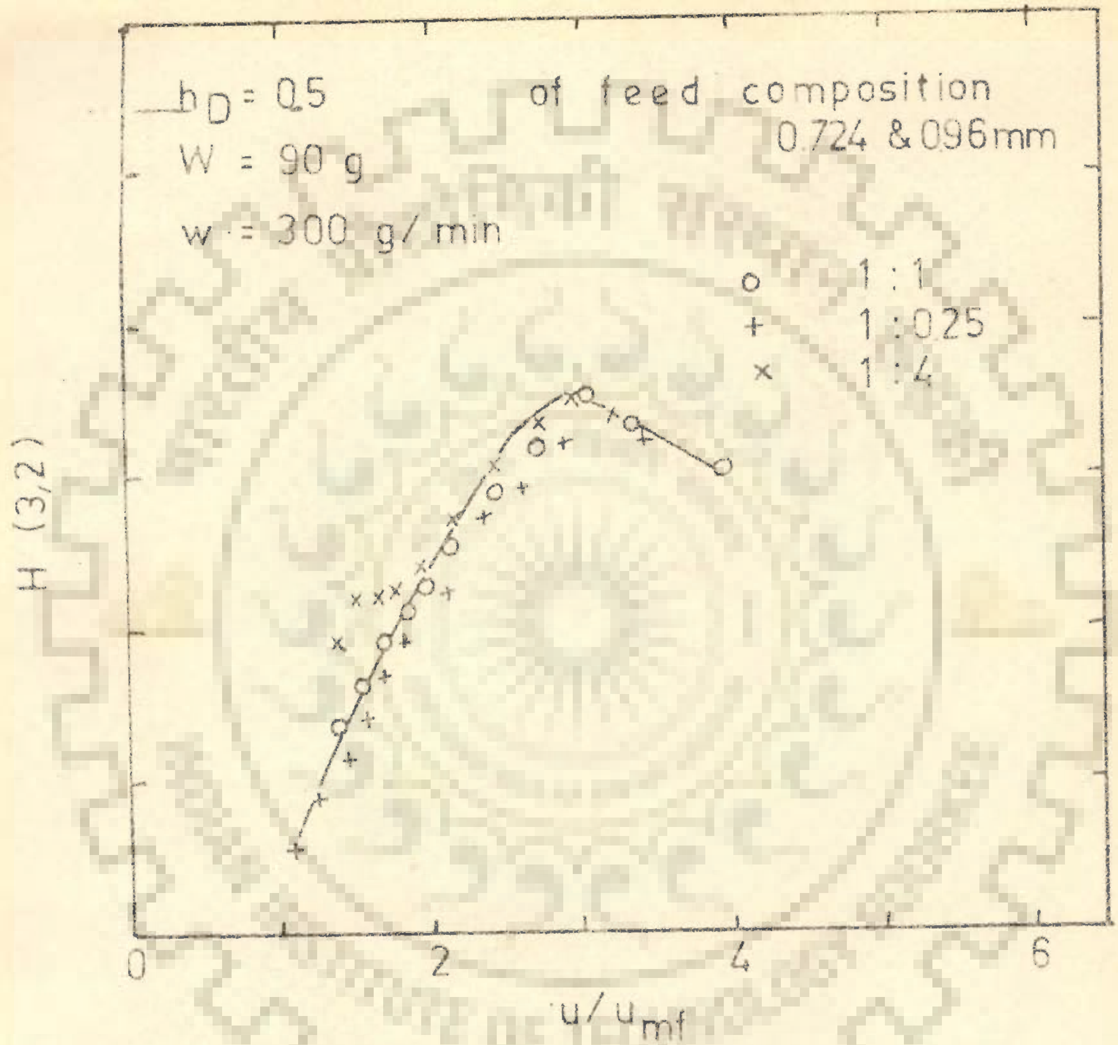


FIG.6.35 EFFECT OF SOLID FEED COMPOSITION ON HOLD UP RATIO

(i) for h_D = 0.5

$b = 0.329$, $c = 0.0213$, $d = -0.00702$

Percent error = ± 3.0

for h_D = 1.0

$b = 0.619$, $c = -0.285$, $d = 0.0235$

Per cent error = ± 4.0

6.5 HOLD UP RATIOS FOR COMBINATION OF 0.393 mm, 0.724 mm AND 0.96 mm PARTICLES

After noting the existence of different hold ups for different size particles in fluidized beds for feeds containing two size particles, it was considered more desirable to investigate the effect for feed containing particles of three different sizes because in actual industrial fluidized beds, feeds with wide size distribution are used. It would have been more meaningful to use feeds of continuous size distributions, but due to experimental limitations feeds with only three sharp cut sizes, 0.393 mm, 0.724 mm and 0.96 mm were prepared. The experimental results with feeds of three sizes in the ratio of 1:1:1 for three bed hold ups are given in Tables F-1 to F-3. Since there are three particle sizes involved here, there will be two hold up ratios, corresponding to each run, one between large and middle size particles, H(3,2), and the other between middle and small size particles, H(2,1), as defined in Equations 6.1 and 6.2. The data of Tables 6.1 and 6.2 is plotted in Fig. 6.36 and from this figure it is clear that, in general, the value of H(3,2) are more than the values of

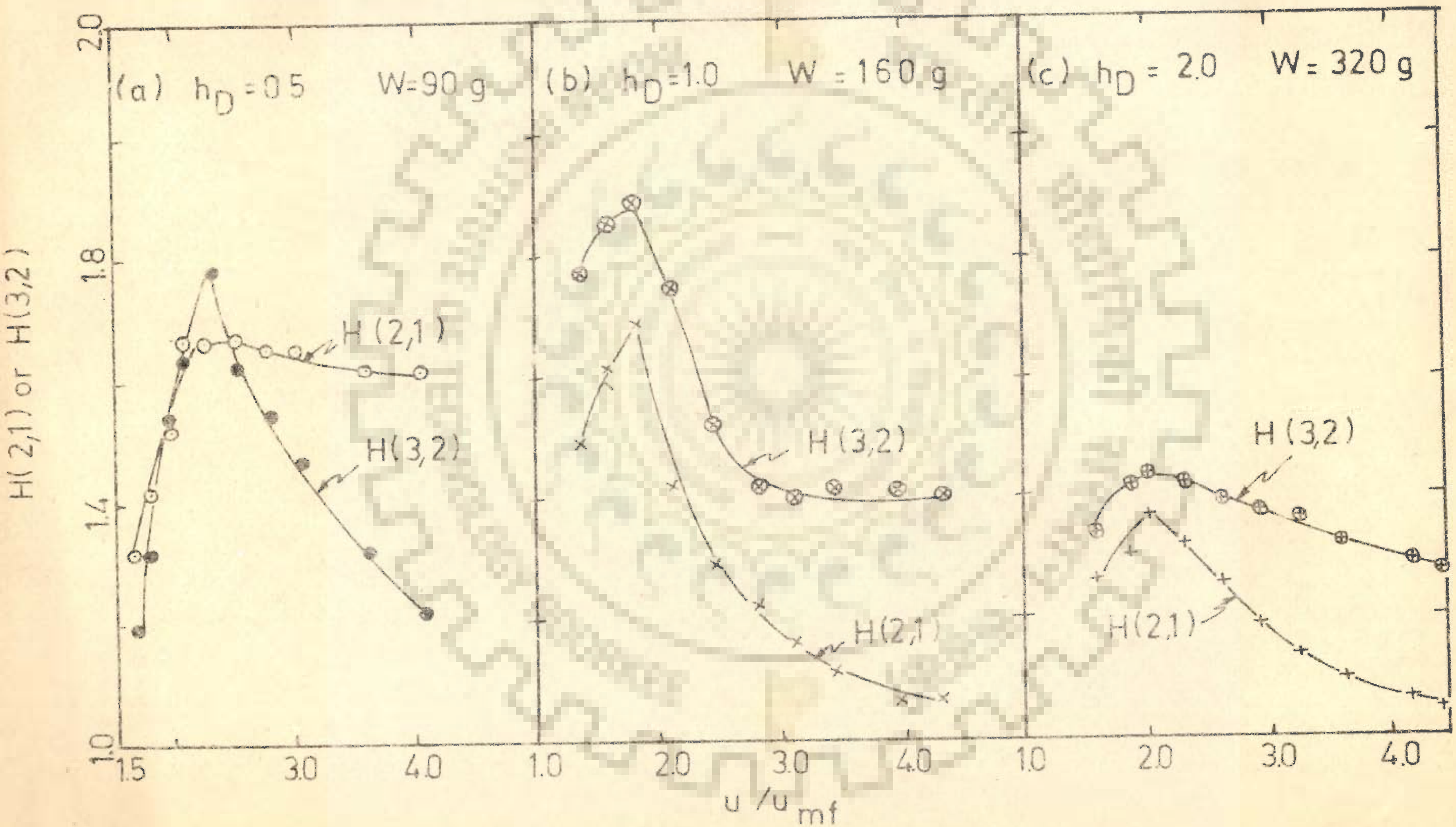
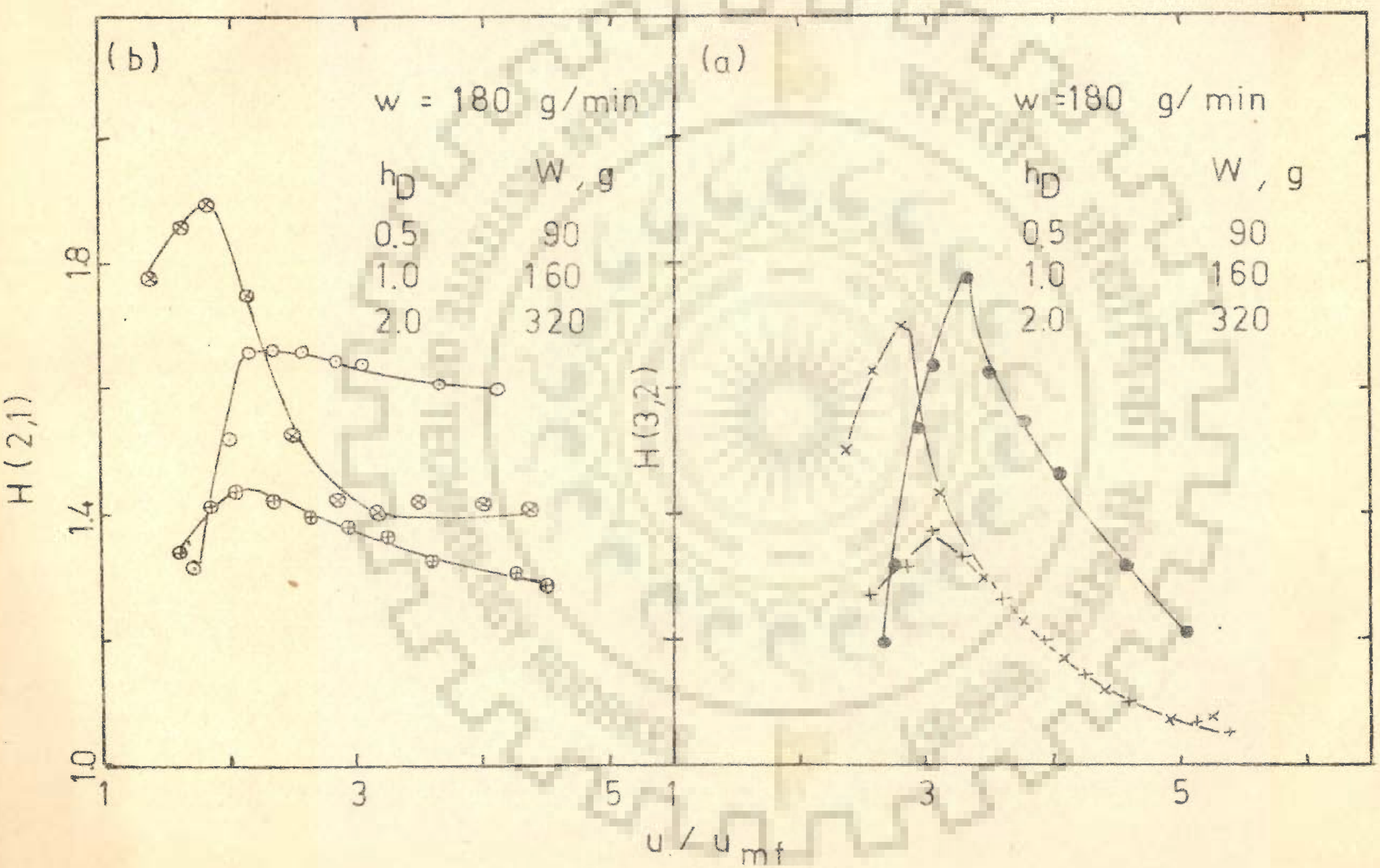


FIG 6.36 EFFECT OF SOLID HOLD UP ON HOLD RATIOS FOR MIXED FEEDS OF THREE SIZES

H(2, 1) except for the bed hold up corresponding to h_D = 0.5, the values of H(2, 1) are higher than H(3, 2) for the u/u_{mf} values in between 2.0 and 2.5. Contrary to this, with feeds of only two sizes, the value of H(2, 1) are in general more than the values of H(3, 2), Figs. 6.23 and 6.32. It appears that there is some kind of interaction when a wider distribution of feed sizes are used and there is a tendency of hold up ratios of large particles to increase due to the presence of smaller particles. The trend of the increase upto 2.5 u_{mf} and decrease with further increase of u/u_{mf} in the values of H(3, 2) and H(2, 1) with increase in air rate is observed here as well.

For feeds with three particle sizes H(2,1) and H(3, 2) are plotted as a function of u/u_{mf} for all bed hold ups in Figs. 6.37(a) and 6.37(b) respectively. Comparison of Figs. 6.37(a) with Fig. 6.23 show that the trend of H(2, 1) variation is same in both the cases H H(1, 2) also vary in the same way but the numerical values of H(2,1) and H(3, 2) are lower in the experiments of mixed feeds with three size particles than for feeds with two size particles. This shows that for mixed feeds with particles of many sizes, the values of hold up ratios are likely to reduce due to mutual interaction. In feeds of continuous size distribution the effect may be still less although this still needs experimental verification.

The effect of solid feed rates on the hold up ratios H(3, 2) and H(2, 1) for two bed hold ups is shown Fig. 6.38 using the data given in Tables F.4 to F.7. A cross plot for H(2, 1) as a



(SOLID FEED 1:1:1 OF 0.393 0.724 & 0.96 mm PARTICLES)
FIG. 6.37 H(2,1) AND H (3,2) FOR DIFFERENT SOLID HOLD UPS

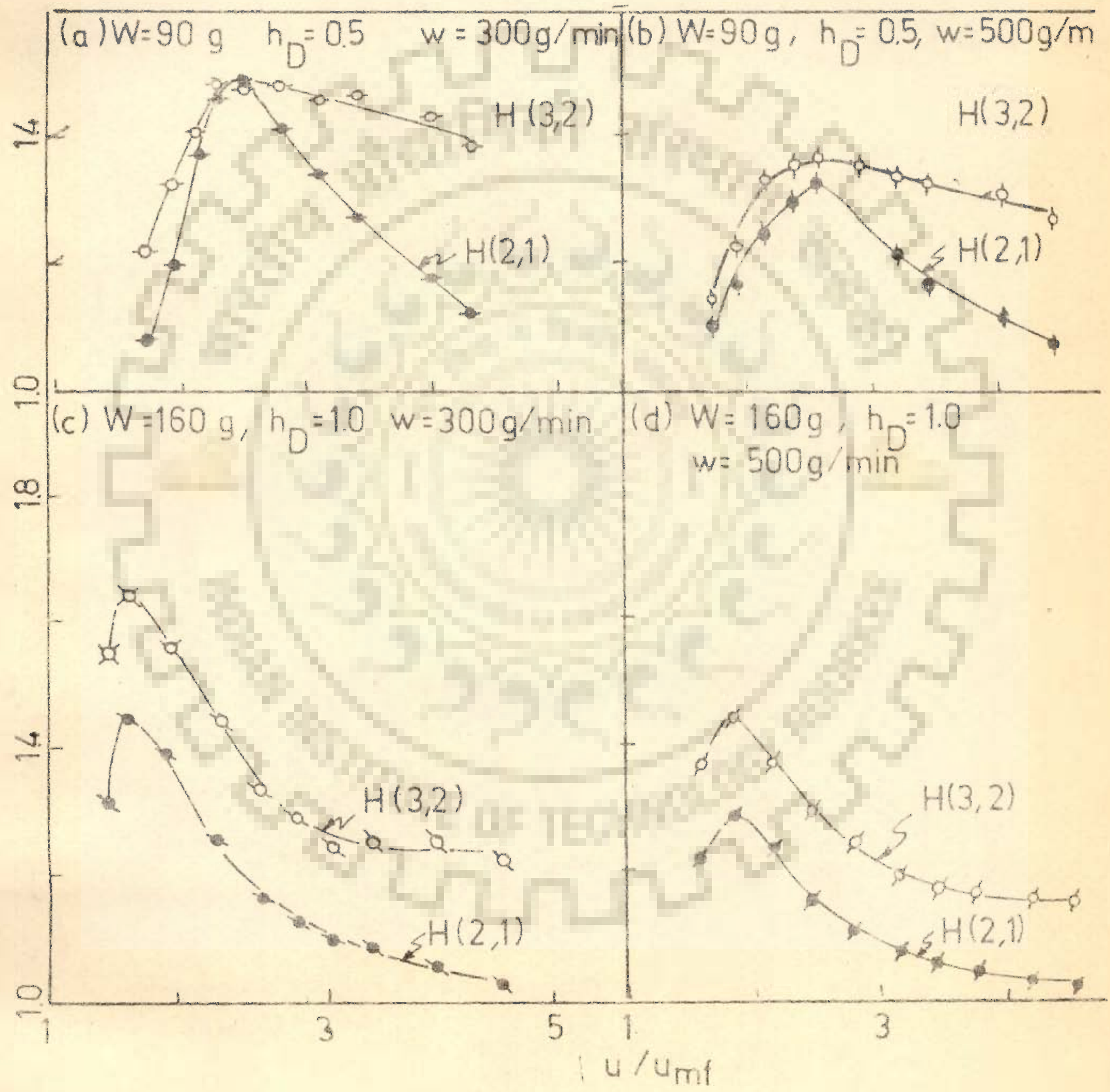


FIG. 6.38 EFFECT OF SOLID FEED RATE ON HOLD UP RATIO FOR FEEDS OF THREE SIZES

function of u/u_{mf} for different solid feed rates is shown in Fig. 6.39(a) and corresponding plot for H(3, 2) is shown in Fig. 6.39(b). The parallel nature of the decreasing portion of the curves is not as good as observed in Figs. 6.27 and 6.33. Again better superimposition of the three curves is obtained by using the value of n as 0.2 in Equation 6.3. The value $H(2,1)/(\tau_B)^{0.2}$ and $H(3,2)/(\tau_B)^{0.2}$ are plotted as a function of u/u_{mf} in Fig. 6.40. The three curves of Fig. 6.39 tend to merge into a single curve and a second order polynomial, Equation 6.12, is used to correlate the decreasing portion of the curves using regression analysis and the computed values of constants are given below:

(i) for hold up ratio H(2, 1)

(a) $h_D = 0.5$

$b = 0.774$, $c = -0.225$, $d = 0.0248$

Percent error = $\pm$ 5.0

(b) $h_D = 1.0$

$b = 0.482$, $c = -0.0999$, $d = 0.01$

Percent error = $\pm$ 10.0

(ii) for hold up ratio H(3, 2)

(a) $h_D = 0.5$

$b = 0.512$, $c = -0.709$, $d = 0.00595$

Percent error = $\pm$ 3.0

(b) $h_D = 1.0$

$b = 0.523$, $c = -0.101$, $d = 0.0106$

Percent error = $\pm$ 5.0

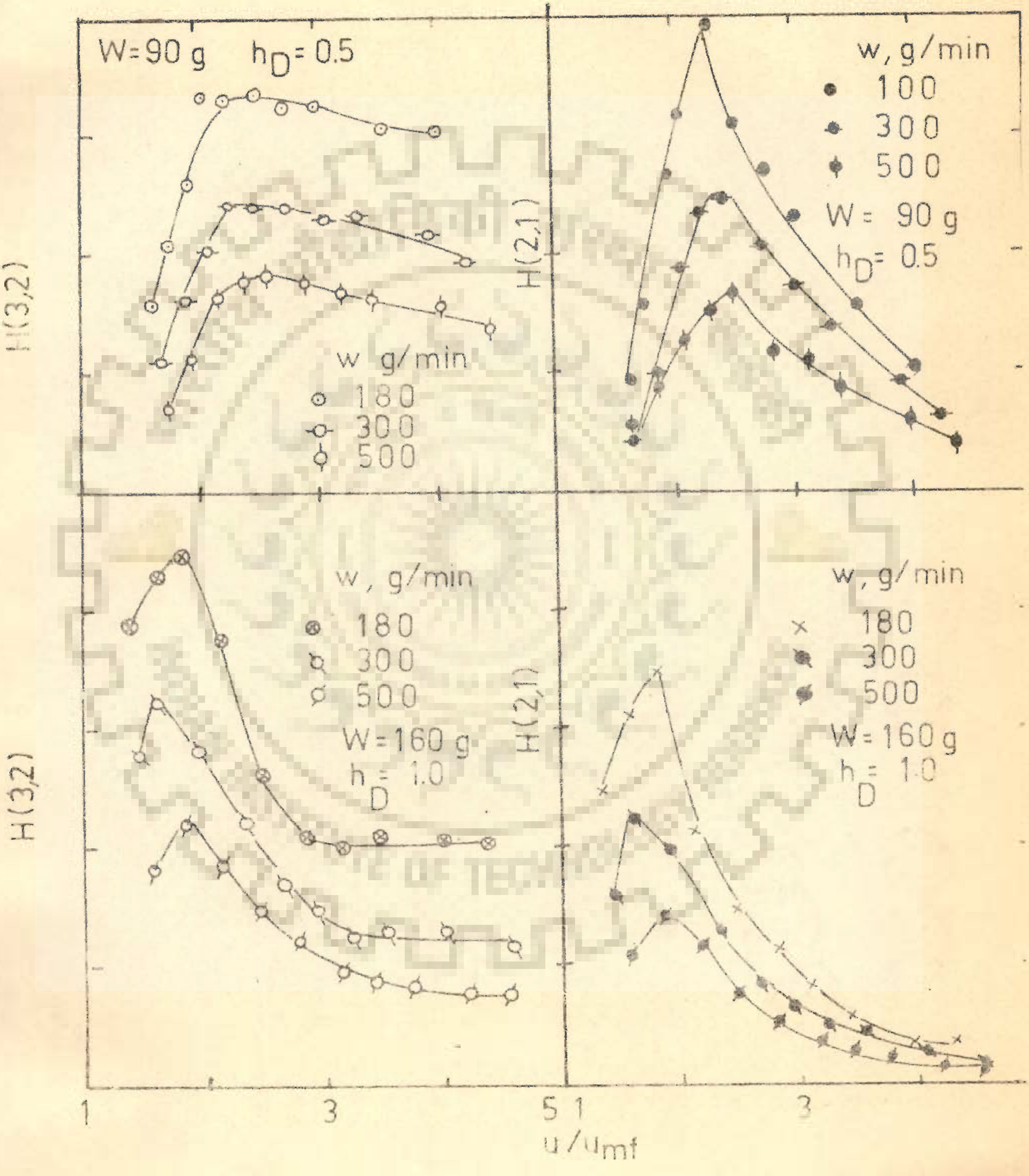


FIG.6.39 H(2,1) & H(3,2) vs u/u_{mf} FOR CHANGING SOLID FEED RATES

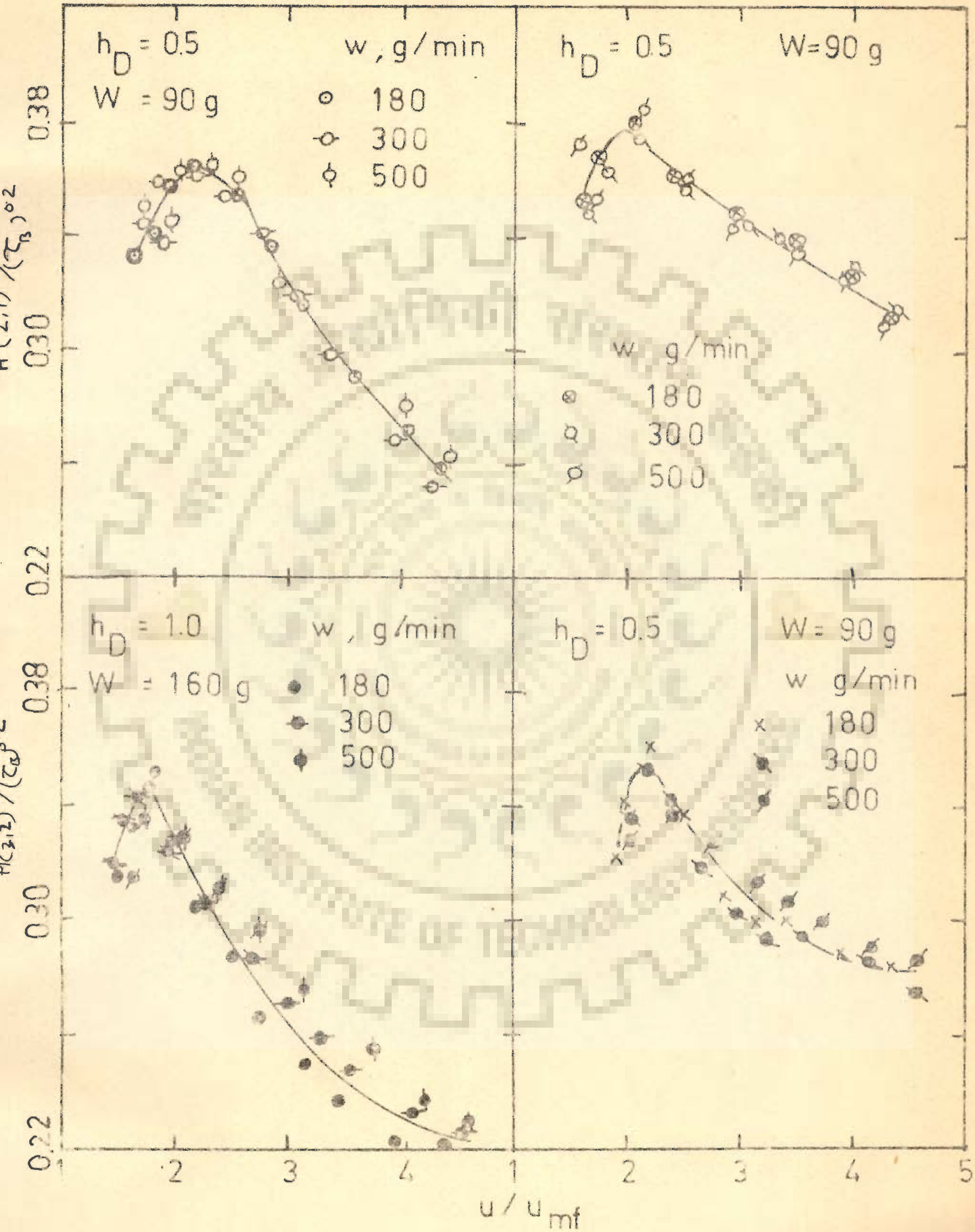


FIG. 6.40 . CORRELATION OF H(2,1) & H(3,2) WITH τ_B

A unified equation for all the runs is not attempted because much more experimental work is required to arrive at an equation which will take into account the effect of particle sizes and their interactions.

The data of Tables F. 8 to F. 11 of hold up ratios for different ratios of three particle sizes are plotted in Fig. 6.41 to show the effect of particle size composition in the solid feed. It can be seen that the scatter is slightly more in this figure as compared to Fig. 6.26 yet the points are quite close.

The important conclusions from experiments on feeds of three particle sizes are:

(i) The trend of the variation of H(2,1) and H(3,2) as a function of u/u_{mf}, w, h_D and solid feed composition obtained for solid feed combinations of three particle sizes is qualitatively very much similar to the one obtained for the combination of feeds of two particle sizes.

(ii) The hold up ratios H(3,2) and H(2,1) for mixed feeds of three particle sizes are lower than those with feeds of two particle sizes due to a possible mutual interaction of the particles.

(iii) A second order polynomial, Equation 6.12, is suitable to represent the experimental results for u/u_{mf} values higher than 2.5 for both H(2,1) and H(3,2). The constant of the polynomial are given below:

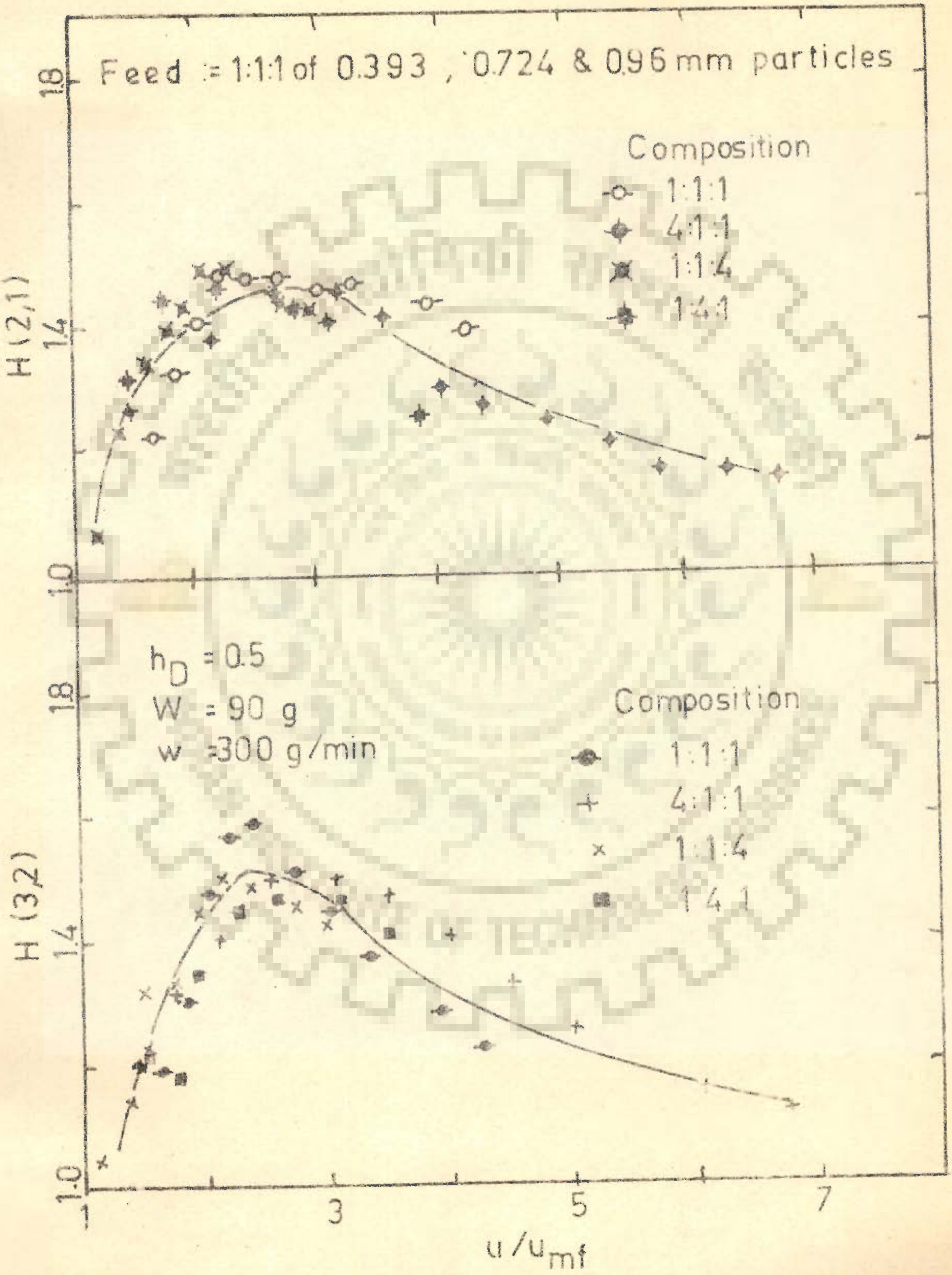


FIG.6.41 EFFECT OF SOLID FEED COMPOSITION O HOLD UP RATIOS

(1) for hold up ratio H(2,1)

(a) $h_D = 0.5$

$b = 0.774, \quad c = -0.25 \quad d = 0.0248$

Maximum error = $\pm$ 5.0%

(b) $h_D = 1.0$

$b = 0.482, \quad c = -0.0999, \quad d = 0.01$

Maximum error = $\pm$ 10.0 %

(2) for hold up ratio H(3,2)

(a) $h_D = 0.5$

$b = 0.512, \quad c = -0.709, \quad d = 0.00595$

Maximum error = $\pm$ 3.0%

(b) for $h_D = 1.0$

$b = 0.523, \quad c = -0.101, \quad d = 0.0106$

Maximum error = $\pm$ 5.0%

In order to obtain a better insight into the mechanism responsible for different mean residence times for different size particles, a study of the radial and axial distribution of solid composition in the fluidized bed was undertaken and these results are discussed in the following section.

6.6 RADIAL AND AXIAL DISTRIBUTION OF PARTICLES IN FLUIDIZED BEDS

The results presented earlier in Section 6.5 reveal a very important and useful phenomena that different size particles spend different mean residence times in the fluidized beds and yet they remain totally backmixed. This phenomena has industrial applications because in solid fluid reactions, it is always desirable that small particles stay in the bed for lesser time as compared to larger particles. In the present investigation this phenomena is distinctly observed and it is, therefore, important to search a mechanism to explain these results for the proper understanding of factors contributing to this phenomena.

The fact that the mean residence time of small particles is lower as compared to that of larger particles in the fluidized bed suggest that for an initial bed composition of 1:1 of both size particles, small particles have a tendency of going to the exit preferentially over the larger particles. Urabe, et. al. (104) studied the axial distribution of size and voidage in continuous fluidized bed of sand at different gas velocities by using an overflow tube for solid removal and observed that the size distribution is roughly constant within the main zone of constant voidage, however, the upper falling density zone becomes progressively richer in fines. In addition, the main zone has less fines at high velocity, indicating that fines are removed rapidly at higher velocity. The conclusions are confirmed by Thomas et. al. (99) and Hiraki et. al. (44). If the concentration of large particles is relatively more at the bottom of the bed, and if it is assumed that the particles are uniformly distributed radially, then the probability of large particles entering the solid discharge system is more than that for smaller particles if the solid discharge is achieved at the bottom of the bed and vice-versa. This suggests that with solid discharge from the bottom of the bed, the mean residence time of the larger particles should be slightly lower than that of smaller particles and for overflow type of solid discharge, the mean residence time of smaller particles should be slightly more than that of larger particles. For the experimental set up used in the present work, the exit connection was at the bottom of the bed so holdup ratios of less than unity should have been obtained. But the experimental results indicated hold up ratios more than unity. If it is hypothesized that the smaller particles in spite of their relatively smaller concentration at the bottom of the bed, tend to enter the solid discharge system in preference to the larger particles because of their higher relative velocities, then the solid discharge system position anywhere in the bed or even with an overflow type of solid discharge should always result in hold up ratios of more than unity because the smaller particles will have higher relative velocities all over the bed. Besides, with the above hypothesis, a change in the size of the solid discharge system should not effectively change the values and the trend of hold up ratio curves. The data for the three opening sizes of solid discharge system, namely 13 mm, 10 mm and 6 mm, given in Table G. 1, are plotted in Fig. 6.42 and it is seen from the figure that the hold up ratios in general are slighly more for smaller opening size but the difference in the hold up ratios is too small to draw any conclusions. Therefore, the results of overflow and downcomer arrangement system as shown in Fig. 4.7(a) are more important to test the above hypothesis. The experimental results

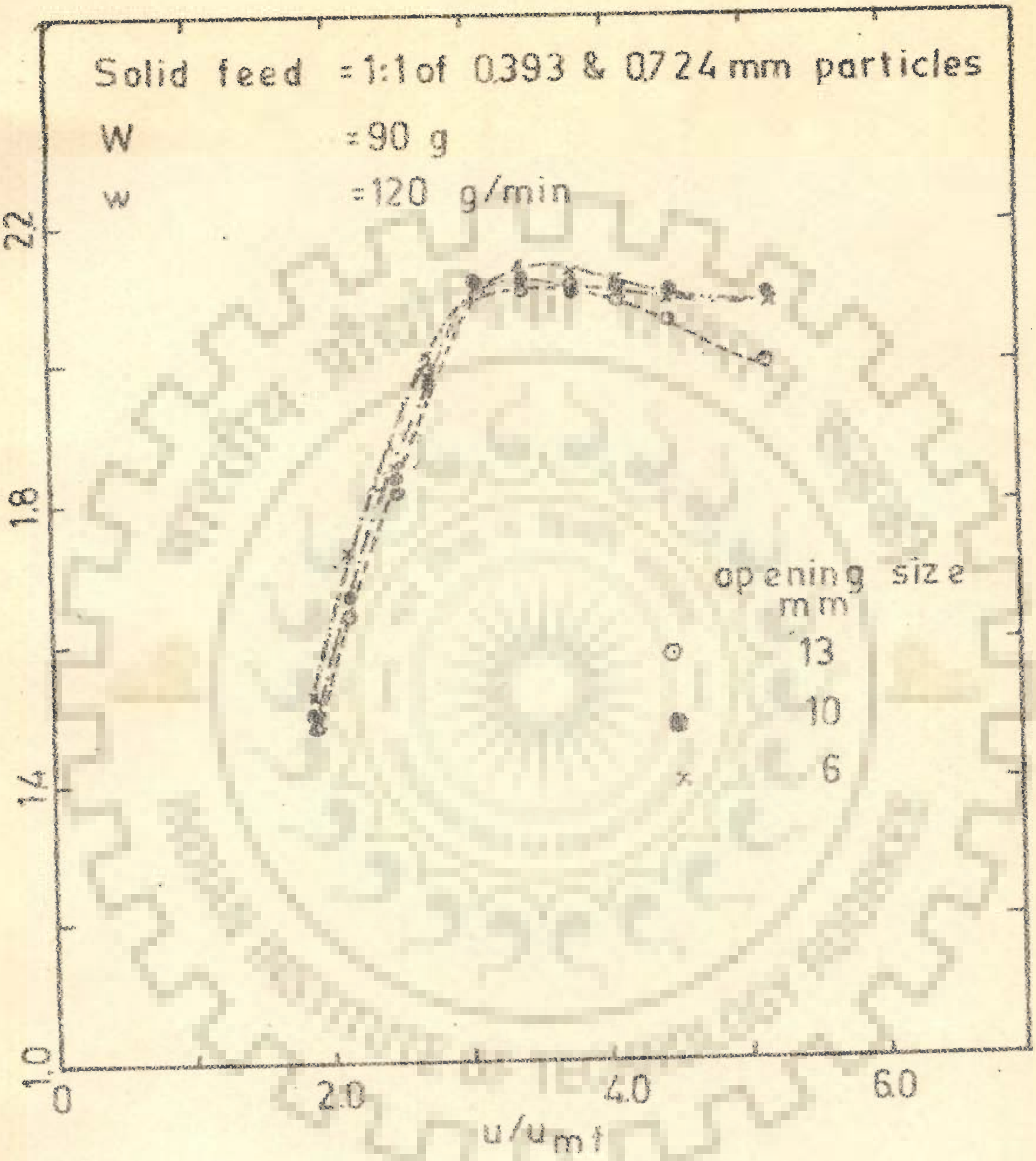


FIG.6.42 EFFECT OF OPENING SIZE ON HOLD UP RATIOS

with such system given in Table G. 2 for feed composition of 1:1:1 of three size particles and two solid feed rates are plotted in Fig. 6.43. It is clear from this plot that the values of hold up ratios are only slightly more than unity for such system. The decreasing trend in the value of hold up ratio with increase in u/u_{mf} is again visible in Fig. 6.43. The values of H(2,1) and H(3,2) are less than 1.1 for all values of u/u_{mf} and at u/u_{mf} values of 2.5 to 5, H(2,1) is about 1.02 and H(3,2) is close to 1.05 at solid feed rates of 180 g/min and decrease even further and reach close to unity at higher solid feed rates. It, therefore, appears that at such velocities the difference in residence times for different size particles is practically absent for overflow and downcomer arrangement. This rules out the hypothesis that the small particles tend to move to solid discharge system preferentially over the larger particles due to their larger relative velocities. A third modification as shown lin Fig. 4.7(b) gave the peripheral opening for the solid discharge with top end at about 7 cm from the bottom of the bed. Using this exit for solids either as an overflow with automatic adjustment of bed height and bed hold up or by controlling the solid discharge rate to maintain static bed height of 7.5 cm corresponding to $h_D = 1.5$, the experimental results given in Table G.3 are plotted in Fig. 6.44. It is clear from this figure that the use of this solid discharge system as an overflow exhibits the same trend in hold up ratios as with the overflow and downcomer arrangement. The hold up ratios remain close to unity for almost all air flow rates. However, with the same solid

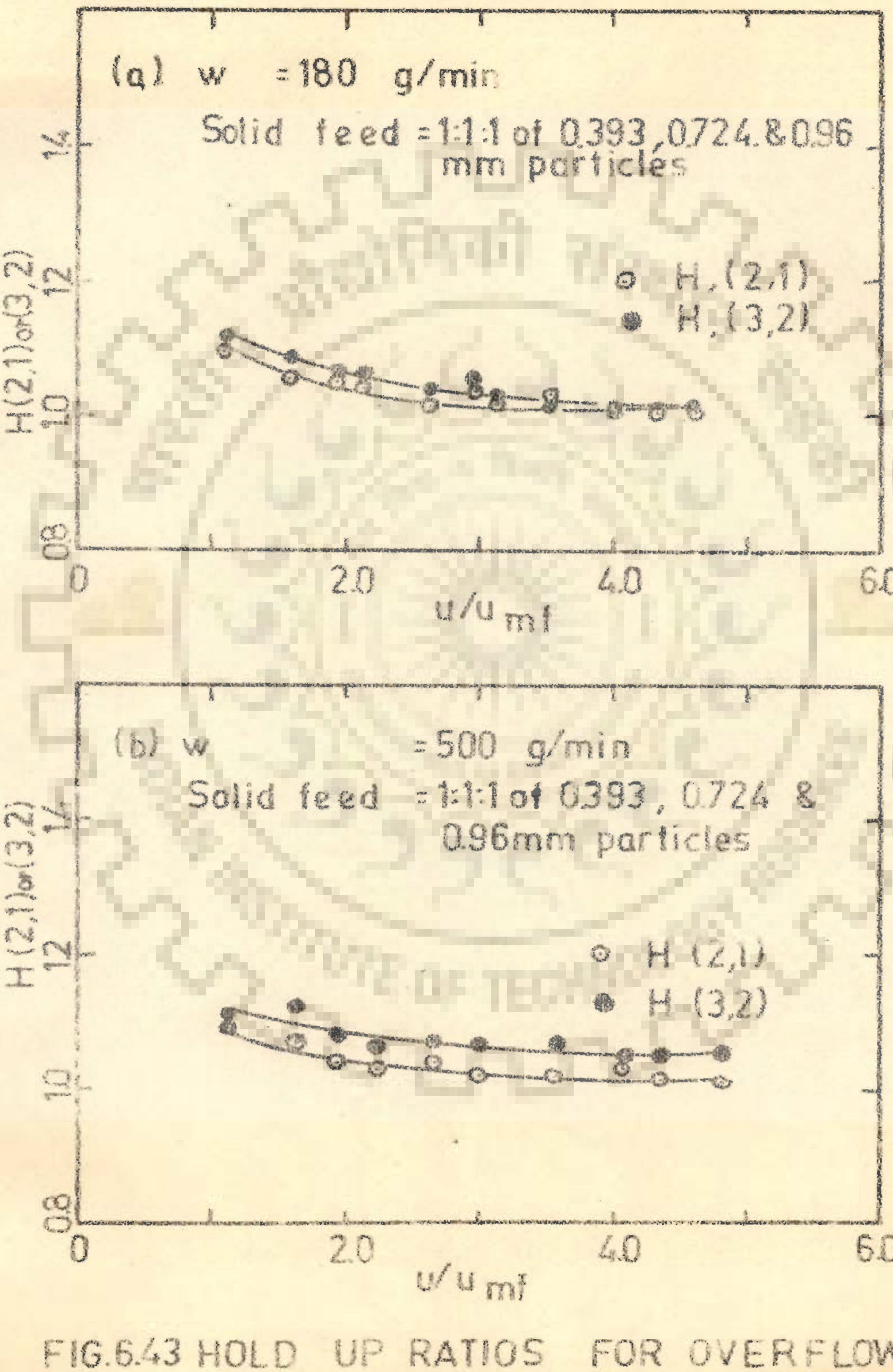


FIG.6.43 HOLD UP RATIOS FOR OVERFLOW AND DOWNCOMER ARRANGEMENT

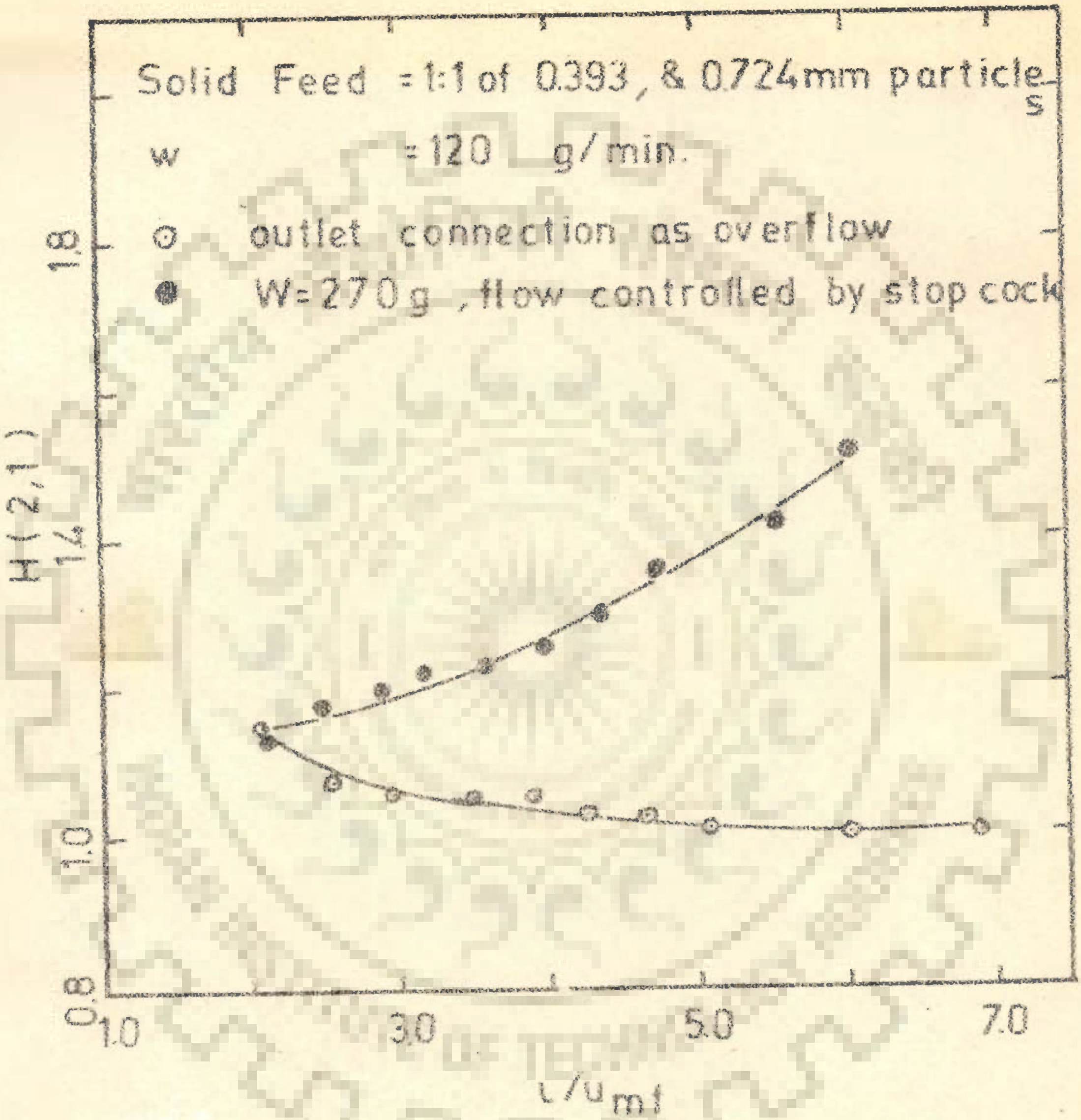


FIG.6.44 HOLD UP RATIOS FOR OUTLET CONNECTION AT 5 cm HEIGHT

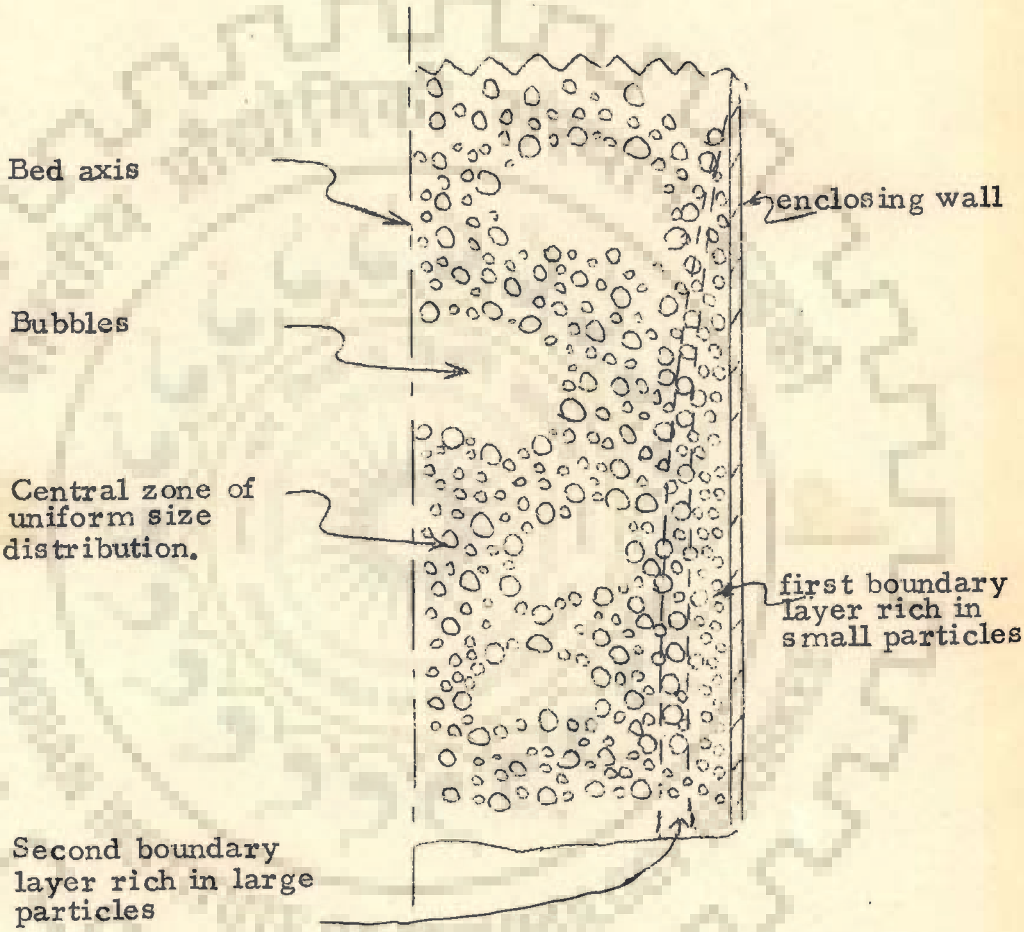
discharge system, when the bed hold up was controlled corresponding to $h_D = 1.5$, the hold up ratios show a different trend. For low values of u/u_{mf}, the hold up ratio matches with the hold up ratio of overflow system but with further increases in u/u_{mf}, hold up ratio increases consistently reaching a value of 1.5 at u/u_{mf} value of about 6.0. This indicates that the effect of different mean residence times for particles of different sizes exists only when the solid discharge system is at the periphery of the bed and the fluidized bed height is much higher than the position of the solid discharge system. In other words, when the bed height is considerably higher than the position of peripheral solid discharge system, the smaller particles tend to leave preferentially over the larger particles. This is possible only when the concentration of small particle is relatively more than larger particles near the periphery of the bed.

Kondukov et. al. (51) observed from their study of trajectories of tagged radio-active particles that particles tend to wander everywhere in the bed and that there is a definite up and down movement, the upward movement is rapid as compared to the downward movement. They also observed that particles near the surface usually remain there for a while before dipping into the bed. The fact that particles spend more time near the surface can mean that in a bed of mixed particle size, the bed composition very close to the surface of the enclosing wall may be different than at other positions in the bed. In other words, there can be a radial distribution of particle concentration in the fluidized bed near its periphery.

6.5.1 Proposed New Model for Radial and Axial Distribution of Particles in Fluidized bed

The picture on mixing of solids developed by various workers (17, 30, 77, 83, 96) indicates that the mixing of solids in gas fluidized beds is caused primarily by the disturbance of gas bubbles passing through the bed. Since the number and size of gas bubbles increase with rise in gas velocity, the extent of solid mixing also increases correspondingly. It, therefore, appears that for most of the radial position, mixing of the solids may be quite uniform in the fluidized beds but very close to the surface of enclosing wall, there may exist some gradient in particle size distribution. The air velocity in the fluidized bed has a more or less flat profile, but close to the surface there is sharp decrease of air velocity because of wall drag. The particles near the surface are not so much in vigorous motion as those in the bed. That, perhaps, is the reason why the particles tend to stay in their position near the surface as observed by Kondukov(51) and Gabor (30). When particles of different sizes are used in the bed, due to the sharp fall in air velocity near thee enclosing wall the gas velocity may drop to a value which is in between the u_{mf} values for small and large size particles. The velocity near the wall may, therefore, be enough just to keep the small particles in random motion but the larger particles may have only downward motion
rapidly
and may tend to / rapidly near the wall. As the particles of large sizes fall, they are likely to encounter some small particles with upward movement since they are still having random motion. The impact of such particles may push the larger particles away from the bed wall

and this may give rise to a kind of boundary layer of particles with higher concentration of small size particles as compared to the main bed. The size distribution near the enclosing wall may appear as shown below:



The increasing and decreasing trend of hold up ratios can be explained on the basis of this model. As the air velocity increases above u_{mf}, the velocity close to the enclosing wall also increases causing more vigorous random movement of the smaller particles. Thus the layer of particles close to the wall gets progressively depleted of large particles, and since with peripheral solid discharge system it is this layer of solid which is removed continuously, the hold up ratio also increases with air rate in the beginning. For velocities beyond 2 u_{mf}, the bed becomes a vigorously bubbling bed and this causes the central zone to spread towards the surface of the enclosing wall due to vigorous particle interaction and also the air velocity close to wall may approach a value near the u_{mf} for larger particles which may result in some random motion for larger particles also. Thus, the layer of particles close to the wall may decrease in thickness and may not be depleted of larger particles as much as at lower air rate, and, therefore, hold up ratio shows a decreasing trend with increase in u/u_{mf} beyond a value of 2 to 3. Another consequence of this model is that the thickness of the boundary layer of particles will not be uniform throughout the radial position because particles are thrown rapidly due to bubble rupture probably with only a slight excess of fines. The thickness of the boundary layer rich in smaller particles is likely to increase as we move down from the top surface and after some distance it may become stabilized into a uniform thickness. Thus, particle concentration close to the enclosing wall is influenced by thebed length. There are bound to be some end effects of the top layer and also of the bottom layer of the bed. Accordingly, the axial position of solid discharge system will have an effect on the hold up ratios. Similarly the bed height will also have an effect on the hold up ratios because the extent of the growth of the boundary layer rich in smaller particles depends on bed height.

On the basis of the model proposed, the boundary layer rich in smaller particles starts growing from the top of the bed and gets stabilized after some distance downwards. Thus if the fluidized bed height is kept constant for different air flow rates (by changing the bed hold up accordingly), an increase of air velocity may decrease the hold up ratios for the region before boundary layer has been stabilized. The experimental results given in Table G.4 on hold up ratios for different bed heights are plotted in Fig. 6.45. Fluidized bed heights depend upon air flow rates as well as solid hold up in the bed. For low air rates a higher solid hold up is required to obtain a given fluidized bed height as compared to hold up required at high air rate for the same bed height. The hold up ratios tend to achieve a constant value in the range 1.33 to 1.35 after the thickness of the boundary layer is stabilized and this is achieved at a bed height of 17 cm for air rate of 80 litres/min. and monotonically increases to a value of 24 cm for air rate of 161 litres/min. with solid discharge system located at 5 cm from the bottom of the bed. Thus, in the region of bed heights before the boundary layer thickness is stabilized, bed heights less than 17 cm, the hold up ratios will decrease with increase in air rate for a constant value of fluidized bed height. After the hold up ratio reaches the maximum value further increase in bed beights at a given air rate is achieved by increasing the solid hold up in the bed still further and hence increasing the tendency of slug formation which results in a small decrease in the hold up ratio as discussed earlier. For fluidized bed heights larger

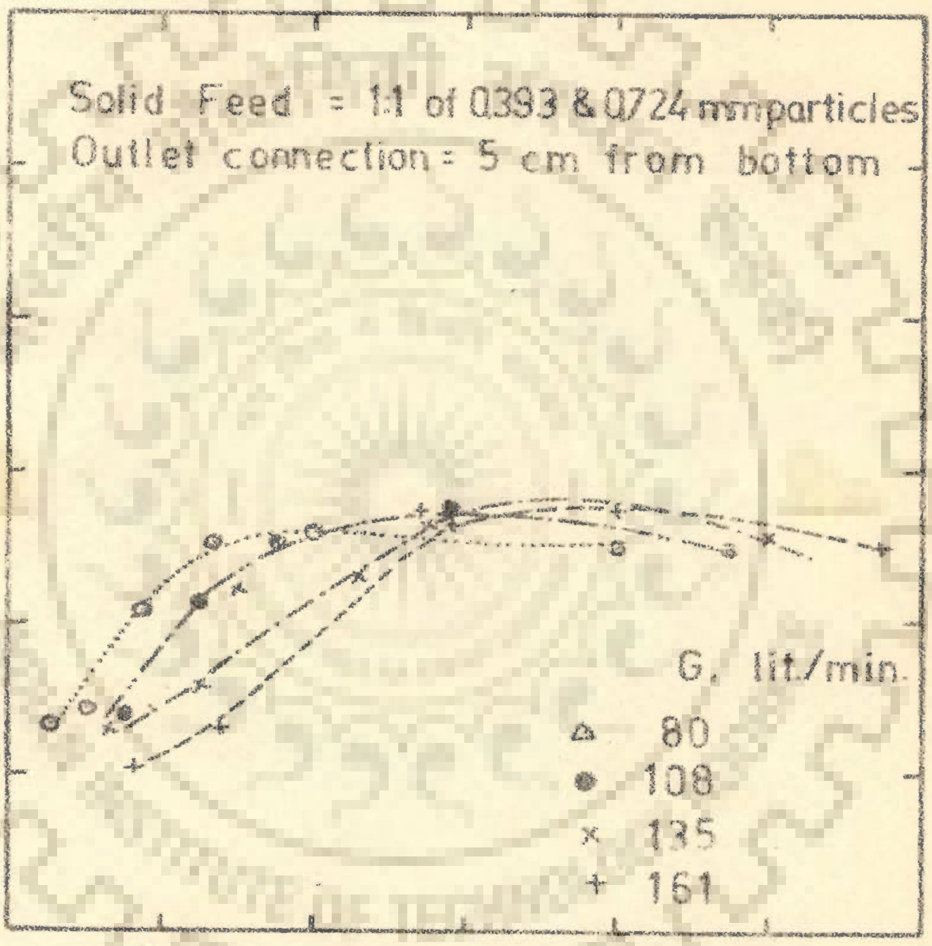


FIG.6.45 EFFECT OF BED HEIGHT ON HOLD UP RATIO

than 17 cm, the actual relative position of the curves for different air rates will depend upon the extent of boundary layer thickness, if the maximum is not reached and the extent of slugging, if the maxima is already achieved.

The existence of boundary layer rich in smaller particles close to the enclosing wall can also be established by changing the radial position of the solid discharge system from the periphery to the inside of the bed. If the solid discharge system can be placed beyond the boundary layer rich in small particles, then the hold up ratios close to unity should be obtained. Further, if the larger particles are concentrated relatively into another layer adjacent to first, because of their being pushed away from the enclosing wall, then hold up ratios below unity can also be expected for the radial position of the solid discharge system just beyond the boundary layer rich in small particles. In other words, there is a possibility of the existence of two boundary layers of particles, one close to the enclosing wall which is rich in small particles followed by another layer adjacent to the first which may be rich in larger size particles. Experimental results on such system with changing radial position of solid discharge system in the beds achieved with the help of polythene tube of suitable dimension, Fig. 4. 7(c), given in Table G. 5, are shown in Fig. 6.46 for radial positions of solid discharge system at 2 mm, 4 mm, 6 mm and 8 mm inside the surface of enclosing wall. This figure reveals very interesting trends. When the position of solid discharge system was 2 mm inside the bed, at low u/u_{mf} values, hold up ratios of less than unity are indeed obtained, confirming that there exists a

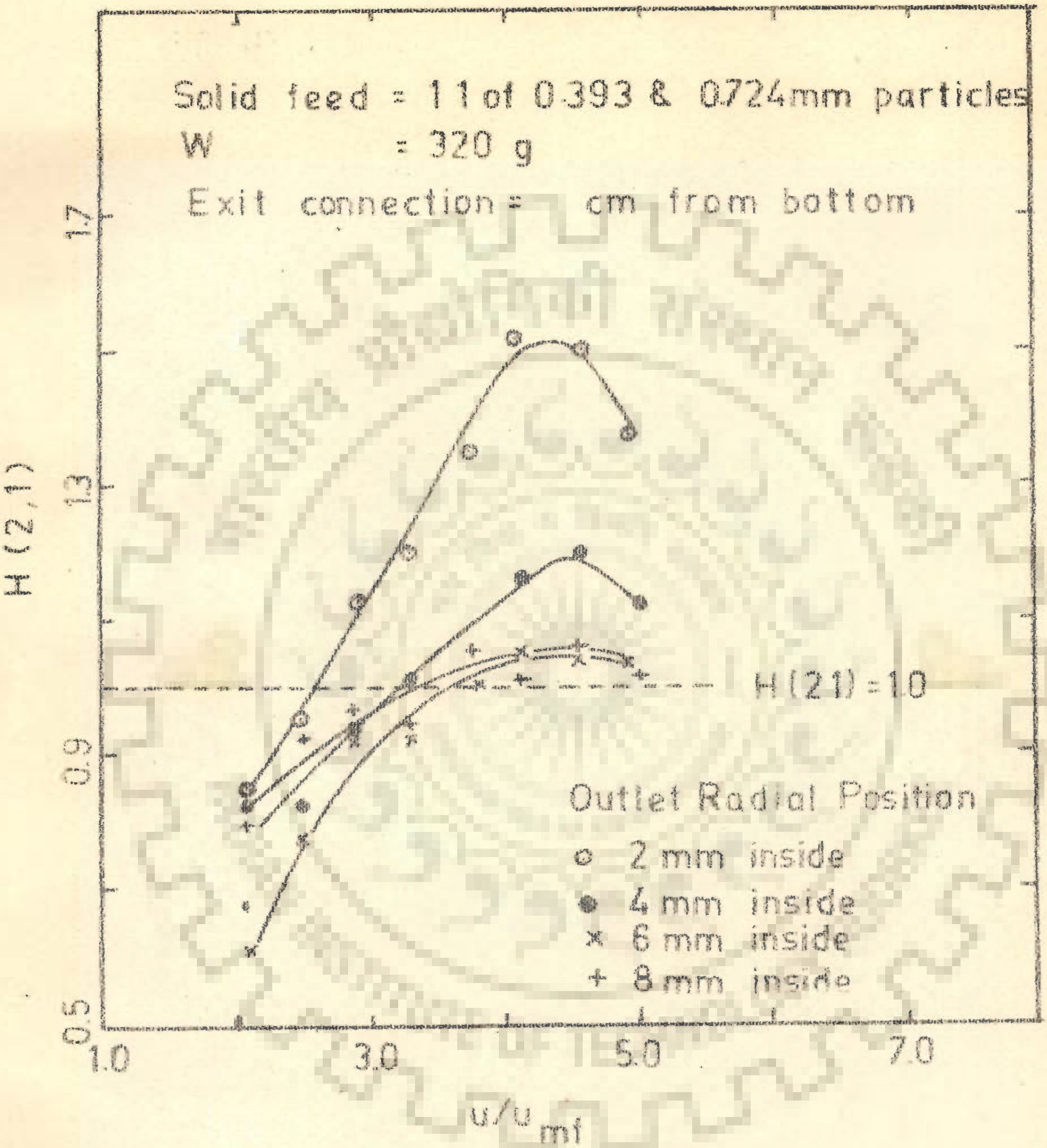


FIG.6.46 EFFECT OF RADIAL POSITION OF OPENING ON HOLD UP RATIO H(2,1)

second boundary layer with slightly greater concentration of larger particles. At higher u/u_{mf}, the second boundary layer seems to get pushed away from the surface of enclosing wall due to an increase in thickness of first layer, the hold up ratio quickly rises till u/u_{mf} value of 4.0 is reached. Beyond this value, the hold up ratio again decreases due to decrease in thickness of the first boundary layer at very high air rates as explained earlier. Similarly for radial position of solid discharge system 4 mm and 6 mm inside the bed, the curve repeats the trend obtained for the radial position 2 mm inside the bed. In general, the values of hold up ratio for radial position 4 mm inside are lower as compared to 2 mm because it appears that through the thickness of boundary layer there is a gradual change of particle concentrations, with maximum concentration of small particles closest to the enclosing wall. The curve for radial position 6 mm inside the bed can also be explained in a similar way. For radial position 8 mm inside the bed, the curve at low values of u/u_{mf} remains above the curve for radial position 6 mm inside the bed, but for u/u_{mf} above 4.0, hold up ratios for two radial positions become nearly the same. This indicates that the radial position of solid discharge system 8 mm inside the bed is beyond the second boundary layer and inside the central zone of uniform composition, so the hold up ratios continue to remain more or less near unity. This also proves that beyond second boundary layer of particles, that is, in the central zone of bed, the mixing caused by bubbles is intensive enough to prevent formation of concentration gradient in that zone. From Fig. 6.46 it can also be concluded that the values where hold up ratio is exactly 1.0 refer to the thickness of boundary layer. Thus the thickness of boundary layer is 2 mm at 2.6 u_{mf} corresponding to a bed height of almost 17 cm. Similarly the thickness of boundary layer is 4 mm and 6 mm for u/u_{mf} of 3.2 and 3.7 respectively.

Finally, to confirm the validity of the proposed model by an independent technique, runs were taken in batch fluidization to obtain the concentration profiles with the help of concentric tubes insert, shown in Fig. 4.8 and explained in Section 5.5.4. The data of Tables G.7 and G.8 is plotted in Fig. 6.47. It is clear from the figure that, in general, the concentration of smaller particles is much higher in the outer most radial zone of the bed, that is, in between 42 mm and 50 mm diameter, than the central and middle radial zones. It is also seen that the large particles have maximum concentration in the middle zone of the bed and in the central zone both the 0.393 mm particles and 0.724 mm particles are more or less in the same proportion. A precise quantitative conclusion is difficult to draw from Fig. 6.47 because the technique used to study radial distribution is not very precise. Yet the observations do give a clear indication of the existence of two boundary layers, one rich in smaller particles close to the retaining wall and the other rich in larger particles

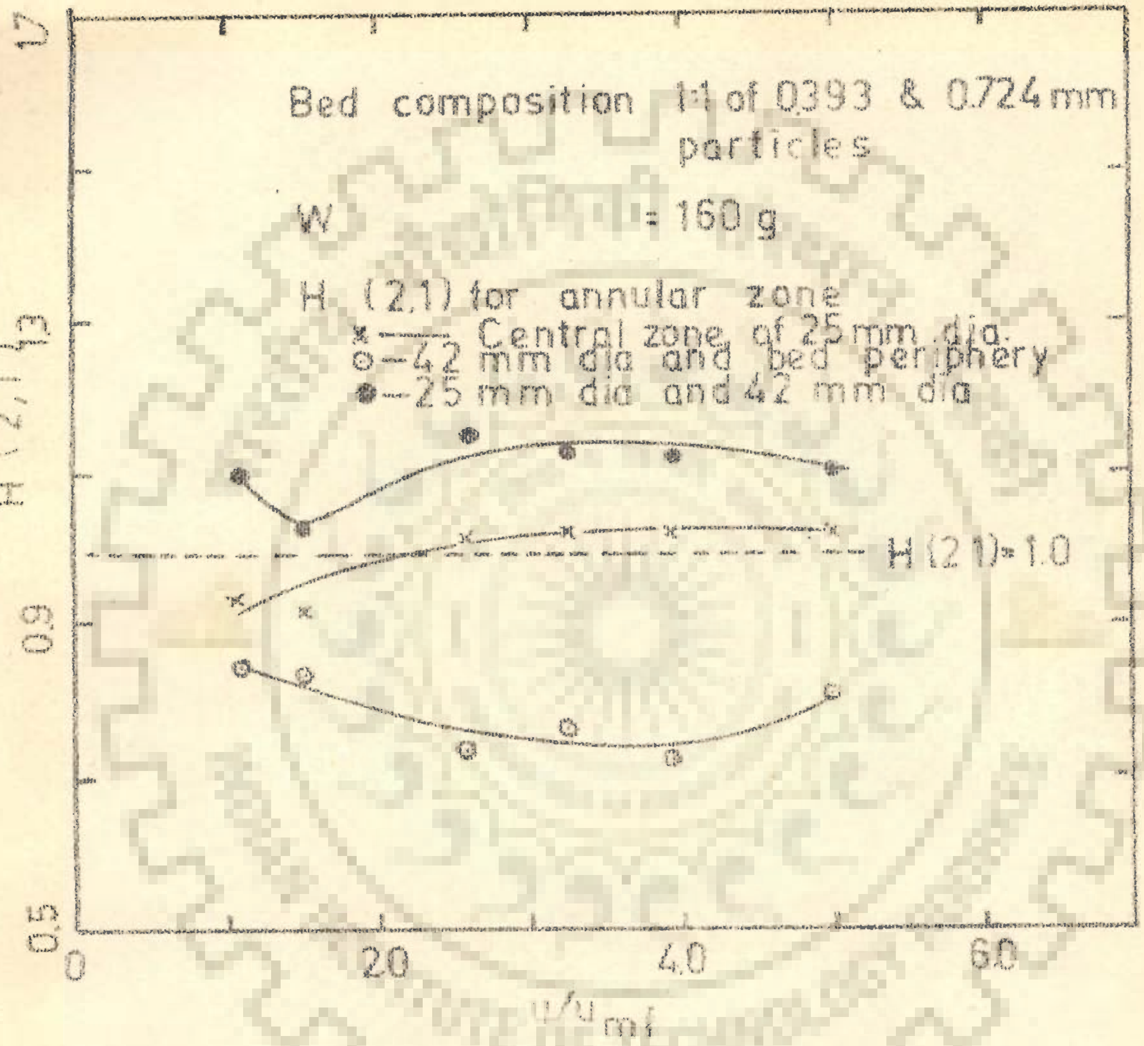


FIG. 6.47 PARTICLE COMPOSITIONS AT DIFFERENT RADIAL POSITIONS

adjacent to the first. From the above discussion on the proposed model, the important conclusions are enumerated below:

(1) Due to velocity profile of the gas in fluidized bed, the solid mixing close to the enclosing wall of the bed is different than in the central zone of the bed.

(2) In the central zone of the bed the mixing is quite intense due to the formation of bubbles at all velocities more than 2.0 u_{mf}. The mixing increases with increase in gas velocity and the bed becomes quite homogeneous in the central zone.

(3) Near the enclosing wall, due to lower gas velocities, the smaller particles tend to remain in random motion, perhaps, because of average gas velocity at this section of the bed falls to a value in between the range of minimum fluidizing velocities of the larger and small particle sizes.

(4) As a result of the postulate 3 of the model, the small particles, which tend to remain fluidized near the enclosing wall, tend to push the larger particles, which are free falling in that region of the bed, away from the surface of the enclosing wall.

(5) As a result of the postulate 4 of the model, two thin boundary layers of particles close to the enclosing wall are formed. The first layer which is situated very close to the surface of the enclosing wall has smaller particles in higher concentration. The second boundary layer which is situated adjacent to the first boundary layer is richer larger particles. However, the second boundary layer is only slightly rich in larger particles as compared to the central well mixed

zone.

(6) In the first boundary layer, the concentration increases with an increase in the value of u/u_{mf} upto 2.0 because of the additional random motion imparted to smaller particles. For u/u_{mf} values larger than 2.0, there is a tendency to decrease the thickness of the two boundary layers due to increased turbulence in central zone and also due to the possibility of some random motion of larger particles in the boundary layer close to the enclosing wall.

(7) The thickness of the boundary layers are different at different axial position of the bed. At the top surface of the bed, the thickness is zero and it grows along the length in the downward direction and may attain a stable thickness at some distance.

(8) Fluidized beds can also exhibit end effects at the top and bottom surfaces.

(9) As can be expected, slugging will adversly effect the stability of the boundary layers.

(10) The bed geometry, design of solid discharge system and the design of air distributor will have considerable influence on hold up ratio.

6.5.2 Interpretation of Experimental Data with the Proposed Model

With the proposed model, the experimental results can be easily explained. In Fig. 6.23, for static bed height corresponding to $h_D = 0.5$ it appears that the increase in the thickness of boundary layer as the bed height increases gives higher hold up ratios. The fall in the hold up ratio is not too sharp at u/u_{mf} values higher than 2.5 because the intense mixing caused by bubbles in central zone is partly off set by the tendency of the boundary layer to grow further to a stable thickness due to low fludized bed height. For a higher hold up corresponding to $h_D = 1.0$, initially higher values of hold up ratio, as compared to $h_D = 0.5$, are obtained because the boundary layer seems to have achieved a stable thickness and the sharp fall in hold up ratios after 2.5 u_{mf} is due to the reduction of boundary layer thickness due to the interaction of particles in central zone of a vigorously bubbling bed. For higher bed hold ups, $h_D = 2.0$ and 4.0, the sudden fall in the hold up ratios is due to slugging in the bed.

The effect of solid feed rates on hold up ratios for a given bed hold up as shown in Fig. 6.27 is quite systematic. With an increase in the solid feed rate, the hold up ratios fall due to exit effects at solid discharge system. At higher solid feed rates, particles are to be withdrawn at a faster rate and to meet this demand, particles will have to be withdrawn from a thicker layer near the solid discharge system resulting in lower hold up ratios. At very high solid feed rates, the layer from which the particles are withdrawn may become so thick so as to include not only the first boundary layer but also the second boundary layer and occasionally particles from the central zone also.

Fig. 6.42 shows a slight increase of hold up ratio due to shortening of the opening size of solid discharge system. This can be explained with the proposed model by considering that the smaller opening size of solid discharge system will make a smaller arc in the bed cross section, thus restricting the withdrawal of particles from a thinner layer close to the surface of the enclosing wall. This may result in slight increase in the value of hold up ratio since layer close to the wall is rich in small particles. The observed values of hold up ratio closer to unity, Fig. 6.43, for overflow and down-corner arrangement are expected from the proposed model because the solids were withdrawn from the discharge system located in the well mixed central zone.

The location of solid discharge system at peripheral position at 5 cm height from the base of the bed and its use as overflow discharge is not likely to give hold up ratios much greater than unity because the proposed model predicts top surface layer of uniform composition with no opportunity for the formation of boundary layers. Since the top layer is only slightly rich in small particles, hold up ratio values greater than unity are expected, Fig. 6.44. When the above discharge system was used to withdraw solids at a regulated rate keeping the fluidized bed height above the discharge level, increase in air flow rates increases the bed height and, therefore, the thickness of the boundary layer. Accordingly increase of hold up ratios is expected at higher u/u_{mf} values.

If the above system is used by maintaining the fluidized bed height above the discharge at a constant position, increase in air flow rates will decrease the hold up ratio because of the interaction of the particles of the central zone. This is clear in Fig. 6.45 upto bed height of 18 cm.

In their brief paper Chechetkin et. al (19) gave no details about the experimental set up and procedure, therefore, it is not possible to check applicability of the proposed model to his findings.

6.6 EFFECT OF BAFFLES ON HOLD UP RATIOS

In Section 6.6.1, a model for explaining fluidized bed behaviour is proposed and the experimental results on hold-up ratio and radial distribution of solids in the fluidized beds are explained in section 6.6.2 using that model. It is also pointed out that the sharp decrease in hold up ratios noticed at u/u_{mf} more than 2.5 is due to slugging in the bed and also due to increased turbulence of the central zone of the bed. It was, therefore, envisaged that since the horizontal and vertical internals affect the bubble size, their presence in the bed should have a significant effect on the hold up ratios in fluidized beds. The experimental results obtained in two kinds of baffles, sieve plate type of horizontal baffles of low and large spacing, and Davis type vertical baffles, shown in Fig. 4.8, are discussed in this section. The experimental data for the horizontal baffles made from screens with small openings and spaced at 13 mm given in Table H.1, Appendix H, is plotted in Fig. 6.48. It is clear from the figure that the use of such baffles reduces the hold up ratios considerably and values lower than unity are obtained for the entire range of u/u_{mf} values investigated. Severe obstruction in the movement of solid particles was observed with these baffles. The fluidization was not smooth and the segregation of particles was observed. It appears

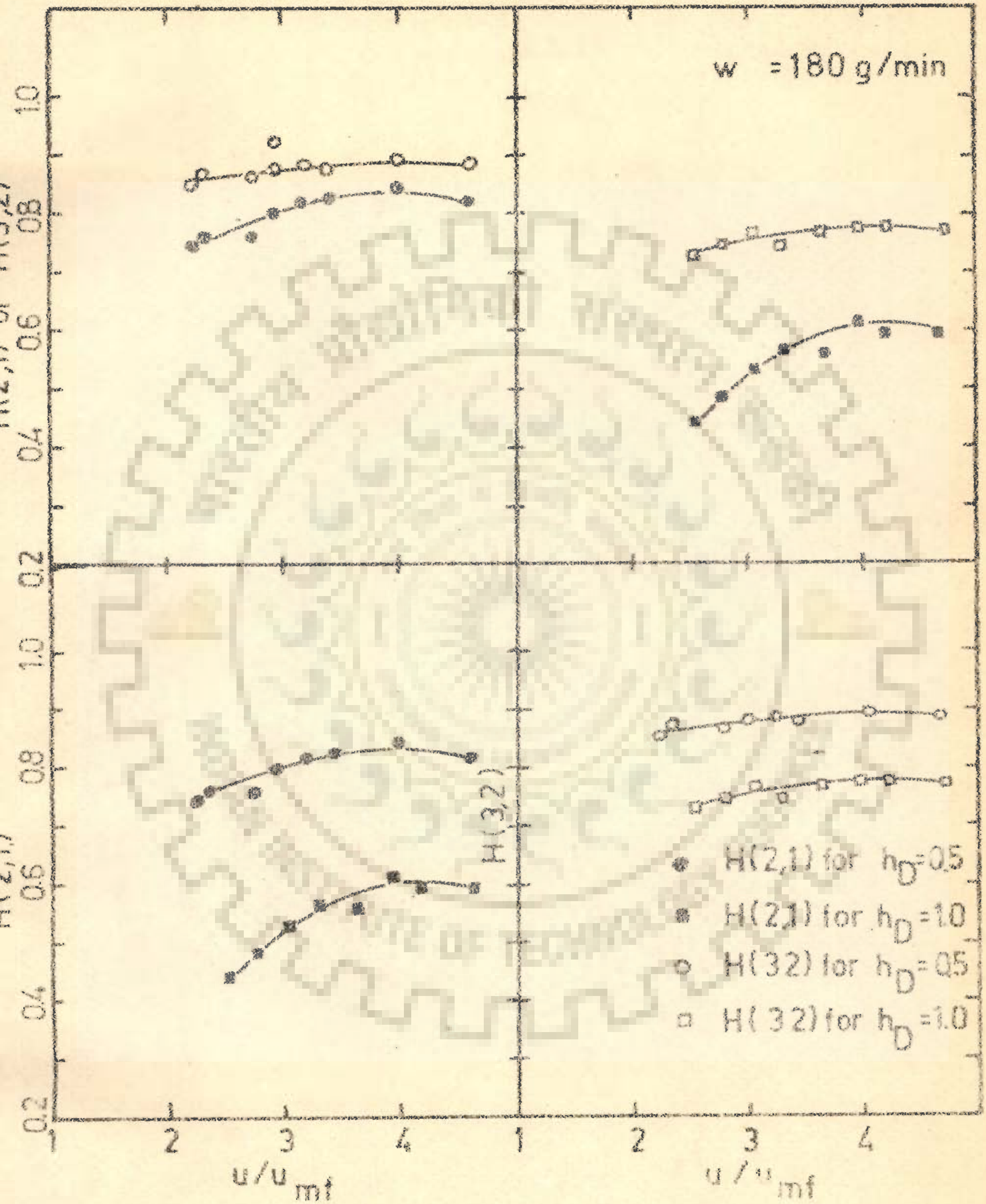


FIG.6.48 USE OF SIEVE PLATE TYPE BAFFLES WITH LOW SPACING
(Solid feed = 1:1:1 of 0.393, 0.724 & 0.96 mm particles)

that the upward movement of the particles of larger size is hindered more due to the presence of such baffles and they tend to segregate at the bottom of the bed and, therefore, enter the solid discharge preferentially over the small particles. The non-smooth fluidization also obstructs the formation of boundary layers of solids. Both these reasons are responsible for obtaining hold up ratios less than unity at the entire range of u/u_{mf} values investigated.

The experimental results for horizontal baffles made from expanded metal wire net of large size opening and spaced at 50 mm given in Table H. 2, Appendix H, are plotted in Fig. 6.49. The hold up ratios obtained in general are less than those without baffles, Fig. 6.37, but the hold up ratios greater than unity are obtained and the trend of the curves for h_D = 0.5 and 1.0 is more or less similar as observed earlier in Fig. 6.37. The fluidization was smooth and the slightly lower values of H(2,1) and H(3,2) with such baffles as compared to those without baffles are obtained as a result of the restriction on the growth of boundary layers of solids due to the presence of the baffles in the bed.

The experimental results for Davis type vertical baffles given in Table H. 3 to H. 5, Appendix H are plotted in Fig. 6.50. The vertical baffles gave smooth fluidization and are unlikely to affect the growth of the boundary layers of solid particles near the enclosing wall. However, the formation of very large bubbles and the slugging in the bed at higher air rates is likely to be reduced by Davis type baffles. Accordingly hold-up ratios greater than unity are observed and they are

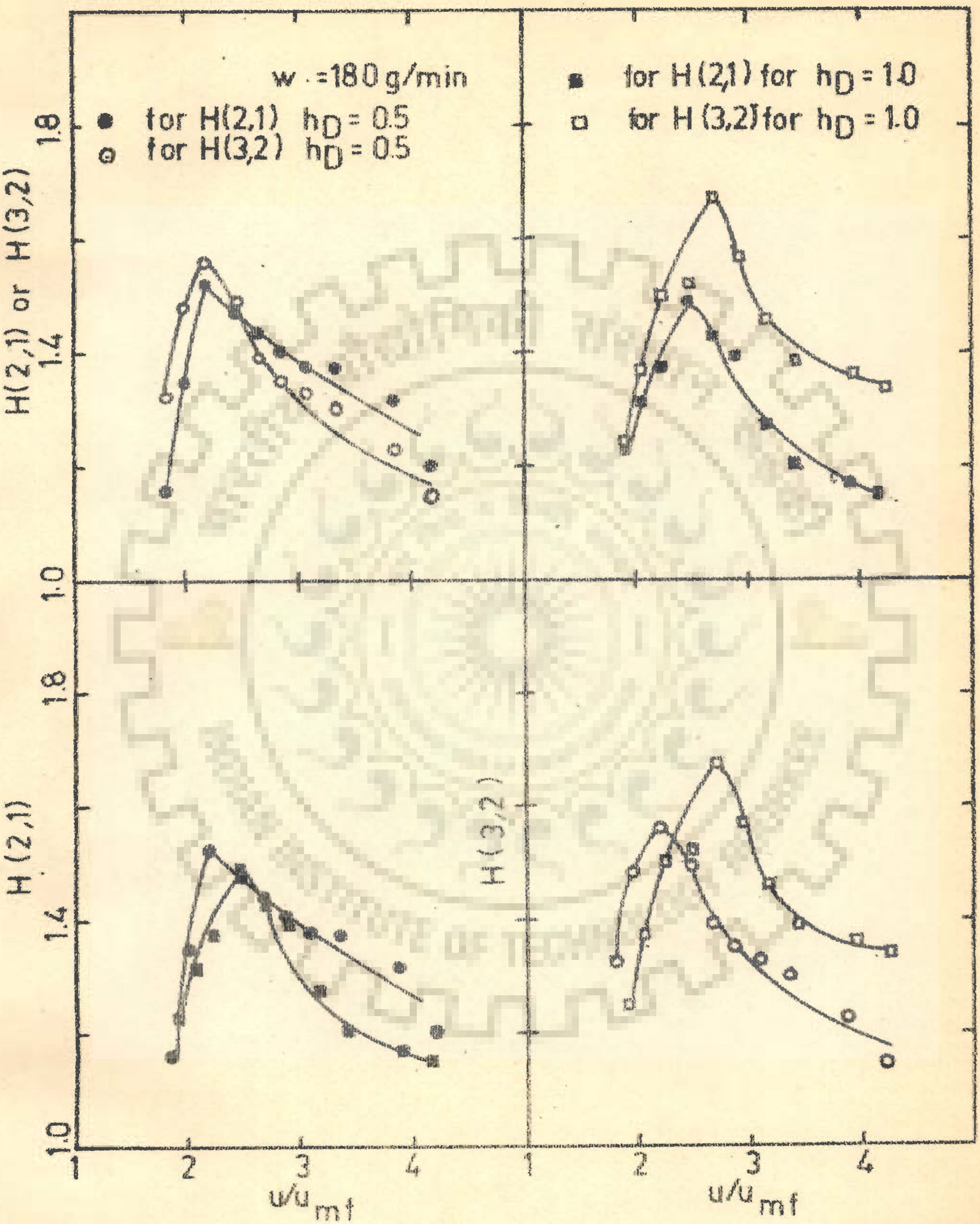


FIG.6.49 USE OF SIEVE PLATE TYPE BAFFLES WITH LARGE SPACING
(Solid feed = 1:1:1 of 0.393, 0.724 & 0.96 mm particles)

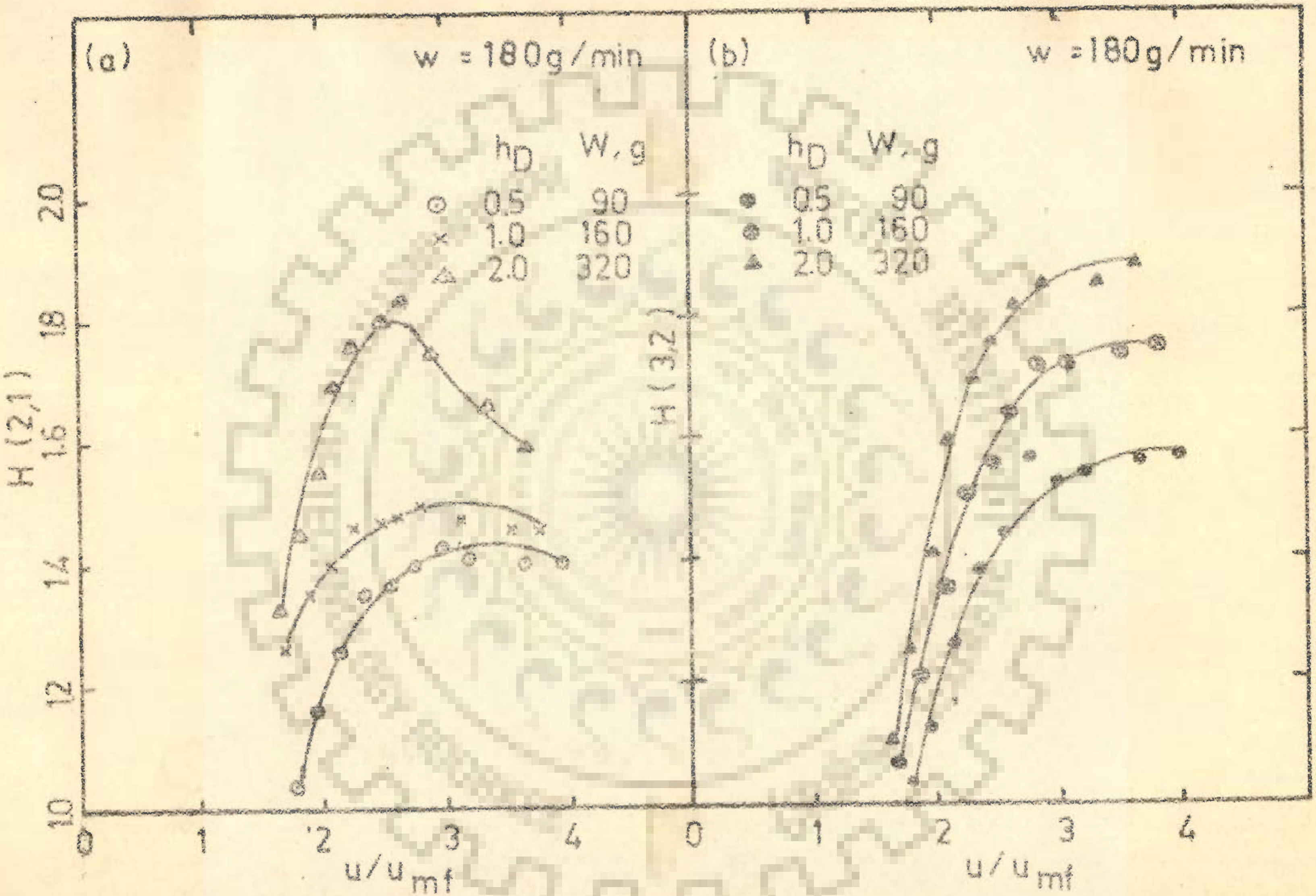


FIG.6.50 HOLD UP RATIO WITH DAVIS TYPE BAFFLES, EFFECT OF BED HOLD UP
(Solid feed = 1:1:1 of 0.393, 0.724, & 0.96 mm particles)

found to increase with increase in bed hold-up corresponding to higher values of h_D. Further, the curves do not intersect with each other, an observation reported earlier, Fig. 6.37, for fluidization studies without baffles and attributed to slug formation in the bed.

The experimental results showing the effect of solid feed rates on the hold up ratios with vertical baffles given in Tables H.6 to H.11 indicate the same trend as observed earlier in Figs. 6.27, 6.33 and 6.39. Fig. 6.51 shows the typical values of H(2,1) and H(3,2) obtained for bed hold up corresponding to $h_D = 0.5$. The parallel nature of the curves beyond the maxima of H(2,1) and H(3,2) for different solid feed rates is maintained with Davis baffles.

Fig. 6.52 shows H(2,1) and H(3,2) values as a function of u/u_{mf} for $h_D = 1.0$ and solid feed rate of 180 g/min. for runs with and without baffles for quick comparison. The important conclusion drawn from the experimental observations with different baffles are given below:

(1) Horizontal baffles with very small openings and at low spacing cause severe hinderence in the upward movement of solid particles of large size and also in the formation of boundary layers. Hence hold up ratios below unity are observed for such baffles.

(2) Horizontal baffles with large openings and at the large spacing do not obstruct fludization and results similar to those without baffles are obtained. Slight decrease in H(2, 1) and H(3,2) values with such baffles is attributed to the restr ction caused on the growth of the boundary layers due to the presence of horizontal baffles in the bed.

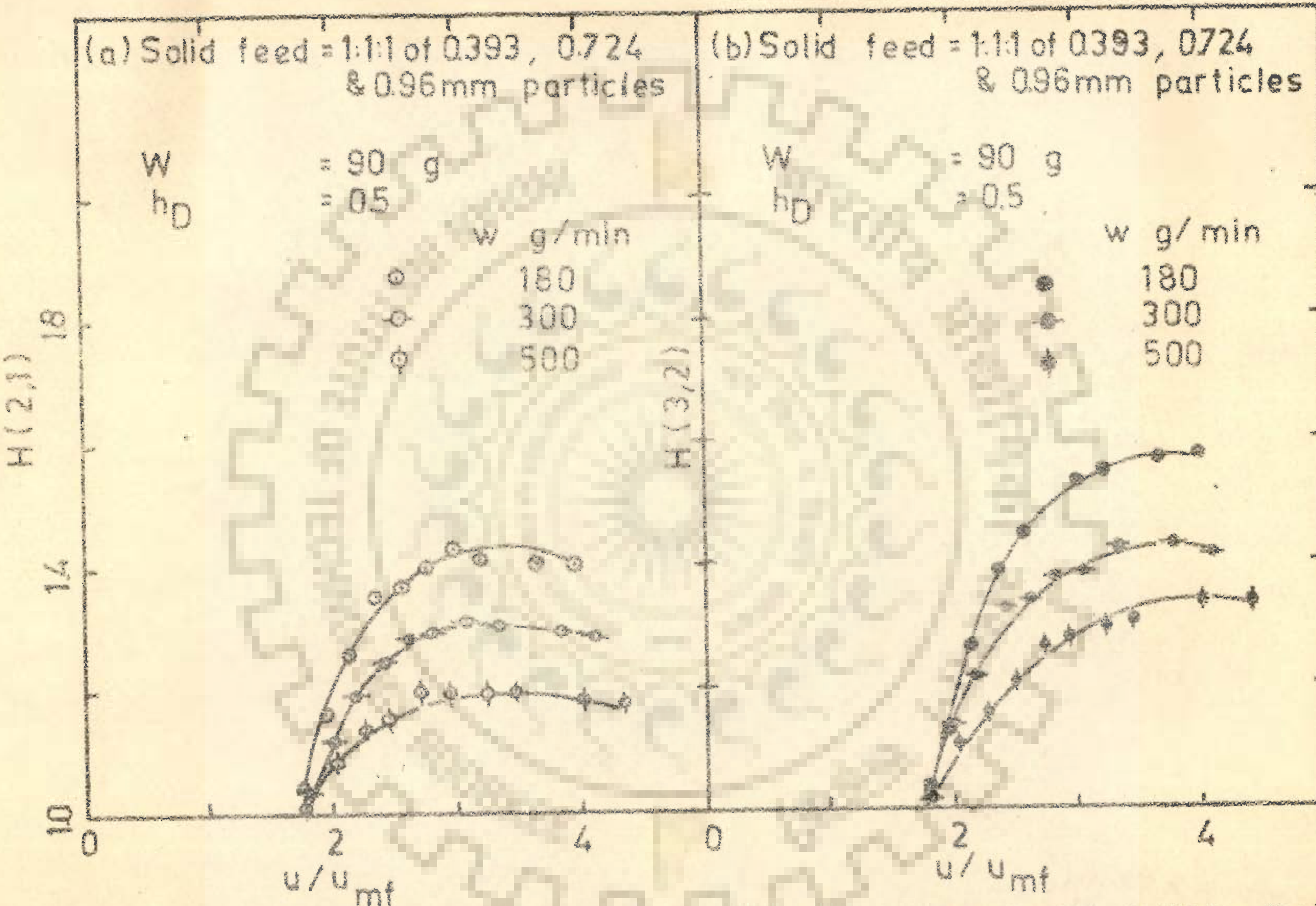


FIG.6.51 EFFECT OF SOLID FEED RATE ON HOLD RATIO WITH DAVIS TYPE BAFFLES FOR h_D = 0.5

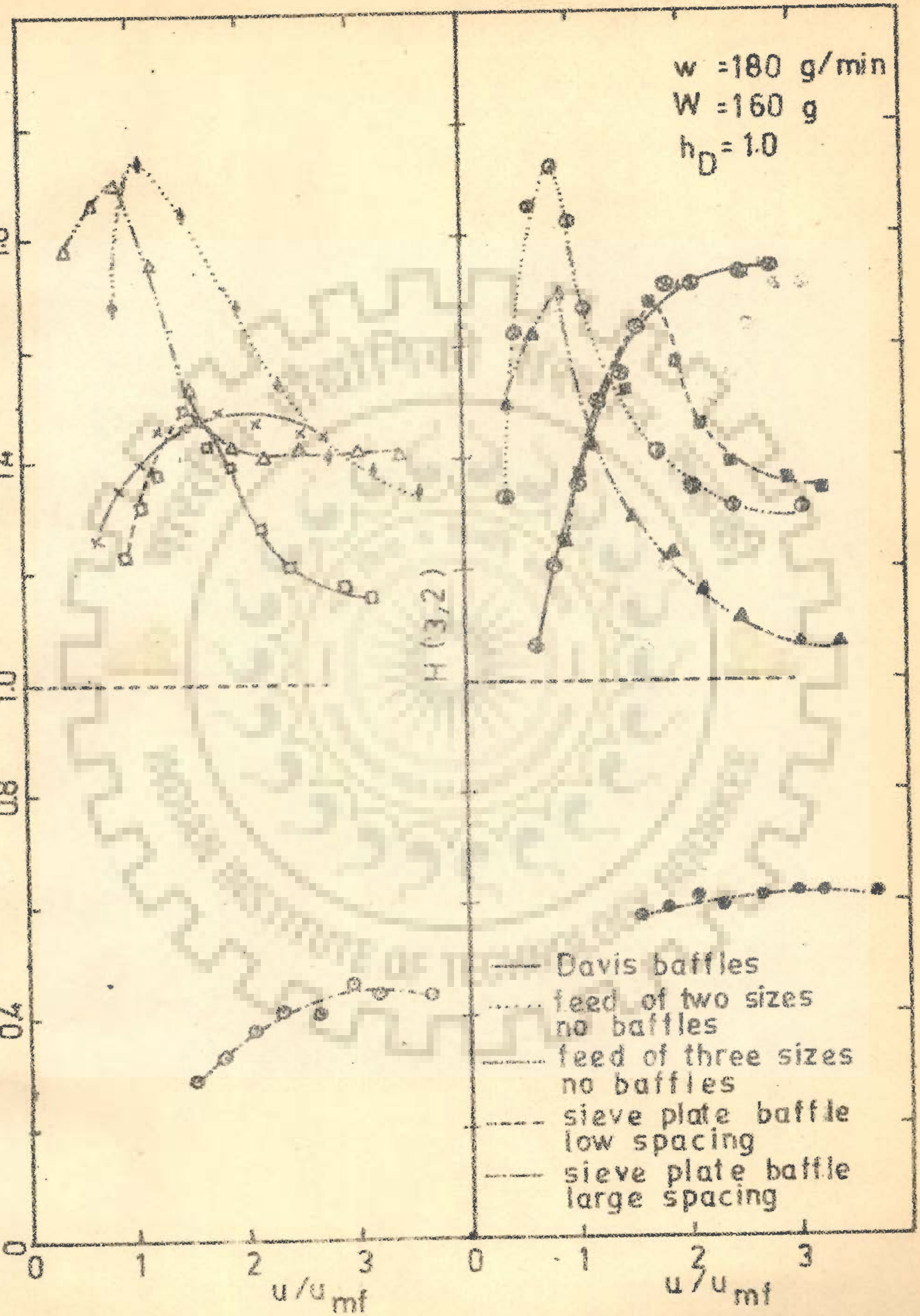


FIG. 6.52 COMPARISON OF HOLD UPS WITH AND WITHOU BAFFLES

(3) The vertical baffles of Davis type do not appear to interfere either with fluidization or with the boundary layer formation. However, the formation of very large bubbles and the slugging in the bed are minimized and sharp decrease in hold up ratios with increase in u/u_{mf} is not observed with Davis baffles.

CHAPTER 7

CONCLUSIONS AND RECOMMENDATIONS

7.1 STUDIES ON MOVING BEDS

The important conclusions drawn from the present study on moving beds with single and mixed particle sizes are summarized below:

(i) For air velocities as high as 0.95 u_{mf} the flow behaviour of solids in the moving beds quite closely resembles the plug flow for single and mixed particle sizes.

(ii) The tracer experiments on moving beds are likely to be influenced by the end effects depending on the design of outlet for solid discharge.

(iii) There is practically no effect either of bed height or the mean residence time or the particle size on the flow pattern of solids in the moving beds upto u/u_{mf} values of 0.95.

(iv) In the experimental range investigated, a dead zone of particles exists in the moving beds, and this dead zone is not much influenced by variables like particle size and bed height but increase in the solid feed rates tends to decrease the dead zone.

(v) The particles of all sizes in a mixed feed travel through the bed in a plug flow at all velocities upto 0.95 u_{mf} and this includes superficial air velocities more than the minimum fluidizing velocity of the small size particles.

7.2 STIMULUS RESPONSE STUDIES ON FLUIDIZED BEDS

The flow behaviour of single and mixed particle sizes through a fluidized bed leads to the following conclusions:

(i) Solid flow pattern resembles ideal backmix flow for air velocity of greater than 1.5 u_{mf}. For air velocity values 1.2 u_{mf} to 1.5 u_{mf}, the flow behaviour is non ideal and exhibits the presence of some short-circuiting, dead zone and plug flow along with the main back mix flow. The results are non-reproducible for air flow rates less than 1.2 u_{mf}.

(ii) In a fluidized bed of mixed sizes with continuous flow of solids, particles of each size experience a different mean residence time even though the flow pattern for solids remains ideal backmix flow for air velocity greater than 1.5 u_{mf}.

(iii) The mean residence time of large size particles is found to be more than the mean residence time of small size particles.

(iv) The bed hold up ratio values for particles of different sizes in a fluidized bed at steady state are as accurate as the stimulus response studies to predict the mean residence time for each size particles.

7.3 STUDIES ON HOLD UP RATIOS IN FLUIDIZED BEDS

(i) The hold up ratios increase with increase of u/u_{mf} upto about 2.5. With a further increase in air flow rates, the hold up ratios exhibit sharp falls.

(ii) There is practically no effect of feed composition on hold up ratio provided the dimensionless form of u/u_{mf} is used to represent flow rates.

(iii) The hold up ratios H(3, 2) and H(2, 1) for mixed feeds of three sizes are lower than those obtained with feeds of two particle sizes possibly due to mutual interaction between particles.

(iv) The effect of solid feed rate on hold up ratio can be represented by a second order polynomial of the following form, for $u/u_{mf} > 2.5$

$$H / (\tau_B)^n = b + c\ (u/u_{mf}) + d\ (u/u_{mf})^2$$

where n, b, c and d are constants. The value of n is obtained by graphical procedure and regression analysis is used to determine the ratio for any two particle sizes, namely, H(2, 1) for middle (0.724 mm) to small (0.393 mm) and H(3, 2) for large (0.96 mm) to middle (0.724 mm) size particles. τ_B is a dimensionless parameter defined as the ratio of volumetric gas flow rate to volumetric solid flow rate, to account for solid feed rate.

Using regression analysis on IBM 1620 computer the values of constants for different particle size combination and static bed heights corresponding to height to diameter ratio h_D are tabulated below:

Particle sizes in feed mm	Hold up ratio	h_D	n	b	c	d	Maximum error %
0.724 & 0.393	H(2,1)	0.5	0.3	0.223	-0.00289	-0.00162	±10
0.724 & 0.393	H(2,1)	1.0	0.3	0.395	-0.0947	0.00832	±10
0.724 & 0.393	H(2,1)	2.0	0.3	0.285	0.0484	0.00359	±10
0.960 & 0.724	H(3,2)	0.5	0.2	0.329	0.0213	-0.00702	± 3
0.960 & 0.724	H(3,2)	1.0	0.2	0.619	-0.186	0.0235	± 4
0.393, 0.724 & 0.96	H(2,1)	0.5	0.2	0.774	-0.225	0.0248	± 5
0.393, 0.724 & 0.96	H(3,2)	0.5	0.2	0.512	-0.709	0.00595	± 3
0.393, 0.724 & 0.96	H(2,1)	1.0	0.2	0.482	0.0999	0.01	± 10

7.4 STUDIES ON RADIAL AND AXIAL DISTRIBUTION OF SOLIDS ANL BASIC FEATURES OF THE PROPOSED MODEL

From the axial and radial distribution of particles, measured by changing the axial and radial position of solid discharge system and by concentric tube insert, proposed to explain the solid flow pattern in fluidized beds containing different size particles, exhibits the following main features:

(i) Due to velocity profile of the gas in fluidized bed, the solid mixing close to the enclosing wall of the bed is different than in the central z one of the bed.

(ii) In the central zone of the bed, the mixing is quite intense due to the formation of bubbles at velocities more than 2.0 u_{mf}. The mixing increases with increase in air rate and the bed becomes quite uniform in the central zone.

(iii) Near the enclosing wall, due to lower gas velocities, the smaller particles tend to remain in random motion, perhaps, because the average gas velocity at this section of the bed falls to a value in between the range of minimum fluidizing velocities of the large and small particle size.

(iv) As a result of postulate 3 of the model, the small particles, which tend to remain fluidized near the enclosing wall, tend to push the large particles, which are free falling in that region of the bed, away from the surface of the enclosing wall.

(v) As a result of the postulate 4 of the model, two thin boundary layers of particles close to the enclosing wall are formed. The first layer which is situated very close to the surface of the enclosing wall has smaller particles in higher concentration. The second boundary layer is slightly rich in larger particles as compared to the central well mixed zone.

(vi) In the first boundary layer, the concentration of small particles increases with an increase in the value of u/u_{mf} upto 2.0 because of the additional random motion imparted to the smaller particles. For u/u_{mf} values larger than 2.0, there is a tendency to decrease the thickness of the two boundary layers due to increased turbulence in central zone and also due to the possibility of some random motion of larger particles in the boundary layer close to the enclosing wall.

(vii) The thickness of the boundary layers are different at different axial position of the bed. At the top surface of the bed, the thickness is zero and it grows along the length in the downward direction and may attain a stable thickness at some distance.

(viii) Fludized beds can also exhibit end effects at the top and bottom surfaces.

(ix) As expected, the slugging adversely effects the stability of the boundary layers.

(x) The bed geometry, design of solid discharge system and the design of air distributor is expected to have considerable influence on hold up ratio.

(xi) The proposed model satisfactorily explains all the experimental observations on hold up ratios and radial and axial distributions of solid particles of different sizes in a fluidized bed.

7.5 STUDIES WITH HORIZONTAL AND VERTICAL BAFFLES

(i) The sieve plate type baffles with small openings and low spacing between the baffles tend to segregate large particles at the bottom of the bed and as such hold up ratios less than one are obtained for solid discharge system fixed at the bottom of the bed.

(ii) The sieve plate type of baffles with large openings and large spacing between the baffles tend to disturb the boundary layers of particles at the enclosing wall of the fluidized bed and some reduction in hold up ratio is observed.

(iii) The Davis type vertical baffles with large openings do not interfere, in general with the boundary layers of the particles in the fluidized bed. However, they prevent the sharp decrease in the hold up ratios for air flow rates larger than 2.5 u_{mf} even for high bed hold ups due to decrease in the slugging tendency of the bed.

7.6 RECOMMENDATIONS

1) The effect of particle size ratios in mixed feeds of two and more particle sizes on hold up ratios should be investigated further for better understanding of the mutual interaction of particles of different sizes in fluidized bed so as to arrive at a more unified correlation for hold up ratios.

2) The effect of physical characteristics of solids such as density, sphericity and porosity on hold up ratios should be investigated because industrial feeds rarely contain spherical particles and particle density generally changes as a result of chemical reaction.

3) Study should be extended to larger bed diameters for scale up purposes.

4) The effect of the position and design of solid discharge system needs further investigation for better design of fluidized bed reactors.

APPENDIX - A

STANDARDIZATION OF EXPERIMENTAL SET-UP

Table A-1

EXPERIMENTAL DETERMINATION OF MINIMUM FLUIDIZING VELOCITY FOR 0.393 mm GLASS BALLOTINI BEADS

Barometric Pressure = 725 mm Hg

Temperature = 21°C, Solid Feed = Glass ballotini beads

Particle Size = 0.393 mm   Pressure at the outlet of Rotameters = 1.5 kgf/cm^2

3 Sets of runs were taken by keeping 3 bed heights.

S. No.	Rotameter type and reading	Air Flow rate LPM (G)	Pressure drop for h_D (in mm water)		
			2	4	6
1.	500$^{(R1)}$	6.15	-	86	115
2.	600	7.35	46	92	135
3.	750	9.25	60	122	180
4.	1000	12.40	75	158	227
5.	1250	15.30	95	202	288
6.	1500	18.30	115	233	338
7.	1600	19.30	122	268	412
8.	1750	20.50	130	275	430
9.	2000	23.40	132	279	430
10.	2250	25.8	130	277	428
11.	2500	28.0	130	275	425

p. t. o.

12.	2750	31.0	130	275	425
13.	3000	34.1	130	275	425
14.	3250	36.0	131	275	425
15.	3500	39.2	130	275	425
16.	1.00(R2)	52.2	130	276	427

Table A-2

EXPERIMENTAL DETERMINATION OF MINIMUM FLUIDIZING VELOCITY FOR 0.724 mm GLASS BALLOTINI BEADS

Barometric Pressure = 725 mm Hg

Temperature = 21°C Solid Feed Glass ballotini

Particle size = 0.724 mm Outlet pr. of rotameters = 1.5 kgf/cm^2

Runs taken at bed-height to bed diameter ratio of (i) 2, (ii) 4(iii)6.

S. No.	Rota.Reading & type	Flow rate LPM	2	4	6
1.	600(R1)	7.35	17	35	50
2.	1000	12.4	30	58	82
3.	0.25(R2)	15.38	37	70	105
4.	0.50	27.2	55	117	180
5.	0.75	37.8	80	162	251
6.	1.00	52.2	110	222	355
7.	1.125	60.0	127	270	418
8.	1.175	63.0	134	280	430
9.	1.25	67.5	130	279	426
10.	1.50	80.2	130	275	425
11.	1.75	94.5	130	275	426
12.	2.25	122.0	132	275	425

Table A-3

EXPERIMENTAL DETERMINATION OF MINIMUM FLUIDIZING VELOCITY FOR 0.960 mm dia GLASS BALLOTINI BEADS

Barometric Pressure = 725 mm Hg

Temperature = 21°C   Solid Feed   Glass Ballotini

Particle size = 0.96 mm,   outlet pr. of rotameter = 1.5 kgf' cm^2

S.No.	Rotameter reading	Flow rate LPM	ΔP for h_D (in mm H_2O) 2	4	6
1.	0.25(R2)	15.38	20	19	60
2.	0.50	27.2	35	74	108
3.	0.75	37.8	49	102	148
4.	1.00	52.2	86	135	202
5.	1.25	67.5	103	173	274
6.	1.50	80.2	122	209	331
7.	1.75	94.5	128	251	393
8.	1.90	103	130	279	431
9.	2.00	107.3	130	275	425
10.	2.25	122.0	130	275	424
11.	2.50	134.9	130	275	425
12.	3.00	161.2	130	2.75	425

Table A-4

EXPERIMENTAL DETERMINATION OF MINIMUM FLUIDIZING VELOCITY FOR MIXED FEEDS OF 0.393 mm and 0.724 mm SIZES

Barometric pressure = 725 mm Hg

Temperature = 21°C

Pressure at the outlet of Rotameters = 1.5 kgf/cm^2

Static bed Ht. to Diameter ratio = 4.0

S. No.	Rotameter reading	Air Flow rate(LPM) G	Pressure Drop for 0.393 & 0.724 mm size ratios of 1: 0.25	1:1	1:4
1.	500(R1)	6.15	81	58	42
2.	600	7.35	95	-	-
3.	750	9.25	105	-	-
4.	1000	12.40	-	115	-
5.	1250	15.30	175	130	88
6.	1500	18.30	192	148	-
7.	1600	19.30	210	-	-
8.	1750	20.50	225	-	-
9.	2000	23.40	254	-	-
10.	2250	25.8	274	-	-
11.	2500	28.0	275	225	142
12.	2750	31.0	275	274	-
13.	3000	34.1	275	275	-
14.	0.75	37.8	-	275	200

15.	1.00	52.2	-	275	276
16.	1.125	60.0	-	-	275
17.	1.175	63.0	-	-	276
18.	1.25	67.5	-	-	275
19.	1.25	80.2	-	-	275

Table A-5

DETERMINATION OF MINIMUM FLUIDIZING VELOCITY FOR MIXED FEEDS OF 0.724 mm and 0.96 mm SIZES

Barometric Pressure = 725 mm Hg

Temperature = 21°C

Pressure at the outlet of rotameter = 1.5 kgf/cm^2

Static Bed ht. to dia. ratio = 40

S. No.	Rotameter reading	Air flow rate LPM	ΔP for 0.724 mm and 0.96 mm ratio of:		
			1:0.25	1:1	1:4
1.	0.25(R2)	15.38	67.5	56	46
2.	0.50	27.2	105	97	78
3.	0.75	37.8	145	132	109
4.	1.00	52.2	198	183	148
5.	1.25	67.5	273	245	197
6.	1.30	70.0	277	-	-
7.	1.40	74.5	-	277	-
8.	1.50	80.2	275	275	235
9.	1.70	91.5	-	-	276
10.	1.75	94.5	276	27.6	275
11.	2.00	104.0	275	275	275
12.	2.00	108.0	275	-	276
13.	2.25	122	-	-	275
14.	2.50	135	-	-	275
15.	2.75	148	-	-	275

A. 1 CALCULATION OF MINIMUM FLUIDIZING VELOCITY FOR A TYPICAL RUN

Bed diameter $D = 5.08$ cm

Particle size $dp = 0.393$ cm

Bed Hold-up $(W) = 90.0$ gm

Static Bed Height $(h) = 3.0$ cm.

Bed volume $(V_b) = \frac{2.5}{4} (5.08)^2 = 50.6 \text{ cm}^3$

Solid density $(\rho s) = 2.488$

Solid volume $(Vs) = 90/2.488 = 36.1 \text{ cm}^3$

Bed Porosity$(\epsilon_{mf}) = \frac{60.8 - 36.1}{60.8} = 0.406$

Air density at S. T. P. $(\rho_g) = 1.29 \times 10^{-3}$ g/c.c.

Air viscosity at S. T. P. $= 0.018$ cp.

$U_{mf} =$

$$\frac{(0.0393)^2}{150} \quad \frac{(2.488 - .0012)}{0.018 \times 10^{-2}} \; 981 \frac{(.406)^3}{0.594}$$

$= 16.3$ cm/sec.

Volumetric air flow rate $= 16.3 \quad \frac{(5.08)^2}{4} \quad \frac{60}{1000}$; L. P. M.

$= 19.8$ L. P. M.

APPENDIX B

MOVING BEDS WITH SINGLE FEED SIZES

Table B-1

STIMULUS RESPONSE IN MOVING BED FOR SINGLE SIZE FEEDS(REPLICATE RUN)

Particle size = 0.724 mm   Tracer Amount = 10g
w =100.0g/min u/u_{mf} = 0.6
Bed heights = 10 cm for tracer input data
= 30 cm for tracer output data

S. No.	Tracer Inpur Data t seconds	C	Tracer Output Data t Seconds	C
1.	75.	0.62E-02	453	0.21
2.	90	0.19E-01	450	0.27
3.	100	0.18	470	0.30
4.	110	0.20	480	0.32
5.	120	0.25+01	490	0.26
6.	130	0.12E+02	500	0.98E+01
7.	140	0.11E+02	510	0.15E+02
8.	150	0.85E+01	520	0.99E+01
9.	160	0.70E+01	530	0.90E+01
10.	170	0.47E+01	540	0.64E+01
11.	180	0.36E+01	550	0.43E+01
12.	240	0.53	560	0.27E+01
13.	300	0.27	570	0.21E+01
14.	390	0.20	600	0.77
15.	540	0.14	630	0.39
16.	660	0.70E-01	660	0.43
17.			780	0.30
18.			960	0.22

Table B-2
1 (REPLICATE RUN)

STIMULUS RESPONSE IN MOVING BED FOR SINGLE SIZE FEEDS

Particle size = 0.724 mm w = 100g/min.
u/u_{mf} = 0.6
Bed heights = 10 cm for tracer input data
30 cm for tracer output data

S.	Tracer Inpur Data		Tracer Output Data	
No.	t	C	t	C
1.	60	0.55	390	0.70E-02
2.	70	0.25	420	0.38E-01
3.	80	0.19	450	0.61E-02
4.	90	0.19	480	0.19
5.	100	0.28E-01	490	0.43E+01
6.	110	0.28	500	0.14E+02
7.	120	0.52E+01	510	0.12E+02
8.	130	0.10E+02	520	0.92E+01
9.	140	0.98E+01	530	0.57E+01
10.	150	0.85E+01	540	0.54E+01
11.	160	0.70E+01	550	0.39E+01
12.	170	0.54E+01	560	0.24E+01
13.	180	0.39E+01	570	0.17E+01
14.	210	0.13E+01	600	0.63
15.	240	0.60	660	0.25
16.	360	0.12	840	0.54E-01
17.	600	0.32	1200	0.31E-01

Table B-3

(REPLICATE RUN)

STIMULUS RESPONSE IN MOVING BED FOR SINGLE SIZE FEEDS

Particle Size = 0.724 mm w = 100 g/min.
u/u_{mf} = 0.6
Bed height = 10 cm for tracer input data
30 cm for tracer output data

S.	Tracer Input Data		Tracer Output Data	
No.	t sec	C	t sec	C
1.	60	0.32	480	0.19E+01
2.	70	0.44	490	0.12E+02
3.	80	0.25	500	0.13E+02
4.	90	0.10	510	0.82E+01
5.	100	0.36	520	0.61E+01
6.	110	0.62E+01	530	0.49E+01
7.	120	0.11+02	540	0.42E+01
8.	130	0.94E+01	550	0.31E+01
9.	140	0.77E+01	560	0.24E+01
10.	150	0.74E+01	570	0.16E+01
11.	160	0.50E+01	580	0.12E+01
12.	170	0.39E+01	590	0.11E+01
13.	180	0.28E+01	600	0.13E+ 01
14.	210	0.77	630	0.40
15.	240	0.47	660	0.66
16.	360	0.24	780	0.48
17.	600	0.30		

Table B-4

STIMULUS RESPONSE IN MOVING BED FOR SINGLE SIZE FEEDS

Particle size = 0.393 | Tracer Amount = 5 g.
w = 34 g/min. | u/umf = 0.2
Bed heights = 10 cm for tracer input data
= 20 cm for tracer output data

S. No.	Tracer Input Data: t seconds	Tracer Input Data: C	Tracer Output Data: t seconds	Tracer Output Data: C
1.	330	0.00	900	0.50E-01
2.	340	0.54E-01	930	0.24
3.	350	0.31	950	0.96
4.	360	0.54E-01	960	0.20E+01
5.	370	0.80E-01	970	0.50E+01
6.	380	0.82E-01	980	0.13E+02
7.	390	0.83	990	0.82E+01
8.	400	0.13E+01	1000	0.84E+01
9.	410	0.62E+01	1010	0.80E+01
10.	420	0.36E+01	1020	0.78E+01
11.	430	0.81E+01	1050	0.50E+01
12.	440	0.12E+02	1080	0.34E+01
13.	450	0.61E+01	1120	0.36E+01
14.	460	0.69E+01	1180	0.28E+01
15.	480	0.55E+01	1260	0.76
16.	540	0.41E+01	1380	0.19
17.	660	0.25E+01		

Table B-5

STIMULUS RESPONSE IN MOVING BED FOR SINGLE SIZE FEEDS

Particle size = 0.96 mm w = 300 g/min
Tracer amount = 10g. u/umf = 0.2
Bed heights = 10 cm for tracer input data
= 30 cm for tracer output data

S.	Tracer Input Data		Tracer Output Data	
No.	t seconds	C	t seconds	C
1.	30	0.16	120	0.17
2.	40	0.10E	1 [illegible]	0.17E+02
3.	50	0.11E+02	140	0.15E+02
4.	60	0.14E+02	150	0.94E+02
5.	70	0.76E+02	160	0.42E+02
6.	80	0.38E+02	170	0.17E+01
7.	90	0.22E+02	180	0.10E+01
8.	120	0.48	190	0.86
9.	150	0.13	200	0.10E+01
=0.	180	0.16	210	0.58
11.	-	-	240	0.32
12.	-	-	300	0.35E-01
13.	-	-	360	0.60E-01
14.	-	-	420	0.00

Table B-6

STIMULUS RESPONSE IN MOVING BED FOR SINGLE SIZE FEEDS

Particle size = 0.393 mm
w = 300. g/min
Bed heights = 10 cm for tracer input data
= 40 cm for tracer output data

Tracer amount = 5g.
u/υ = 0.2

S. No.	Tracer Input Data t seconds	C	Tracer Output Data t seconds	C
1.	30	0.44	330	0.40
2.	40	0.44	340	0.28
3.	50	0.25	350	0.43
4.	60	0.56	360	0.71
5.	70	0.12E+02	370	0.15E+02
6.	80	0.12E+02	380	0.12E+02
7.	90	0.11E+02	390	0.85E+01
8.	100	0.15E+01	400	0.67E+01
9.	110	0.12E+01	410	0.40E+01
10.	120	0.10E+01	420	0.14E+01
11.	150	0.44	450	0.52
12.	180	0.35	480	0.68
13.	240	0.44	540	0.27

Table B-7

STIMULUS RESPONSE IN MOVING BED FOR SINGLE SIZE FEEDS

Particle Size = 0.393 mm
w = 300 g/min
Bed height = 10 cm for tracer input data
= 20 cm for tracer output data

Tracer amount 5 g
$u/u_{mf} = 0.95\ u_{mf}$

S. No.	Tracer Input Data t seconds	C	Tracer Output Data t seconds	C
1.	30	0.28	120	0.23
2.	40	0.30	130	0.25
3.	50	0.26	140	0.30
4.	60	0.11E+02	150	0.24
5.	70	0.76E+01	160	0.40
6.	80	0.26E+01	170	0.13E+02
7.	90	0.74	180	0.57E+01
8.	120	0.40	190	0.40E+01
9.	150	0.28	200	0.49
10.	180	0.22	210	0.48
11.			220	0.33
12.			230	0.33
13.			240	0.30
14.			270	0.19
15.			300	0.20

Table B-8

STIMULUS RESPONSE IN MOVING BED FOR SINGLE SIZE FEEDS

particle size = 0.393 mm Tracer Amount 5 g

w = 34. g/min u/u_{mf} = 0.95

Bed heights = 10 cm for tracer input data

= 40 cm for tracer output data

S. No.	Tracer Input Data t seconds	C	Tracer Output Data t seconds	C
1.	270	0.46	1800	0.20
2.	300	0.12E+01	1950	0.82
3.	330	0.14	1965	0.34E+01
4.	360	0.60	1980	0.74E+01
5.	380	0.62	2000	0.10E+02
6.	390	0.30E+01	2010	0.12E+02
7.	400	0.79E+01	2020	0.11E+02
8.	410	0.10E+02	2030	0.10E+02
9.	420	0.11E+02	2040	0.84E+01
10.	430	0.12E+02	2050	0.76E+01
11.	440	0.10E+02	2060	0.70E+01
12.	450	0.86E+01	2070	0.48E+01
13.	480	0.58E+01	2130	0.37E+01
14.	510	0.46E+01	2160	0.280E+01
15.	570	0.26+01	2220	0.48E+01
16.	690	0.62E+01	2400	0.40
17.	840	0.46		

Table B-9

STIMULUS RESPONSE IN MOVING BED FOR SINGLE SIZE FEEDS

Particle size = 0.96 mm
$\dot{w}$ = 300 g/min
Bed heights = 10 cm for tracer input data
= 40 cm for tracer output data

Tracer amount 10 g
u/u_{mf} = 0.95

S. No.	Tracer Input Data t Seconds	C	Tracer Output Data t Seconds	C
1.	40	0.75E-01	240	0.28E-01
2.	50	0.85E+01	300	0.46
3.	60	0.60E+01	310	0.21
4.	70	0.38E+01	320	0.20E-01
5.	80	0.11E+01	330	0.13
6.	90	0.87	340	0.12
7.	100	0.24	350	0.11
8.	110	0.36	360	0.15
9.	120	0.20	370	0.62E+01
10.	150	0.20	380	0.82E+01
11.	180	0.11	390	0.61E+01
12.	240	0.11	400	0.35E+01
14.			420	0.40E+01
15.			450	0.24E+01
16.			480	0.20E+01
17.			540	0.19E+01

APPENDIX - C

MOVING BEDS WITH MIXED FEED SIZES

Table C-1

MOVING BED WITH MIXED FEED OF 0.393 mm AND 0.724 mm PARTICLES

Solid Feed = 1:1 ratio of 0.393 mm & 0.724 sizes
Solid Feed rate = 90 g/min of each size
u/u_{mf} = 0.95 (u_{mf} for average diameter)
Tracer Amount = 10 g.

S. No.	Time Seconds	C for tracer size: 0.393 mm only	0.724 mm only	1:1 of 0.393 mm and 0.724 mm
(a) For bed hold up 340 g (Input signal curve)				
1.	60	0.15	0.10	0.10
2.	70	0.43	0.10	0.10
3.	80	0.61 E+01	0.22E+01	0.55E+01
4.	90	0.12E+02	0.13E+02	0.12E+02
5.	100	0.1E+ 02	0.10E+02	0.12E+02
6.	110	0.10E+02	0.83E+01	0.10E+02
7.	120	0.11E+02	0.80E+01	0.94E+01
8.	150	0.75E+01	0.42E+01	0.80E+01
9.	180	0.34E+01	0.64	0.25E+01
10.	240	0.95	0.23	0.72
11.	300	0.21	0.10	0.44
(b) For bed hold up 680 g (output tracer curve)				
1.	220	0.10E-01	0.22- 01	0.11E-01
2.	230	0.21	0.12	0.16
3.	240	0.13	0.53	0.57

4.	250	0.42E+01	0.12E+01	0.42E+01
5.	260	0.13E+02	0.12E+02=	0.11E+02
6.	270	0.11E+02	0.12E+02	0.10E+02
7.	280	0.12E+02	0.83E+01	0.75E+01
8.	300	0.910 E+01	0.76E+01	0.70E+01
9.	330	0.54E+01	0.41E+01	0.32E+01
10.	360	0.40	0.82	0.76
11.	420	0.20	0.15	0.10
12.	480	0.42	0.10E-01	0.81E-02

Table C-2

MOVING BED WITH MIXED FEED OF 0.724 mm AND 0.96 mm PARTICLES

Solid feed = 1:1 Ratio of 0.724 mm & 9.96 mm sizes
Solid feed rate = 90 g/min of each size
u/u_{mf} = 0.95, Tracer Amount = 10.0g

S. No.	Time (Seconds)	C for tracer size: 0.724 mm only	0.96 mm only	1:1 of two sizes
(a) For bed hold up of 340 g (Input signal curve)				
1.	60	0.10E-01	0.21	0.42E-01
2.	70	0.18	0.28	0.11
3.	80	0.43E+01	0.40E+01	0.92
4.	90	0.11E+02	0.82E+01	0.62E+01
5.	100	0.10E+02	0.12E+02	0.13E+02
6.	110	0.11E+02	0.11E+02	0.10E+02
7.	120	0;95E+01	0.10E+02	0.11E+02
8.	150	0.87E+01	0.75E+01	0.38E+01
9.	180	0.45E+01	0.26E+01	0.43E+01
10.	240	0.13	0.22	0.13E+01
11.	300	0.31	0.11E-01	0.12
(b) For bed hold up 680 g (output tracer curve)				
1.	220	0.76E-01	0.28	0.43
2.	230	0.11	0.61	0.21

Continued

3.	240	0.82	0.12E+01	0.51E+01
4.	250	0.22E+01	0.64E+01	0.92E+01
5.	260	0.11E+02	0.12E+02	0.11E+02
6.	270	0.12E+02	0.11E+01	0.10E+02
7.	280	0.10+02	0.97E+01	0.64E+01
8.	300	0.62E+01	0.37E+01	0.22E+01
9.	330	0.22E+01	0.43E+01	0.58
10.	360	0.13+01	0.64	0.76
11.	420	0.12	0.29	0.42
12.	480	0.97E-01	0.11	0.24

Table C-3

MOVING BED WITH MIXED FEED OF THREE SIZES

Solid feed = 1:1:1 of 0.393 mm, 0.724 mm and 0.96 mm particles
Solid feed rate = 60 g/min of each size (i.e. 180 g/min of total)
u/u_{mf} = 0.95 Tracer Amount = 10 g.

S. No.	Time Seconds	C for tracer sizes			
		0.393 mm only	0.724 mm only	0.96 mm only	1:1:1 of three sizes
(a) For bed hold up 340 g (Input signal curve)					
1.	60	0.10	0.35E-01	0.48	0.56
2.	70	0.20	0.16	0.27	0.82
3.	80	0.11E+01	0.88	0.13E+01	0.91
4.	90	0.12E+02	0.11E+02	0.11E+02	0.97E+01
5.	100	0.11E+02	0.12E+02	0.11E+02	0.10E+02
6.	110	0.10E+02	0.98E+01	0.10E+02	0.87E+01
7.	120	0.95E+01	0.67E+01	0.54E+01	0.64E+01
8.	150	0.67E+01	0.23E+01	0.42E+01	0.54E+01
9.	180	0.54E+01	0.11E+01	0.97	0.21E+01
10.	240	0.23E+01	0.11E+01	0.13E+01	0.97
11.	300	0.11E+01	0.56	0.24	0.56
(b) For bed hold up 680 g (Output tracer curve)					
1.	220	0.60E-01	0.51E-01	0.23E-01	0.12
2.	230	0.58E-01	0.17	0.19E-01	0.21
3.	240	0.37	0.53	0.32	0.56

Continued

4.	250	0.31E+01	0.92	0.76	0.18E+01
5.	260	0.11E+02	0.12E+02	0.12E+02	0.12E+02
6.	270	0.10E+02	0.87E+01	0.11E+02	0.97E+01
7.	280	0.10E+02	0.83E+01	0.97E+01	0.72E+01
8.	300	0.87E+01	0.73E+01	0.90+01	0.71E+01
9.	330	0.64E+01	0.56E+01	0.41E+01	0.62E+01
10.	360	0.52+01	0.15E+01	0.85	0.22E+01
11.	420	0.32E+01	0.12E+01	0.71	0.82
12.	480	0.64	0.27	0.12	0.87E-01

APPENDIX - D

FLUIDIZED BEDS WITH SINGLE AND MIXED FEED SIZES

Table D-1

FLUIDIZED BED WITH SINGLE FEED SIZES

In all, four runs are tabulated here. The conditions of the runs are:

Run No.	$\frac{u}{u_{mf}}$	h cm	W grams / min.	d_p mm
1.	1.2	10	34	0.393
2.	1.2	10	300	0.96
3.	1.2	30	300	0.393
4.	6.0	10	300	0.393

S. No.	Time (Seconds)	C for run number			
		1	2	3	4
1.	10	0.0	0.10	0.02	1.03
2.	20	0.0	0.41	0.20	0.87
3.	30	0.013	1.18	0.59	0.85
4.	40	0.039	0.91	0.80	0.80
5.	50	0.150	0.59	1.35	0.79
6.	60	0.350	0.65	1.39	0.75
7.	70	0.50	0.12	1.46	0.71
8.	80	0.91	0.18	1.41	0.73
9.	90	1.30	0.13	1.40	0.60
10.	100	1.30	0.13	1.40	0.60
11.	120	1.57	0.10	0.95	0.57
12.	140	1.40	0.031	0.51	0.52

Continued

13.	160	1.00	0.041	0.51	0.46
14.	180	1.02	0.013	0.40	0.43
15.	200	0.93	0.030	0.42	0.36
16.	220	0.80	0.03[illegible]	0.31	0.37
17.	240	0.85	0.011	0.25	0.34
18.	260	0.70	0.013	0.28	0.25
19.	280	0.72	0.010	0.21	0.22
20.	300	0.70	0.0082	0.20	0.22
21.	360	0.60	0.0082	0.15	0.20
22.	420	0.60	-	0.15	0.12
23.	480	0.43	-	0.071	0.056
24.	600	0.40	-	0.042	0.039
25.	720	0.29	-	0.051	0.040
26.	840	0.23	-	0.035	0.022
27.	960	0.20	-	0.024	0.002

Table D-2

FLUIDIZED BED WITH MIXED FEEDS OF SIZES 0.393 mm and 0.724mm

u/u_{mf} = 3.0, Tracer amount = 10 g total
Solid feed rate = 90 grams per minute of 0.393 mm particles and 90 grams per minute of 0.724 mm particles
Total Solid hold up = 340 g
Inlet and Outlet stream compositions 1:1 of 0.393 mm and 0.724 mm dia particles.

S. No.	Time in Seconds	C for tracer sizes: 0.393 mm only	0.724 mm only	50% of each size
1.	10	0.92	1.20	1.01
2.	20	0.81	0.95	0.87
3.	30	0.71	0.85	0.78
4.	40	0.67	0.78	0.70
5.	50	0.61	0.67	0.64
6.	60	0.50	0.67	0.62
7.	70	0.45	0.58	0.55
8.	80	0.44	0.58	0.52
9.	90	0.36	0.55	0.47
10.	100	0.34	0.52	0.41
11.	110	0.33	0.45	0.40
12.	120	0.25	0.44	0.32
13.	150	0.18	0.36	0.28
14.	180	0.14	0.27	0.23
15.	210	0.10	0.21	0.16
16.	240	0.062	0.16	0.11

17.	270	0.048	0.13	0.083
18.	300	0.037	0.12	0.070
19.	360	0.020	0.076	0.042
20.	420	0.006	0.042	0.021
21.	480	0.006	0.034	0.017

Table - D-3

FLUIDIZED BED WITH MIXED FEED SIZES OF 0.724 mm AND 0.960 mm

u/u_{mf} = 3.0, Tracer Amount = 10 g

Solid feed rate: 90 grams per min. of 0.724 mm size
and 90 grams per min of 0.96 mm size.

Total solid hold up = 340 g

Inlet and Outlet streams composition: 1:1 ratio of 0.724 mm and 0.926 mm dia particles

S. No.	Time in Seconds	C for tracer sizes		
		0.724 mm only	0.96 mm only	50% 0.724 and 50% 0.960 mm
1.	10	0.96	1.05	1.12
2.	20	0.96	1.05	1.12
3.	30	0.74	0.82	0.79
4.	40	0.64	0.78	0.70
5.	50	0.60	0.69	0.65
6.	60	0.52	0.61	0.62
7.	70	0.41	0.55	0.55
8.	80	0.38	0.56	0.47
9.	90	0.32	0.53	0.48
10.	100	0.31	0.50	0.40
11.	110	0.26	0.46	0.37
12.	120	0.22	0.42	0.32
13.	150	0.19	0.35	0.26
14.	180	0.15	0.28	0.25
15.	210	0.065	0.22	0.14

16.	240	0. 042	0. 14	0. 13
17.	270	0. 046	0. 14	0. 065
18.	300	0. 021	0. 10	0. 10
19.	360	0. 047	0. 11	0. 032
20.	420	0. 011	0. 021	0. 020
21.	480	0. 008	0. 040	0. 03

Table D-4

FLUIDIZED BED WITH MIXED FEEDS OF THREE SIZES

u/u_{mf} = 3.0, Tracer Amount = 10.0 g

Solid feed rate: 60 g per min. each of the sizes 0.393 mm, 0.724 mm and 0.96 mm with a total of 180 g/min.

Inlet and outlet stream compositions: 1:1:1 ratios of the three sizes i.e., 0.393 mm, 0.724 mm and 0.96mm

S. No.	Time in Seconds.	C for tracer sizes			
		0.393 mm only	0.724 mm only	0.96 mm only	33-1/3 % of each size
1.	10	0.84	0.96	1.20	1.07
2.	20	0.73	0.82	0.93	0.88
3.	30	0.62	0.70	0.86	0.79
4.	40	0.51	0.65	0.78	0.62
5.	50	0.47	0.61	0.70	0.58
6.	60	0.40	0.52	0.68	0.57
7.	70	0.31	0.44	0.64	0.52
8.	80	0.25	0.40	0.62	0.48
9.	90	0.23	0.40	0.54	0.43
10.	100	0.20	0.35	0.56	0.37
11.	110	0.15	0.27	0.50	0.37
12.	120	0.15	0.29	0.46	0.31
13.	150	0.094	0.19	0.44	0.18
14.	180	0.058	0.15	0.30	0.14
15.	210	0.037	0.09	0.23	0.14

16.	240	0.018	0.07	0.21	0.12
17.	270	0.012	0.042	0.17	0.062
18.	300	0.0088	0.03	0.13	0.057
19.	360	0.0037	0.023	0.11	0.033
20.	420	-	0.006	0.088	0.015
21.	480	-	0.0045	0.05	0.013

APPENDIX E

HOLD UP RATIOS IN FLUIDIZED BED WITH FEEDS OF TWO SIZE

TABLE E.1

SOLID FEED = 60.0 G/MIN EACH OF 0.393MM AND 0.724MM GLASS BALLOTINIS
SOLID FEED RATE = 120.0 G/MIN
BED HOLD UP = 90.0 GRAM
STATIC BED HEIGHT= 2.5CM

AVG. d_p MM	τ_S MIN	τ_{GA} MIN	τ_{GB} MIN	τ_A	τ_B	U/U_{mf}	H(2,1)
.538	.750	1.111E-03	5.359E-04	675.06	1399.50	1.882	1.448
.548	.750	1.050E-03	4.465E-04	713.64	1679.40	2.182	1.640
.557	.750	1.008E-03	3.827E-04	744.00	1959.30	2.476	1.818
.566	.750	9.383E-04	3.349E-04	799.27	2239.20	2.746	1.984
.569	.750	8.971E-04	2.965E-04	836.00	2529.46	3.076	2.100
.568	.750	8.707E-04	2.679E-04	861.29	2799.00	3.414	2.102
.569	.750	8.353E-04	2.444E-04	897.79	3068.53	3.726	2.111
.569	.750	8.182E-04	2.246E-04	916.55	3338.06	4.051	2.090
.567	.750	7.875E-04	2.067E-04	952.29	3628.33	4.432	2.065
.567	.750	7.600E-04	1.808E-04	986.76	4146.66	5.075	2.000
.569	.750	7.541E-04	1.682E-04	994.47	4457.66	5.421	2.000

TABLE E.2

SOLID FEED = 60.0 G/MIN EACH OF 0.393MM AND 0.724MM GLASS BALLOTINIS
SOLID FEED RATE = 120.0 G/MIN
BED HOLD UP = 160.0 GRAM
STATIC BED HEIGHT= 5.0CM

AVG. d_P MM	τ'_S MIN	T_{GA} MIN	T_{GB} MIN	τ_A	τ_B	U/U_{mf}	H(2,1)
.550	1.333	2.282E-03	9.527E-04	584.26	1399.49	1.806	1.681
.572	1.333	2.201E-03	7.939E-04	605.51	1679.39	2.020	2.190
.571	1.333	2.037E-03	6.805E-04	654.38	1959.29	2.368	2.149
.565	1.333	1.876E-03	5.954E-04	710.46	2239.19	2.758	2.000
.555	1.333	1.794E-03	5.271E-04	743.11	2529.46	3.221	1.767
.550	1.333	1.741E-03	4.763E-04	765.59	2798.99	3.622	1.666
.551	1.333	1.684E-03	4.345E-04	791.55	3068.53	3.948	1.645
.544	1.333	1.636E-03	3.994E-04	814.71	3338.06	4.403	1.553
.543	1.333	1.621E-03	3.674E-04	822.30	3628.33	4.806	1.531
.541	1.333	1.520E-03	3.215E-04	877.12	4146.66	5.516	1.489
.544	1.333	1.508E-03	2.991E-04	883.97	4457.66	5.880	1.480

TABLE E.3

SOLID FEED = 60.0 G/MIN EACH OF 0.393MM AND 0.724MM GLASS BALLOTINIS
SOLID FEED RATE = 120.0 G/MIN
BED HOLD UP = 320.0 GRAM
STATIC BED HEIGHT= 10.0CM

AVG. d_p MM	τ_S MIN	T_{GA} MIN	T_{GB} MIN	τ_A	τ_B	U/U_{mf}	H(2,1)
.531	2.000	4.804E-03	1.905E-03	416.29	1049.62	1.927	1.328
.544	2.000	4.504E-03	1.587E-03	444.04	1259.55	2.214	1.538
.549	2.000	4.075E-03	1.361E-03	490.78	1469.47	2.538	1.659
.548	2.000	3.847E-03	1.190E-03	519.85	1679.40	2.917	1.626
.545	2.000	3.654E-03	1.054E-03	547.20	1897.10	3.319	1.544
.547	2.000	3.528E-03	9.527E-04	566.86	2099.25	3.651	1.584
.536	2.000	3.423E-03	8.690E-04	584.16	2301.40	4.152	1.525
.535	2.000	3.336E-03	7.988E-04	599.50	2503.55	4.539	1.391
.535	2.000	3.242E-03	7.349E-04	616.72	2721.25	4.926	1.400
.538	2.000	3.040E-03	6.430E-04	657.84	3110.00	5.588	1.403
.538	2.000	3.016E-03	5.982E-04	662.98	3343.25	5.992	1.399

TABLE E.4

SOLID FEED = 60.0 G/MIN EACH OF 0.393MM AND 0.724MM GLASS BALLOTINIS
SOLID FEED RATE = 120.0 G/MIN
BED HOLD UP = 620.0 GRAM
STATIC BED HEIGHT= 20.0CM

AVG. d_p MM	τ_S MIN	τ_{GA} MIN	τ_{GB} MIN	τ_A	τ_B	U/U_{mf}	H(2,1)
.530	5.166	9.608E-03	3.691E-03	537.70	1399.50	1.937	1.303
.534	5.166	8.883E-03	3.076E-03	581.63	1679.40	2.287	1.384
.537	5.166	8.150E-03	2.636E-03	633.93	1959.30	2.645	1.428
.539	5.166	7.506E-03	2.307E-03	688.26	2239.20	3.000	1.468
.538	5.166	7.309E-03	2.042E-03	706.80	2529.46	3.404	1.444
.536	5.166	7.056E-03	1.845E-03	732.19	2799.00	3.789	1.414
.535	5.166	6.847E-03	1.683E-03	754.54	3068.53	4.173	1.390
.535	5.166	6.672E-03	1.547E-03	774.36	3338.06	4.542	1.388

TABLE E.5

SOLID FEED = 90.0 G/MIN EACH OF 0.393MM AND 0.724MM GLASS BALLOTINIS
SOLID FEED RATE = 180.0 G/MIN
BED HOLD UP = 90.0 GRAM
STATIC BED HEIGHT= 2.5CM

AVG. P MM	S MIN	GA MIN	GB MIN	A	B	U/U	H(2,1)
.535	.500	1.111E-03	5.359E-04	450.04	933.00	1.904	1.389
.547	.500	1.050E-03	4.465E-04	475.76	1119.60	2.191	1.578
.550	.500	1.008E-03	3.827E-04	496.00	1306.20	2.533	1.691
.556	.500	9.383E-04	3.349E-04	532.85	1492.80	2.842	1.790
.559	.500	8.971E-04	2.965E-04	557.33	1686.31	3.172	1.870
.560	.500	8.707E-04	2.679E-04	574.19	1866.00	3.505	1.903
.554	.500	8.353E-04	2.444E-04	598.52	2045.68	3.909	1.830
.556	.500	8.182E-04	2.246E-04	611.03	2225.37	4.226	1.806
.556	.500	7.875E-04	2.067E-04	634.86	2418.88	4.598	1.800
.554	.500	7.600E-04	1.808E-04	657.84	2764.44	5.289	1.757
.554	.500	7.541E-04	1.682E-04	662.98	2971.77	5.685	1.674

TABLE E.6

SOLID FEED = 150.0 G/MIN EACH OF 0.393MM AND 0.724MM GLASS BALLOTINIS
SOLID FEED RATE = 300.0 G/MIN
BED HOLD UP = 90.0 GRAM
STATIC BED HEIGHT= 2.5CM

AVG. d_P MM	τ_S MIN	τ_{GA} MIN	τ_{GB} MIN	τ_A	τ_B	U/U_{mf}	H(2,1)
.533	.300	1.111E-03	5.359E-04	270.02	559.80	1.915	1.358
.541	.300	1.050E-03	4.465E-04	285.45	671.76	2.237	1.500
.544	.300	1.008E-03	3.827E-04	297.60	783.72	2.582	1.558
.545	.300	8.971E-04	2.965E-04	334.40	1011.78	3.321	1.579
.544	.300	8.707E-04	2.679E-04	344.51	1119.60	3.690	1.557
.543	.300	8.353E-04	2.444E-04	359.11	1227.41	4.067	1.546
.542	.300	8.182E-04	2.246E-04	366.62	1335.22	4.430	1.520
.541	.300	7.875E-04	2.067E-04	380.91	1451.33	4.826	1.507
.542	.300	7.600E-04	1.808E-04	394.70	1658.66	5.509	1.477

SET NO. 47

AVG.DP TS TGA TGB TA TB U/UMF R(1,2)

TABLE E.7

SOLID FEED = 250.0 G/MIN EACH OF 0.393MM AND 0.724MM GLASS BALLOTINIS
SOLID FEED RATE = 500.0 G/MIN
BED HOLD UP = 90.0 GRAM
STATIC BED HEIGHT= 2.5CM

AVG. d_P MM	τ_S MIN	T_{GA} MIN	T_{GB} MIN	τ_A	τ_B	U/U	H(2,1)
.527	.180	1.111E-03	5.359E-04	162.01	335.88	1.956	1.259
.531	.180	1.050E-03	4.465E-04	171.27	403.05	2.313	1.328
.532	.180	1.008E-03	3.827E-04	178.56	470.23	2.686	1.350
.534	.180	9.383E-04	3.349E-04	191.82	537.40	3.053	1.378
.533	.180	8.971E-04	2.965E-04	200.64	607.07	3.461	1.361
.529	.180	8.707E-04	2.679E-04	206.70	671.76	3.882	1.310
.529	.180	8.353E-04	2.444E-04	215.47	736.44	4.252	1.300
.528	.180	8.182E-04	2.246E-04	219.97	801.13	4.655	1.270
.527	.180	7.875E-04	2.067E-04	228.55	870.80	5.074	1.257
.524	.180	7.600E-04	1.808E-04	236.82	995.20	5.861	1.194
.523	.180	7.541E-04	1.682E-04	238.67	1069.84	6.314	1.170

TABLE E.8

SOLID FEED = 50.0 G/MIN EACH OF 0.393MM AND 0.724MM GLASS BALLOTINIS
SOLID FEED RATE = 100.0 G/MIN
BED HOLD UP = 160.0 GRAM
STATIC BED HEIGHT= 5.0CM

AVG. d_p MM	τ_S MIN	T_{GA} MIN	T_{GB} MIN	τ_A	τ_B	U/U_{mf}	H(2,1)
.549	.888	1.111E-03	9.527E-04	800.07	933.00	1.815	1.650
.563	.888	1.050E-03	7.939E-04	845.79	1119.60	2.084	1.945
.558	.888	1.008E-03	6.805E-04	881.78	1306.20	2.464	1.851
.551	.888	9.383E-04	5.954E-04	947.29	1492.80	2.890	1.683
.545	.888	8.971E-04	5.271E-04	990.82	1686.31	3.324	1.536
.536	.888	8.707E-04	4.763E-04	1020.78	1866.00	3.786	1.417
.535	.888	8.353E-04	4.345E-04	1064.05	2045.68	4.175	1.388
.532	.888	8.182E-04	3.994E-04	1086.28	2225.37	4.579	1.347
.531	.888	7.875E-04	3.674E-04	1128.65	2418.88	4.997	1.328
.531	.888	7.600E-04	3.215E-04	1169.49	2764.44	5.722	1.319
.531	.888	7.541E-04	2.991E-04	1178.63	2971.77	6.151	1.320

TABLE E.1

SOLID FEED = 150.0 G/MIN EACH OF 0.393MM AND 0.724MM GLASS BALLOTINIS
SOLID FEED RATE = 300.0 G/MIN
BED HOLD UP = 160.0 GRAM
STATIC BED HEIGHT= 5.0CM

AVG. d_P MM	τ_S MIN	T_{GA} MIN	T_{GB} MIN	τ_A	τ_B	U/U_{mf}	H(2,1)
.541	.533	1.111E-03	9.527E-04	480.04	559.80	1.864	1.500
.555	.533	1.050E-03	7.939E-04	507.47	671.76	2.139	1.766
.549	.533	1.008E-03	6.805E-04	529.07	783.72	2.538	1.661
.537	.533	9.383E-04	5.954E-04	568.37	895.68	3.021	1.430
.531	.533	8.971E-04	5.271E-04	594.49	1011.78	3.491	1.318
.524	.533	8.707E-04	4.763E-04	612.47	1119.60	3.947	1.219
.523	.533	8.353E-04	4.345E-04	638.43	1227.41	4.346	1.200
.522	.533	8.182E-04	3.994E-04	651.76	1335.22	4.750	1.180
.522	.533	7.875E-04	3.674E-04	677.19	1451.33	5.164	1.179
.524	.533	7.600E-04	3.215E-04	701.69	1658.66	5.851	1.187
.524	.533	7.541E-04	2.991E-04	707.18	1783.06	6.287	1.189

TABLE E.16

SOLID FEED = 250.0 G/MIN EACH OF 0.393MM AND 0.724MM GLASS BALLOTINIS
SOLID FEED RATE = 580.0 G/MIN
BED HOLD UP = 320.0 GRAM
STATIC BED HEIGHT= 10.0CM

AVG. d_p MM	τ_S MIN	T_{GA} MIN	T_{GB} MIN	τ_A	τ_B	U/U_{mf}	H(2,1)
.533	.320	1.111E-03	9.527E-04	288.02	335.88	1.916	1.358
.541	.320	1.050E-03	7.939E-04	304.48	403.05	2.237	1.500
.527	.320	1.008E-03	6.805E-04	317.44	470.23	2.736	1.248
.521	.320	9.383E-04	5.954E-04	341.02	537.40	3.193	1.171
.533	.320	8.971E-04	5.271E-04	356.69	607.07	3.463	1.358
.517	.320	8.707E-04	4.763E-04	367.48	671.76	4.044	1.102
.513	.320	8.353E-04	4.345E-04	383.05	736.44	4.496	1.060
.513	.320	8.182E-04	3.994E-04	391.06	801.13	4.891	1.060
.514	.320	7.875E-04	3.674E-04	406.31	870.80	5.300	1.059
.517	.320	7.600E-04	3.215E-04	421.01	995.20	6.002	1.068
.517	.320	7.541E-04	2.991E-04	424.30	1069.84	6.449	1.058

TABLE E.11

SOLID FEED = 90.0 G/MIN EACH OF 0.393MM AND 0.724MM GLASS BALLOTINIS
SOLID FEED RATE = 180.0 G/MIN
BED HOLD UP = 320.0 GRAM
STATIC BED HEIGHT= 10.0CM

AVG. d_P MM	τ_S MIN	$\bar{T}_{GA}$ MIN	$\bar{T}_{GB}$ MIN	τ_A	τ_B	U/U_{mf}	H(2,1)
.527	1.777	1.111E-03	1.905E-03	1600.15	933.00	1.956	1.258
.537	1.777	1.050E-03	1.587E-03	1691.59	1119.60	2.266	1.430
.544	1.777	1.008E-03	1.361E-03	1763.57	1306.20	2.584	1.515
.541	1.777	9.383E-04	1.190E-03	1894.58	1492.80	2.982	1.500
.537	1.777	8.971E-04	1.054E-03	1981.64	1686.31	3.419	1.439
.534	1.777	8.707E-04	9.527E-04	2041.57	1865.99	3.816	1.378
.531	1.777	8.353E-04	8.690E-04	2128.10	2045.68	4.235	1.318
.529	1.777	8.182E-04	7.988E-04	2172.56	2225.37	4.635	1.289
.529	1.777	7.875E-04	7.349E-04	2257.30	2418.88	5.043	1.270
.530	1.777	7.600E-04	6.430E-04	2338.99	2764.44	5.725	1.269
.533	1.777	7.541E-04	5.982E-04	2357.26	2971.77	6.104	1.260

TABLE E.12

SOLID FEED = 150.0 G/MIN EACH OF 0.393MM AND 0.724MM GLASS BALLOTINIS
SOLID FEED RATE = 300.0 G/MIN
BED HOLD UP = 320.0 GRAM
STATIC BED HEIGHT= 10.0CM

AVG. d_p MM	τ_S MIN	T_{GA} MIN	T_{GB} MIN	τ_A	τ_B	U/U_{mf}	H(2,1)
.522	1.066	1.111E-03	1.905E-03	960.09	559.80	1.991	1.180
.529	1.066	1.050E-03	1.587E-03	1014.95	671.75	2.327	1.300
.531	1.066	1.008E-03	1.361E-03	1058.14	783.72	2.700	1.325
.530	1.066	9.383E-04	1.190E-03	1136.75	895.68	3.093	1.314
.529	1.066	8.971E-04	1.054E-03	1188.98	1011.78	3.509	1.309
.525	1.066	8.707E-04	9.527E-04	1224.94	1119.59	3.937	1.261
.523	1.066	8.353E-04	8.690E-04	1276.86	1227.41	4.346	1.200
.523	1.066	8.182E-04	7.988E-04	1303.53	1335.22	4.730	1.198
.526	1.066	7.875E-04	7.349E-04	1354.38	1451.33	5.082	1.190
.525	1.066	7.600E-04	6.430E-04	1403.39	1658.66	5.835	1.170
.524	1.066	7.541E-04	5.982E-04	1414.36	1783.06	6.287	1.293

TABLE E.13

SOLID FEED = 250.0 G/MIN EACH OF 0.393MM AND 0.724MM GLASS BALLOTINIS
SOLID FEED RATE = 500.0 G/MIN
BED HOLD UP = 320.0 GRAM
STATIC BED HEIGHT= 10.0CM

AVG. d_p MM	τ_S MIN	T_{GA} MIN	T_{GB} MIN	τ_A	τ_B	U/U_{mf}	H(2,1)
.516	.640	1.111E-03	1.905E-03	576.05	335.88	2.029	1.101
.519	.640	1.050E-03	1.587E-03	608.97	403.05	2.411	1.169
.522	.640	1.008E-03	1.361E-03	634.88	470.23	2.782	1.189
.521	.640	9.383E-04	1.190E-03	682.05	537.40	3.194	1.170
.518	.640	8.971E-04	1.054E-03	713.39	607.07	3.651	1.177
.519	.640	8.707E-04	9.527E-04	734.96	671.76	4.020	1.140
.518	.640	8.353E-04	8.690E-04	766.11	736.44	4.429	1.120
.519	.640	8.182E-04	7.988E-04	782.12	801.13	4.803	1.104
.518	.640	7.875E-04	7.349E-04	812.62	870.80	5.240	1.078
.518	.640	7.600E-04	6.430E-04	842.03	995.20	5.983	1.069
.518	.640	7.541E-04	5.982E-04	848.61	1069.84	6.421	1.048

TABLE E.14

SOLID FEED = 240 G/MIN OF 0.393MM AND 60 G/MIN OF 0.724MM BALLOTINIS
SOLID FEED RATE = 300.0 G/MIN
BED HOLD UP = 90.0 GRAM
STATIC BED HEIGHT = 2.5CM

AVG. d_p MM	τ_S MIN	T_{GA} MIN	T_{GB} MIN	τ_A	τ_B	U/U_{mf}	H(2,1)
.450	.300	1.111E-03	5.359E-04	270.02	559.80	2.609	1.538
.452	.300	1.050E-03	4.465E-04	285.45	671.76	3.103	1.611
.453	.300	1.008E-03	3.827E-04	297.60	783.72	3.611	1.633
.451	.300	9.383E-04	3.349E-04	319.71	895.68	4.150	1.612
.447	.300	8.971E-04	2.965E-04	334.40	1011.78	4.763	1.581
.450	.300	8.707E-04	2.679E-04	344.51	1119.60	5.206	1.558
.450	.300	8.353E-04	2.444E-04	359.11	1227.41	5.708	1.511
.445	.300	8.182E-04	2.246E-04	366.62	1335.22	6.349	1.520
.444	.300	7.875E-04	2.067E-04	380.91	1451.33	6.937	1.454
.447	.300	7.600E-04	1.808E-04	394.70	1658.66	7.819	1.449
.447	.300	7.541E-04	1.682E-04	397.78	1783.06	8.417	1.438

TABLE E.15

SOLID FEED = 60 G/MIN OF 0.393MM AND 240 G/MIN OF 0.724MM BALLOTINIS
SOLID FEED RATE = 300.0 G/MIN
BED HOLD UP = 90.0 GRAM
STATIC BED HEIGHT= 2.5CM

AVG. d_P MM	τ_S MIN	T_{GA} MIN	T_{GB} MIN	τ_A	τ_B	U/U_{mf}	H(2,1)
.635	.300	1.111E-03	5.359E-04	270.02	559.80	1.394	1.183
.643	.300	1.050E-03	4.465E-04	285.45	671.76	1.636	1.289
.647	.300	1.008E-03	3.827E-04	297.60	783.72	1.887	1.377
.645	.300	9.383E-04	3.349E-04	319.71	895.68	2.169	1.471
.649	.300	8.971E-04	2.965E-04	334.40	1011.78	2.418	1.593
.648	.300	8.707E-04	2.679E-04	344.51	1119.60	2.687	1.552
.653	.300	8.353E-04	2.444E-04	359.11	1227.41	2.900	1.564
.650	.300	8.182E-04	2.246E-04	366.62	1335.22	3.181	1.600
.655	.300	7.875E-04	2.067E-04	380.91	1451.33	3.418	1.595
.651	.300	7.600E-04	1.808E-04	394.70	1658.66	3.950	1.557
.647	.300	7.541E-04	1.682E-04	397.78	1783.06	4.295	1.520

TABLE E.16

SOLID FEED = 60 G/MIN OF 0.393MM AND 240 G/MIN OF 0.724MM BALLOTINIS
SOLID FEED RATE = 300.0 G/MIN
BED HOLD UP = 160.0 GRAM
STATIC BED HEIGHT= 5.0CM

AVG. d_P MM	τ_S MIN	T_{GA} MIN	T_{GB} MIN	τ_A	τ_B	U/U_{mf}	H(2,1)
.454	.533	1.111E-03	9.527E-04	480.04	559.80	2.566	1.676
.452	.533	1.050E-03	7.939E-04	507.47	671.76	3.105	1.607
.446	.533	1.008E-03	6.805E-04	529.07	783.72	3.711	1.457
.441	.533	9.383E-04	5.954E-04	568.37	895.68	4.322	1.330
.442	.533	8.971E-04	5.271E-04	594.49	1011.78	4.867	1.237
.441	.533	8.707E-04	4.763E-04	612.47	1119.60	5.406	1.210
.437	.533	8.353E-04	4.345E-04	638.43	1227.41	6.021	1.188
.438	.533	8.182E-04	3.994E-04	651.76	1335.22	6.531	1.190
.438	.533	7.875E-04	3.674E-04	677.19	1451.33	7.107	1.200
.435	.533	7.600E-04	3.215E-04	701.69	1658.66	8.196	1.188
.433	.533	7.541E-04	2.991E-04	707.18	1783.06	8.899	1.079

TABLE E.17

SOLID FEED = 240 G/MIN OF 0.393MM AND 60 G/MIN OF 0.724MM BALLOTINIS
SOLID FEED RATE = 300.0 G/MIN
BED HOLD UP = 160.0 GRAM
STATIC BED HEIGHT = 5.0CM

AVG. d_P MM	τ_S MIN	T_{GA} MIN	T_{GB} MIN	τ_A	τ_B	U/U_{mf}	H(2,1)
.637	.533	1.111E-03	9.527E-04	480.04	559.80	1.385	1.260
.645	.533	1.050E-03	7.939E-04	507.47	671.76	1.623	1.500
.655	.533	1.008E-03	6.805E-04	529.07	783.72	1.841	1.727
.656	.533	9.383E-04	5.954E-04	568.37	895.68	2.102	1.739
.647	.533	8.971E-04	5.271E-04	594.49	1011.78	2.435	1.480
.644	.533	8.707E-04	4.763E-04	612.47	1119.60	2.718	1.298
.636	.533	8.353E-04	4.345E-04	638.43	1227.41	3.044	1.248
.635	.533	8.182E-04	3.994E-04	651.76	1335.22	3.324	1.221
.632	.533	7.875E-04	3.674E-04	677.19	1451.33	3.646	1.209
.634	.533	7.600E-04	3.215E-04	701.69	1658.66	4.144	1.198
.632	.533	7.541E-04	2.991E-04	707.18	1783.06	4.474	1.199

TABLE E.18

SOLID FEED = 90.0 G/MIN EACH OF 0.724MM AND 0.960MM GLASS BALLOTINIS
SOLID FEED RATE = 180.0 G/MIN
BED HOLD UP = 90.0 GRAM
STATIC BED HEIGHT= 2.5CM

AVG. d_P MM	τ_S MIN	T_{GA} MIN	T_{GB} MIN	τ_A	τ_B	U/U_{mf}	H(2,1)
.839	.500	6.943E-04	3.349E-04	720.07	1492.80	1.344	1.262
.848	.500	6.977E-04	2.965E-04	716.57	1686.31	1.488	1.320
.844	.500	7.056E-04	2.679E-04	708.58	1866.00	1.661	1.378
.843	.500	6.847E-04	2.444E-04	730.20	2045.68	1.823	1.417
.845	.500	6.798E-04	2.246E-04	735.50	2225.37	1.974	1.450
.850	.500	6.717E-04	2.067E-04	744.32	2418.88	2.124	1.502
.852	.500	6.181E-04	1.808E-04	808.82	2764.44	2.418	1.569
.854	.500	5.855E-04	1.607E-04	853.92	3110.00	2.712	1.628
.855	.500	5.512E-04	1.446E-04	906.95	3455.55	3.002	1.697
.855	.500	5.527E-04	1.315E-04	904.53	3801.11	3.307	1.657
.853	.500	4.989E-04	1.113E-04	1002.18	4492.22	3.926	1.600

TABLE E.19

SOL+D FEED = 90.0 G/MIN EACH OF 0.724MM AND 0.960MM GLASS BALLOTINIS
SOLID FEED RATE = 180.0 G/MIN
BED HOLD UP = 160.0 GRAM
STATIC BED HEIGHT= 5.0CM

AVG. d_p MM	τ_S MIN	T_{GA} MIN	T_{GB} MIN	τ_A	τ_B	U/U_{mf}	H(2,1)
.840	.888	6.943E-04	5.954E-04	1280.12	1492.80	1.339	1.330
.853	.888	6.977E-04	5.271E-04	1273.91	1686.31	1.471	1.638
.860	.888	7.056E-04	4.763E-04	1259.69	1866.00	1.606	1.853
.944	.888	6.847E-04	4.345E-04	1298.14	2045.68	1.806	1.930
.860	.888	6.798E-04	3.994E-04	1307.56	2225.37	1.912	1.830
.857	.888	6.717E-04	3.674E-04	1323.24	2418.88	2.095	1.676
.850	.888	6.181E-04	3.215E-04	1437.90	2764.44	2.429	1.530
.846	.888	5.855E-04	2.858E-04	1518.09	3110.00	2.757	1.411
.843	.888	5.512E-04	2.572E-04	1612.35	3455.55	3.085	1.349
.843	.888	5.527E-04	2.338E-04	1608.05	3801.11	3.393	1.319
.842	.888	4.989E-04	1.978E-04	1781.65	4492.22	4.019	1.328

TABLE E.20

SOL+D FEED = 90.0 G/MIN EACH OF 0.724MM AND 0.960MM GLASS BALLOTINIS
SOLID FEED RATE = 180.0 G/MIN
BED HOLD UP = 320.0 GRAM
STATIC BED HEIGHT= 10.0CM

AVG. P MM	τ_S MIN	T_{GA} MIN	T_{GB} MIN	τ_A	τ_B	U/U_{mf}	H(2,1)
.834	1.777	6.943E-04	1.190E-03	2560.25	1492.80	1.357	1.229
.938	1.777	6.977E-04	1.054E-03	2547.83	1686.31	1.469	1.310
.843	1.777	7.056E-04	9.527E-04	2519.39	1865.99	1.663	1.349
.842	1.777	6.847E-04	8.690E-04	2596.28	2045.68	1.826	1.380
.844	1.777	6.798E-04	7.988E-04	2615.12	2225.37	1.979	1.407
.845	1.777	6.717E-04	7.349E-04	2646.49	2418.88	2.149	1.400
.844	1.777	6.181E-04	6.430E-04	2875.81	2764.44	2.461	1.377
.842	1.777	5.855E-04	5.716E-04	3036.19	3109.99	2.779	1.340
.838	1.777	5.512E-04	5.144E-04	3224.71	3455.55	3.114	1.300
.841	1.777	5.527E-04	4.676E-04	3216.11	3801.11	3.406	1.280
.842	1.777	4.989E-04	3.957E-04	3563.31	4492.22	4.019	1.267

TABLE E.21

SOL+D FEED = 90.0 G/MIN EACH OF 0.724MM AND 0.960MM GLASS BALLOTINIS
SOLID FEED RATE = 180.0 G/MIN
BED HOLD UP = 620.0 GRAM
STATIC BED HEIGHT= 20.0CM

AVG. d_P MM	τ_S MIN	T_{GA} MIN	T_{GB} MIN	τ_A	τ_B	U/U_{mf}	H(2,1)
.838	3.444	6.943E-04	2.307E-03	4960.48	1492.80	1.346	1.205
.838	3.444	6.977E-04	2.042E-03	4936.42	1686.31	1.521	1.259
.843	3.444	7.056E-04	1.845E-03	4881.32	1866.00	1.664	1.298
.844	3.444	6.847E-04	1.683E-03	5030.30	2045.68	1.820	1.320
.844	3.444	6.798E-04	1.547E-03	5066.80	2225.37	1.981	1.318
.842	3.444	6.717E-04	1.423E-03	5127.57	2418.88	2.163	1.330
.841	3.444	6.181E-04	1.245E-03	5571.88	2764.44	2.478	1.308
.840	3.444	5.855E-04	1.107E-03	5882.62	3109.99	2.793	1.289
.838	3.444	5.512E-04	9.967E-04	6247.88	3455.55	3.115	1.279
.842	3.444	5.527E-04	9.061E-04	6231.22	3801.11	3.400	1.267
.840	3.444	4.989E-04	7.667E-04	6903.91	4492.22	4.034	1.258

TABLE E.22

SOL+D FEED = 150.0 G/MIN EACH OF 0.724MM AND 0.960MM GLASS BALLOTINIS
SOLID FEED RATE = 300.0 G/MIN
BED HOLD UP = 90.0 GRAM
STATIC BED HEIGHT= 2.5CM

AVG. d_P MM	τ_S MIN	T_{GA} MIN	T_{GB} MIN	τ_A	τ_B	U/U_{mf}	H(2,1)
.829	.300	6.943E-04	3.349E-04	432.04	895.68	1.372	1.100
.834	.300	6.977E-04	2.965E-04	429.94	1011.78	1.535	1.159
.837	.300	7.056E-04	2.679E-04	425.14	1119.60	1.687	1.220
.839	.300	6.847E-04	2.444E-04	438.12	1227.41	1.840	1.270
.841	.300	6.798E-04	2.246E-04	441.30	1335.22	1.991	1.309
.842	.300	6.717E-04	2.067E-04	446.59	1451.33	2.161	1.342
.846	.300	6.181E-04	1.808E-04	485.29	1658.66	2.450	1.430
.851	.300	5.855E-04	1.607E-04	512.35	1866.00	2.727	1.555
.850	.300	5.512E-04	1.446E-04	544.17	2073.33	3.034	1.541
.853	.300	5.527E-04	1.315E-04	542.71	2280.66	3.318	1.501
.850	.300	4.989E-04	1.113E-04	601.30	2695.33	3.946	1.480

TABLE E.23

SOLID FEED = 250.0 G/MIN EACH OF 0.724MM AND 0.960MM GLASS BALLOTINIS
SOLID FEED RATE = 500.0 G/MIN
BED HOLD UP = 90.0 GRAM
STATIC BED HEIGHT= 2.5CM

AVG. d_P MM	τ_S^* MIN	T_{GA} MIN	T_{GB} MIN	τ_A	τ_B	U/U_{mf}	H(2,1)
.830	.180	6.943E-04	3.349E-04	259.22	537.40	1.369	1.080
.831	.180	6.977E-04	2.965E-04	257.96	607.07	1.543	1.111
.835	.180	7.056E-04	2.679E-04	255.08	671.76	1.692	1.166
.835	.180	6.847E-04	2.444E-04	262.87	736.44	1.857	1.172
.835	.180	6.798E-04	2.246E-04	264.78	801.13	2.019	1.217
.839	.180	6.717E-04	2.067E-04	267.95	870.80	2.178	1.247
.841	.180	6.181E-04	1.808E-04	291.17	995.20	2.477	1.309
.839	.180	5.855E-04	1.607E-04	307.41	1119.60	2.796	1.343
.843	.180	5.512E-04	1.446E-04	326.50	1244.00	3.081	1.379
.843	.180	5.527E-04	1.315E-04	325.63	1368.40	3.388	1.368
.842	.180	4.989E-04	1.113E-04	360.78	1617.20	4.017	1.349

TABLE E.24

SOL+D FEED = 150.0 G/MIN EACH OF 0.724MM AND 0.960MM GLASS BALLOTINIS
SOLID FEED RATE = 300.0 G/MIN
BED HOLD UP = 160.0 GRAM
STATIC BED HEIGHT= 5.0CM

AVG. d_P MM	τ_S MIN	T_{GA} MIN	T_{GB} MIN	τ_A	τ_B	U/U_{mf}	H(2,1)
.838	.533	6.943E-04	5.954E-04	768.07	895.68	1.346	1.230
.841	.533	6.977E-04	5.271E-04	764.35	1011.78	1.509	1.372
.849	.533	7.056E-04	4.763E-04	755.81	1119.60	1.641	1.551
.851	.533	6.847E-04	4.345E-04	778.88	1227.41	1.793	1.562
.850	.533	6.798E-04	3.994E-04	784.53	1335.22	1.956	1.526
.847	.533	6.717E-04	3.674E-04	793.94	1451.33	2.140	1.448
.843	.533	6.181E-04	3.215E-04	862.74	1658.66	2.464	1.349
.838	.533	5.855E-04	2.858E-04	910.85	1866.00	2.803	1.268
.838	.533	5.512E-04	2.572E-04	967.41	2073.33	3.118	1.228
.836	.533	5.527E-04	2.338E-04	964.83	2280.66	3.442	1.209
.836	.533	4.989E-04	1.978E-04	1068.99	2695.33	4.071	1.200

TABLE E.25

SOL+D FEED = 250.0 G/MIN EACH OF 0.724MM AND 0.960MM GLASS BALLOTINIS
SOLID FEED RATE = 500.0 G/MIN
BED HOLD UP = 160.0 GRAM
STATIC BED HEIGHT= 5.0CM

AVG. d_P MM	τ_S MIN	T_{GA} MIN	T_{GB} MIN	τ_A	τ_B	U/U_{mf}	H(2,1)
.833	.320	6.943E-04	5.954E-04	460.84	537.40	1.362	1.139
.837	.320	6.977E-04	5.271E-04	458.61	607.07	1.525	1.220
.840	.320	7.056E-04	4.763E-04	453.49	671.76	1.674	1.298
.841	.320	6.847E-04	4.345E-04	467.33	736.44	1.831	1.319
.837	.320	6.798E-04	3.994E-04	470.72	801.13	2.010	1.261
.835	.320	6.717E-04	3.674E-04	476.36	870.80	2.194	1.206
.835	.320	6.181E-04	3.215E-04	517.64	995.20	2.511	1.151
.831	.320	5.855E-04	2.858E-04	546.51	1119.60	2.849	1.142
.831	.320	5.512E-04	2.572E-04	580.44	1244.00	3.163	1.109
.831	.320	5.527E-04	2.338E-04	578.90	1368.40	3.483	1.101
.831	.320	4.989E-04	1.978E-04	641.39	1617.20	4.112	1.111

TABLE E.26

SOL+D FEED = 240 G/MIN OF 0.724MM AND 60 G/MIN OF 0.960MM BALLOTINIS
SOLID FEED RATE = 300.0 G/MIN
BED HOLD UP = 90.0 GRAM
STATIC BED HEIGHT= 2.5CM

AVG. d_p MM	τ_S MIN	T_{GA} MIN	T_{GB} MIN	τ_A	τ_B	U/U_{mf}	H(2,1)
.895	.300	6.943E-04	3.349E-04	432.04	895.68	1.193	1.437
.897	.300	6.977E-04	2.965E-04	429.94	1011.78	1.342	1.462
.896	.300	7.056E-04	2.679E-04	425.14	1119.60	1.488	1.500
.898	.300	6.847E-04	2.444E-04	438.12	1227.41	1.625	1.538
.900	.300	6.798E-04	2.246E-04	441.30	1335.22	1.763	1.566
.900	.300	6.717E-04	2.067E-04	446.59	1451.33	1.913	1.613
.900	.300	6.181E-04	1.808E-04	485.29	1658.66	2.188	1.652
.900	.300	5.855E-04	1.607E-04	512.35	1866.00	2.463	1.640
.902	.300	5.512E-04	1.446E-04	544.17	2073.33	2.725	1.645
.900	.300	5.527E-04	1.315E-04	542.71	2280.66	3.013	1.629
.902	.300	4.989E-04	1.113E-04	601.30	2695.33	3.545	1.610

TABLE E.27

SOL+D FEED = 60 G/MIN OF 0.724MM AND 240 G/MIN OF 0.960MM BALLOTINIS
SOLID FEED RATE = 300.0 G/MIN
BED HOLD UP = 90.0 GRAM
STATIC BED HEIGHT= 2.5CM

AVG. d_P MM	τ_S MIN	T_{GA} MIN	T_{GB} MIN	τ_A	τ_B	U/U_{mf}	H(2,1)
.905	.300	6.943E-04	3.349E-04	432.04	895.68	1.169	1.110
.905	.300	6.977E-04	2.965E-04	429.94	1011.78	1.321	1.147
.909	.300	7.056E-04	2.679E-04	425.14	1119.60	1.450	1.207
.910	.300	6.847E-04	2.444E-04	438.12	1227.41	1.589	1.261
.913	.300	6.798E-04	2.246E-04	441.30	1335.22	1.717	1.324
.913	.300	6.717E-04	2.067E-04	446.59	1451.33	1.864	1.366
.914	.300	6.181E-04	1.808E-04	485.29	1658.66	2.128	1.424
.916	.300	5.855E-04	1.607E-04	512.35	1866.00	2.383	1.526
.918	.300	5.512E-04	1.446E-04	544.17	2073.33	2.638	1.555
.919	.300	5.527E-04	1.315E-04	542.71	2280.66	2.900	1.615
.919	.300	4.989E-04	1.113E-04	601.30	2695.33	3.422	1.638

APPENDIX F

HOLD UP RATIOS IN FLUIDIZED BED WITH FEEDS OH THREE SIZES

TABLE F.1

SOLID FEED = 60.0 G/MIN EACH OF 0.393MM, 0.724MM AND 0.96MM BALLOTINIS
SOLID FEED RATE = 180.0 G/MIN
BED HOLD UP = 93.3 GRAM
STATIC BED HEIGHT= 2.5CM

AVG. d_P MM	τ_S MIN	T_{GA} MIN	T_{GB} MIN	τ_A	τ_B	U/U_{mf}	H(2,1)	H(3,1)	H(3,2)
.643	.500	9.258E-04	4.465E-04	540.05	1119.60	1.635	1.186	1.555	1.311
.673	.500	9.008E-04	3.827E-04	555.05	1306.20	1.757	1.315	1.851	1.407
.686	.500	8.820E-04	3.349E-04	566.86	1492.80	1.935	1.531	2.333	1.523
.704	.500	8.306E-04	2.965E-04	601.92	1686.31	2.087	1.626	2.703	1.661
.704	.500	8.107E-04	2.679E-04	616.72	1866.00	2.310	1.775	2.935	1.653
.709	.500	7.942E-04	2.444E-04	629.48	2045.68	2.502	1.618	2.697	1.666
.704	.500	7.650E-04	2.238E-04	653.52	2233.67	2.768	1.537	2.531	1.646
.692	.500	7.528E-04	2.067E-04	664.16	2418.88	3.090	1.461	2.401	1.640
.686	.500	6.891E-04	1.808E-04	725.56	2764.44	3.584	1.311	1.977	1.607
.667	.500	6.976E-04	1.682E-04	716.73	297 .77	4.053	1.195	1.875	1.600

TABLE F.2

SOLID FEED = 60.0 G/MIN EACH OF 0.393MM, 0.724MM AND 0.96MM BALLOTINIS
SOLID FEED RATE = 180.0 G/MIN
BED HOLD UP = 160.3 GRAM
STATIC BED HEIGHT= 5.0CM

AVG. d_p MM	τ_S MIN	T_{GA} MIN	T_{GB} MIN	τ_A	τ_B	U/U_{mf}	H(2,1)	H(3,1)	H(3,2)
.707	.888	9.258E-04	7.939E-04	960.09	1119.60	1.376	1.497	2.662	1.777
.706	.888	9.008E-04	6.805E-04	986.76	1306.20	1.610	1.618	3.006	1.857
.704	.888	8.820E-04	5.954E-04	1007.75	1492.80	1.847	1.696	2.948	1.798
.692	.888	8.306E-04	5.271E-04	1070.09	1686.31	2.153	1.427	2.500	1.751
.673	.888	8.107E-04	4.763E-04	1096.40	1866.00	2.508	1.298	1.977	1.522
.660	.888	7.942E-04	4.345E-04	1119.08	2045.68	2.851	1.223	1.736	1.419
.653	.888	7.650E-04	3.979E-04	1161.82	2233.67	3.168	1.166	1.641	1.406
.649	.888	7.528E-04	3.674E-04	1180.74	2418.88	3.468	1.121	1.598	1.425
.646	.888	6.891E-04	3.215E-04	1289.88	2764.44	3.998	1.068	1.509	1.412
.642	.888	6.976E-04	2.991E-04	1274.19	2971.77	4.346	1.077	1.519	1.409

TABLE F.3

SOLID FEED = 60.0 G/MIN EACH OF 0.393MM, 0.724MM AND 0.96MM BALLOTINIS
SOLID FEED RATE = 180.0 G/MIN
BED HOLD UP = 320.3 GRAM
STATIC BED HEIGHT= 10.0CM

AVG. d_p MM	τ_S MIN	T_{GA} MIN	T_{GB} MIN	τ_A	τ_B	U/U_{mf}	H(2,1)	H(3,1)	H(3,2)
.650	1.777	9.258E-04	1.587E-03	1920.18	1119.60	1.603	1.269	1.700	1.340
.645	1.777	9.008E-04	1.361E-03	1973.52	1306.20	1.894	1.308	1.863	1.424
.661	1.777	8.820E-04	1.190E-03	2015.51	1492.80	2.074	1.372	1.975	1.438
.660	1.777	8.306E-04	1.054E-03	2140.18	1686.31	2.346	1.324	1.889	1.427
.654	1.777	8.107E-04	9.527E-04	2192.80	1865.99	2.641	1.259	1.762	1.398
.645	1.777	7.942E-04	8.690E-04	2238.17	2045.68	2.968	1.187	1.639	1.380
.640	1.777	7.650E-04	7.958E-04	2323.65	2233.67	3.288	1.142	1.562	1.367
.634	1.777	7.528E-04	7.349E-04	2361.48	2418.88	3.627	1.032	1.367	1.324
.628	1.777	6.891E-04	6.430E-04	2579.77	2764.44	4.209	1.064	1.379	1.295
.633	1.777	6.976E-04	5.982E-04	2548.39	2971.77	4.457	1.047	1.339	1.279

TABLE F.4

SOLID FEED = 100.0 G/MIN EACH OF 0.393MM, 0.724MM AND 0.96MM BALLOTINIS
SOLID FEED RATE = 300.0 G/MIN
BED HOLD UP = 93.3 GRAM
STATIC BED HEIGHT= 2.5CM

AVG. d_P MM	τ_S MIN	T_{GA} MIN	T_{GB} MIN	τ_A	τ_B	U/U_{mf}	H(2,1)	H(3,1)	H(3,2)
.629	.300	9.258E-04	4.465E-04	324.03	671.76	1.702	1.076	1.307	1.214
.646	.300	9.008E-04	3.827E-04	333.03	783.72	1.893	1.197	1.587	1.325
.664	.300	8.820E-04	3.349E-04	340.11	895.68	2.056	1.371	1.926	1.405
.678	.300	8.306E-04	2.965E-04	361.15	1011.78	2.233	1.463	2.170	1.483
.680	.300	8.107E-04	2.679E-04	370.03	1119.60	2.460	1.487	2.205	1.482
.672	.300	7.942E-04	2.444E-04	377.69	1227.41	2.757	1.414	2.096	1.482
.667	.300	7.650E-04	2.238E-04	392.11	1340.20	3.051	1.344	1.957	1.456
.662	.300	7.528E-04	2.067E-04	398.50	1451.33	3.350	1.278	1.872	1.464
.652	.300	6.891E-04	1.808E-04	435.33	1658.66	3.932	1.180	1.691	1.433
.650	.300	6.976E-04	1.682E-04	430.04	1783.06	4.248	1.122	1.555	1.385

TABLE F.5

SOLID FEED = 166.7 G/MIN EACH OF 0.393MM, 0.724MM AND 0.96MM BALLOTINIS
SOLID FEED RATE = 500.0 G/MIN
BED HOLD UP = 93.3 GRAM
STATIC BED HEIGHT= 2.5CM

d_P MM	τ_S MIN	τ_{GA} MIN	τ_{GB} MIN	τ_A	τ_B	U/U_{mf}	H(2,1)	H(3,1)	H(3,2)
.624	.179	9.258E-04	4.465E-04	194.03	402.25	1.728	1.098	1.260	1.148
.644	.179	9.008E-04	3.827E-04	199.42	469.29	1.904	1.225	1.500	1.224
.649	.179	8.820E-04	3.349E-04	203.66	536.33	2.143	1.247	1.662	1.332
.659	.179	8.306E-04	2.965E-04	216.26	605.86	2.353	1.293	1.741	1.345
.663	.179	8.107E-04	2.679E-04	221.57	670.41	2.579	1.324	1.804	1.363
.654	.179	7.942E-04	2.444E-04	226.16	734.97	2.897	1.211	1.643	1.356
.649	.179	7.650E-04	2.238E-04	234.80	802.51	3.206	1.212	1.561	1.288
.649	.179	7.528E-04	2.067E-04	238.62	869.06	3.476	1.163	1.632	1.403
.642	.179	6.891E-04	1.808E-04	260.68	993.21	4.048	1.115	1.460	1.309
.637	.179	6.976E-04	1.682E-04	257.51	1067.70	4.411	1.075	1.373	1.277

TABLE F.6

SOLID FEED = 100.0 G/MIN EACH OF 0.393MM, 0.724MM AND 0.96MM BALLOTINIS
SOLID FEED RATE = 300.0 G/MIN
BED HOLD UP = 160.3 GRAM
STATIC BED HEIGHT= 5.0CM

AVG. d_P MM	τ_S MIN	τ_{GA} MIN	τ_{GB} MIN	τ_A	τ_B	U/U_{mf}	H(2,1)	H(3,1)	H(3,2)
.681	.533	9.258E-04	7.939E-04	576.05	671.76	1.472	1.317	2.042	1.550
.689	.533	9.008E-04	6.805E-04	592.05	783.72	1.679	1.446	2.378	1.644
.679	.533	8.820E-04	5.954E-04	604.65	895.68	1.974	1.389	2.169	1.561
.658	.533	8.306E-04	5.271E-04	642.05	1011.78	2.361	1.254	1.800	1.435
.648	.533	8.107E-04	4.763E-04	657.84	1119.60	2.685	1.167	1.560	1.336
.645	.533	7.942E-04	4.345E-04	671.45	1227.41	2.971	1.127	1.460	1.294
.641	.533	7.650E-04	3.979E-04	697.09	1340.20	3.279	1.096	1.366	1.245
.638	.533	7.528E-04	3.674E-04	708.44	1451.33	3.579	1.083	1.362	1.257
.631	.533	6.891E-04	3.215E-04	773.93	1658.66	4.174	1.050	1.323	1.260
.624	.533	6.976E-04	2.991E-04	764.51	1783.06	4.580	1.024	1.254	1.224

TABLE F.7

SOLID FEED = 166.7 G/MIN EACH OF 0.393MM, 0.724MM AND 0.96MM BALLOTINIS
SOLID FEED RATE = 500.0 G/MIN
BED HOLD UP = 160.3 GRAM
STATIC BED HEIGHT= 5.0CM

AVG. d_P MM	τ_S MIN	T_{GA} MIN	T_{GB} MIN	τ_A	τ_B	U/U_{mf}	H(2,1)	H(3,1)	H(3,2)
.656	.320	9.258E-04	7.939E-04	345.63	403.05	1.577	1.219	1.655	1.356
.660	.320	9.008E-04	6.805E-04	355.23	470.23	1.817	1.289	1.861	1.443
.649	.320	8.820E-04	5.954E-04	362.79	537.40	2.140	1.238	1.610	1.300
.644	.320	8.306E-04	5.271E-04	385.23	607.07	2.455	1.155	1.499	1.297
.634	.320	8.107E-04	4.763E-04	394.70	671.76	2.797	1.102	1.372	1.244
.626	.320	7.942E-04	4.345E-04	402.87	736.44	3.131	1.078	1.290	1.195
.626	.320	7.650E-04	3.979E-04	418.25	804.12	3.424	1.061	1.248	1.175
.624	.320	7.528E-04	3.674E-04	425.06	870.80	3.730	1.040	1.210	1.163
.629	.320	6.891E-04	3.215E-04	464.35	995.20	4.205	1.028	1.190	1.158
.622	.320	6.976E-04	2.991E-04	458.71	1069.84	4.602	1.016	1.176	1.158

TABLE F.8

SOLID FEED = 200.0 G/MIN OF 0.393MM, 50.0 G/MIN OF 0.724MM AND 50.0 G/MIN OF 0.96MM BALLOTINIS

SOLID FEED RATE = 300.0 G/MIN

BED HOLD UP = 93.3 GRAM

STATIC BED HEIGHT= 2.5CM

d_p MM	τ_S MIN	τ_{GA} MIN	τ_{GB} MIN	τ_A	τ_B	U/U_{mf}	H(2,1)	H(3,1)	H(3,2)
.513	.300	9.258E-04	4.465E-04	324.03	671.76	2.465	1.315	1.578	1.200
.523	.300	9.008E-04	3.827E-04	333.03	783.72	2.774	1.454	1.910	1.314
.522	.300	8.820E-04	3.349E-04	340.11	895.68	3.180	1.462	2.039	1.394
.523	.300	8.306E-04	2.965E-04	361.15	1011.78	3.586	1.362	1.988	1.459
.518	.300	8.107E-04	2.679E-04	370.03	1119.60	4.041	1.299	1.959	1.507
.523	.300	7.942E-04	2.444E-04	377.69	1227.41	4.349	1.271	1.921	1.511
.513	.300	7.650E-04	2.238E-04	392.11	1340.20	4.923	1.242	1.839	1.481
.507	.300	7.528E-04	2.067E-04	398.50	1451.33	5.431	1.210	1.760	1.454
.502	.300	6.891E-04	1.808E-04	435.33	1658.66	6.339	1.161	1.607	1.383
.504	.300	6.976E-04	1.682E-04	430.04	1783.06	6.753	1.147	1.594	1.390

TABLE F.9

SOLID FEED = 50.0 G/MIN OF 0.393MM, 50.0 G/MIN OF 0.724MM AND200.0 G/MIN OF 0.96MM BALLOTINIS
SOLID FEED RATE = 300.0 G/MIN
BED HOLD UP = 93.3 GRAM
STATIC BED HEIGHT= 2.5CM

AVG. d_P MM	τ_S MIN	T_{GA} MIN	T_{GB} MIN	τ_A	τ_B	U/U_{mf}	H(2,1)	H(3,1)	H(3,2)
.750	.300	9.258E-04	4.465E-04	324.03	671.76	1.234	1.071	1.117	1.042
.780	.300	9.008E-04	3.827E-04	333.03	783.72	1.342	1.242	1.427	1.148
.798	.300	8.820E-04	3.349E-04	340.11	895.68	1.472	1.321	1.621	1.226
.806	.300	8.306E-04	2.965E-04	361.15	1011.78	1.632	1.460	1.929	1.320
.812	.300	8.107E-04	2.679E-04	370.03	1119.60	1.781	1.521	2.099	1.380
.810	.300	7.942E-04	2.444E-04	377.69	1227.41	1.961	1.444	2.090	1.447
.814	.300	7.650E-04	2.238E-04	392.11	1340.20	2.123	1.368	2.052	1.500
.808	.300	7.528E-04	2.067E-04	398.50	1451.33	2.333	1.277	1.902	1.489
.791	.300	6.891E-04	1.808E-04	435.33	1658.66	2.765	1.157	1.689	1.459
.789	.300	6.976E-04	1.682E-04	430.04	1783.06	2.989	1.130	1.619	1.432

TABLE F.10

SOLID FEED = 50.0 G/MIN OF 0.393MM, 200.0 G/MIN OF 0.724MM AND 50.0 G/MIN OF 0.96MM BALLOTINIS

SOLID FEED RATE = 300.0 G/MIN

BED HOLD UP = 93.3 GRAM

STATIC BED HEIGHT= 2.5CM

AVG. d_p MM	τ_s MIN	T_{GA} MIN	T_{GB} MIN	τ_A	τ_B	U/U_{mf}	H(2,1)	H(3,1)	H(3,2)
.670	.300	9.258E-04	4.465E-04	324.03	671.76	1.516	1.017	1.344	1.322
.665	.300	9.008E-04	3.827E-04	333.03	783.72	1.793	1.070	1.256	1.173
.666	.300	8.820E-04	3.349E-04	340.11	895.68	2.042	1.200	1.175	.979
.683	.300	8.306E-04	2.965E-04	361.15	1011.78	2.205	1.481	1.407	.950
.683	.300	8.107E-04	2.679E-04	370.03	1119.60	2.437	1.397	1.450	1.037
.685	.300	7.942E-04	2.444E-04	377.69	1227.41	2.659	1.455	1.493	1.026
.688	.300	7.650E-04	2.238E-04	392.11	1340.20	2.884	1.465	1.636	1.116
.688	.300	7.528E-04	2.067E-04	398.50	1451.33	3.118	1.453	1.597	1.099
.687	.300	6.891E-04	1.808E-04	435.33	1658.66	3.580	1.309	1.726	1.318
.682	.300	6.976E-04	1.682E-04	430.04	1783.06	3.892	1.227	1.508	1.228

TABLE F.11

SOLID FEED = 134.0 G/MIN OF 0.393MM, 34.0 G/MIN OF 0.724MM AND 134 G/MIN OF 0.96MM BALLOTINIS
SOLID FEED RATE = 300.0 G/MIN
BED HOLD UP = 93.3 GRAM
STATIC BED HEIGHT= 2.5CM

d_P MM	τ_S MIN	TGA MIN	TGB MIN	τ_A	τ_B	U/U_{mf}	H(2,1)	H(3,1)	H(3,2)
.540	.300	9.258E-04	4.465E-04	324.03	671.76	2.247	.744	.762	1.024
.558	.300	9.008E-04	3.827E-04	333.03	783.72	2.464	.857	.986	1.150
.609	.300	8.820E-04	3.349E-04	340.11	895.68	2.403	1.154	1.427	1.236
.620	.300	8.306E-04	2.965E-04	361.15	1011.78	2.632	1.176	1.555	1.322
.626	.300	8.107E-04	2.679E-04	370.03	1119.60	2.862	1.113	1.531	1.375
.636	.300	7.942E-04	2.444E-04	377.69	1227.41	3.047	1.246	1.814	1.455
.662	.300	7.650E-04	2.238E-04	392.11	1340.20	3.093	1.473	2.113	1.433
.640	.300	7.528E-04	2.067E-04	398.50	1451.33	3.560	1.266	1.831	1.445
.655	.300	6.891E-04	1.808E-04	435.33	1658.66	3.906	1.370	2.055	1.500
.654	.300	6.976E-04	1.682E-04	430.04	1783.06	4.211	1.333	1.983	1.487

APPENDIX G

AXIAL AND RADIAL DISTRIBUTION OF SOLIDS IN FLUIDIZED BED

Table G-1

EFFECT OF SIZE OF OUTLET CONNECTION ON HOLD UP RATIOS

(Outlet connection at bottom of bed. The size of opening was changed by inserting polythene tube of required amount of bore).

Solid feed rate = 60 g per min each of 0.393 mm and 0.724 mm sizes

Total bed hold up = 90 g.

S. No.	u/u_{mf}	Hold up ratios for		
		dia. of outlet 13 mm	dia of outlet 10 mm	dia of outlet 6 mm
1.	1.9	1.48	1.49	1.52
2.	2.17	1.64	1.67	1.73
3.	2.51	1.82	1.835	1.85
4.	2.79	1.99	1.97	2.01
5.	3.11	2.10	2.12	2.11
6.	3.44	2.105	2.12	2.14
7.	3.78	2.11	2.105	2.12
8.	4.12	2.09	2.11	2.10
9.	4.48	2.06	2.10	2.09
10.	5.17	2.0	2.10	2.09

Table G - 2

HOLD UP RATIOS FOR OVERFLOW AND DOWN COMER ARRANGEMENT

Solid feed composition = 1:1:1 proportions of three sizes

Height of overflow tube = 5 cm.

Runs taken for solid feed rates of 180 g/ min and 500 g/ min

S. No.	u/u_{mf}	Runs for W=180 g/ min			Runs for w = 500 g/ min	
		H(1, 2)	H(2, 3)	u/u_{mf}	H(1, 2)	H(2, 3)
1.	1.139	1.09	1.10	1.134	1.12	1.11
2.	1.613	1.07	1.12	1.621	1.06	1.09
3.	1.922	1.04	1.08	1.926	1.06	1.07
4.	2.245	1.03	1.06	2.24	1.07	1.05
5.	2.655	1.04	1.07	2.641	1.02	1.04
6.	2.997	1.02	1.06	2.992	1.05	1.06
7.	3.231	1.06	1.06	3.229	1.03	1.03
8.	3.522	1.02	1.07	3.520	1.04	1.02
9.	4.023	1.03	1.03	4.001	1.02	1.01
10.	4.305	1.01	1.05	4.320	1.01	1.01
11.	4.712	1.01	1.05	4.721	1.01	1.01

Table G - 3

EXPERIMENTS WITH OUTLET CONNECTION

Observation recorded for 2 sets at following conditions:

(a) Set No. 1

i) Outlet connection used as overflow
ii) Solid feed rate = 60 g per min each of 0.393 and 0.724 mm sizes

(b) Set No. 2

i) Exit feed of solids controlled by stopcock at the outlet connections.
ii) Bed hold up = 270 g.
iii) Solid feed rate = same as set 1.

S. No.	Set No. 1			Set No. 2		
	u/u_{mf}	Static bed height cm	Hold up ratio H(1, 2)	u/u_{mf}	Fluidized bed ht.	hold up ratio H(1, 2)
1.	2.105	6.4	1.15	2.105	12.5	1.13
2.	2.53	6.3	1.07	2.451	14.0	1.18
3.	2.96	5.8	1.06	2.88	15.2	1.20
4.	3.47	5.5	1.055	3.16	16.5	1.22
5.	3.86	5.0	1.06	3.59	17.4	1.23
6.	4.21	4.8	1.03	3.98	18.5	1.26
7.	4.619	4.6	1.03	4.35	19.6	1.30
8.	5.07	4.5	1.01	4.71	20.8	1.355
9.	6.24	4.4	1.00	5.55	22.2	1.42
10.	6.72	4.3	1.00	6.01	25	1.517

TABLE G-4
EFFECT OF BED HEIGHT ON HOLD-UP RATIO

i) Solid Feed Rate : 60 g per minute each of 0.393 mm and 0.724 mm
ii) Particles withdrawn through outlet connection at 5 cm height and controlled by stopcock

S. No.	P mm H2O	h cm	Fluidized bed ht. in cm for G = 80 LPM	108 LPM	135 LPM	161 LPM	H(1, 2) for G = 80 LPM	108 LPM	135 LPM	161 LPM
1.	76	5.5	11	12	12.7	13.2	1.065	1.07	1.06	1.01
2.	82	6.5	12	13	15	15.5	1.065	1.07	1.12	1.06
3.	100	8.0	13.5	15	16	20.8	1.215	1.23	1.25	1.35
4.	118	9.0	15.3	17	19	21.5	1.305	1.305	1.26	1.33
5.	140	10.0	18	21.5	21	26	1.32	1.35	1.33	1.35
6.	194	13.7	26	29	30	33	1.30	1.30	1.31	1.30

Table G - 5

EFFECT OF THE RADIAL POSITION OF OUTLET CONNECTION ON HOLD UP RATIOS

i) Feed conditions: 60 g/min each of 0.393 mm and 0.724 mm sizes

ii) Bed hold up = 320 g.

iii) Outlet connection at 5 cm from bottom of the bed.

S. No.	h cm	Fluidized bed height cm	u/u_{mf}	Hold up ratio H(1 ↑) for opening position			
				2mm inside	4 mm inside	6 mm inside	8 mm inside
1.	10	16	2.06	0.97	0.965	0.922	0.96
2.	10	18	2.46	0.99	0.965	0.55	0.985
3.	10	19	2.88	1.025	0.987	0.988	0.995
4.	10	20	3.30	1.04	1.002	0.987	0.986
5.	10	21	3.72	1.065	1.007	0.992	1.01
6.	10	22.5	4.12	1.104	1.032	1.01	1.00
7.	10	24	4.52	1.10	1.04	1.006	1.012
8.	10	25	4.91	1.075	1.024	1.007	1.003

Table G- 6

AXIAL AND RADIAL DISTRIBUTION OF SOLIDS

(Experiments with batch fluidization only)
Total bed hold up = 160 g.
Overall bed composition = 1:1 of .393 mm and .724 mm particles

S. No.	u/u_{mf}	Axial Position ↓ / Radial Position →	Hold Up Ratio: Central core 25 mm dia	Middle annular layer between 25 mm dia and 42 mm dia.	Outer layer between 42 mm dia and 50 mm dia.
		Top 20%	0.555	0.95	-
1.	2.08	Middle 40%	0.85	1.27	-
		bottom 40%	1.52	1.13	-
		Overall	0.937	1.1	0.845
		Top 20%	0.834	1.25	-
2.	2.49	Middle 40%	0.978	1.08	-
		Bottom 40%	1.08	0.99	-
		Overall	0.923	1.12	0.836
		Top 20%	0.96	1.19	-
3.	3.32	Middle 40%	0.96	1.13	-
		Bottom 40%	1.14	1.10	-
		Overall	1.02	1.15	0.738
		Top 20%	1.11	1.19	-
4.	4.26	Middle 40%	1.08	1.12	-
		Bottom 40%	1.185	1.09	-
		Overall	1.025	1.13	0.766
		Top 20%	1.09	1.095	-
5.	4.95	Middle 40 %	0.995	1.20	-
		Bottom 40%	0.995	1.075	-
		Overall	1.022	1.12	0.772
		Top 20%	1.15	1.03	-
6.	6.61	Middle 40 %	1.14	1.00	-
		Bottom 40%	1.17	1.4	-
		Overall	1.155	1.02	0.801

Table G-7

AXIAL AND RADIAL DISTRIBUTION OF SOLIDS

Experiments with batch fluidization of 0.724 mm and 0.96 mm particles
Total bed hold up = 160 g
Bed composition = 1:1 of 0.724 and 0.96 mm sizes

S. No.	u/u_{mf}	Radial Position → / Axial Position ↓	Hold up Ratio: Central Core of 25 mm dia	Hold up Ratio: Annulus between 25 mm and 42 mm dia	Hold up Ratio: Outer layer between 42 mm and 50 mm dia
		Top 20%	1.305	1.125	-
1.	1.04	Middle 40%	1.168	1.27	-
		Bottom 40%	1.078	1.00	-
		Overall	1.182	1.122	0.682
		Top 20%	1.105	1.39	-
2.	1.39	Middle 40%	1.072	1.04	-
		Bottom 40%	1.2	1.00	-
		Overall	1.18	1.13	0.685
		Top 20%	1.00	1.19	-
3.	2.08	Middle 40%	1.07	1.07	-
		Bottom 40%	1.1	1.2	-
		Overall	1.055	1.15	0.640
		Top 20%	1.045	1.27	-
4.	2.73	Middle 40%	1.00	1.125	
		Bottom 40%	1.30	1.19	-
		Overall	1.09	1.182	0.635
		Top 20%	1.268	1.222	
5.	3.47	Middle 40%	1.08	1.284	
		Bottom 40%	1.00	1.182	
		Overall	1.11	1.23	0.676
		Top 20%	1.06	1.18	
6.	4.13	Middle 40%	1.08	1.28	
		Bottom 40%	1.00	1.09	
		Overall	1.07	1.17	0.710

Table G-8

AXIAL AND RADIAL DISTRIBUTION OF SOLIDS

Experiments with batch fluidization of 0.393 mm, 0.724 mm & 0.96 mm sizes
Bed hold up = 160 g.
Overall Bed composition = 1:1:1 of 0.393, 0.724 mm and 0.96 mm sizes.

S. No.	u/u_{mf}	Radial Position → / Axial Position ↓	H (1,2) Central core 25 dia.	H (1,2) Annulus between 25 & 42 mm dia	H (1,2) Overlayer bet. 42 & 50 mm dia.	H(2,3) Central core 25 mm dia.	H(2,3) Annulus bet. 25 & 42 mm dia	H(2,3) outer layer bet. 42 & 50 mm dia.
1.	1.14	Top 20%	1.0	1.11	-	0.944	1.05	-
		Middle 40%	0.889	0.945	-	1.00	1.00	-
		Bottom 40%	1.4	1.077		1.43	1.143	
		Overall	1.03	1.03	0.834	1.06	1.08	0.900
2.	2.01	Top 20%	1.00	1.13	-	1.00	1.28	-
		Middle 40%	0.91	1.143		1.00	1.19	-
		Bottom 40%	1.125	1.07	-	1.12	1.06	-
		Overall	1.00	1.135	0.740	1.03	1.18	0.600
3.	3.13	Top 20%	1.22	1.00	-	1.18	1.196	-
		Middle 40%	0.90	1.23	-	0.89	1.125	-
		Bottom 40%	1.125	1.07	-	1.22	1.266	-
		Overall	1.072	1.1	0.742	1.102	1.19	0.700
4.	4.03	Top 20%	1.08	1.125	-	1.077	1.22	-
		Middle 40%	0.90	1.00	-	1.00	1.125	-
		Bottom 40%	1.04	1.07	-	1.102	1.13	-
		Overall	1.012	1.065	0.742	1.065	1.162	0.82
5.	4.81	Top 20%	1.035	1.078	-	1.14	1.24	-
		Middle 40%	1.142	1.06	-	1.05	1.18	-
		Bottom 40%	1.24	1.125	-	1.124	1.14	-
		Overall	1.122	1.09	0.778	1.11	1.184	0.83

APPENDIX H

HOLD UP RATIOS IN FLUIDIZED BEDS WITH BAFFLES

Table H-1

RUNS ON SIEVE PLATE BAFFLES OF LOW SPACING

Solid Feed: 1:1:1 of 0.393 mm, 0.724 mm and 0.960 mm particles.

Solid feed rate = 180 g/min.

S.	$h_D = 0.5$			$h_D = 1.0$		
No.	u/u_{mf}	H(2,1)	H(3,2)	u/u_{mf}	H(2,1)	H(3,2)
1.	2.211	0.746	0.850	2.534	0.443	0.723
2.	2.322	0.761	0.869	2.786	0.484	0.745
3.	2.763	0.759	0.855	3.066	0.533	0.764
4.	2.926	0.801	0.872	3.374	0.566	0.740
5.	3.192	0.822	0.883	3.668	0.555	0.762
6.	3.402	0.829	0.878	3.962	0.617	0.771
7.	4.018	0.846	0.890	4.298	0.592	0.773
8.	4.648	0.820	0.887	4.662	0.592	0.768

Table H-2

RUNS ON SIEVE PLATE OF LARGE SPACING

Solid feed: 1:1:1 of 0.393 mm, 0.724 mm and 0.960 mm particles

Solid feed rate = 180 g/min.

S.	h_D = 0.5			h_D = 1.0		
	u/u_{mf}	H(2,1)	H(3,2)	u/u_{mf}	H(2,1)	H(3,2)
1.	1.813	1.16	1.32	1.806	1.23	1.14
2.	1.989	1.35	1.48	2.046	1.31	1.37
3.	2.192	1.52	1.56	2.213	1.37	1.50
4.	2.422	1.47	1.49	2.447	1.49	1.52
5.	2.623	1.44	1.39	2.663	1.43	1.67
6.	2.829	1.40	1.35	2.854	1.39	1.57
7.	3.074	1.38	1.33	3.112	1.27	1.46
8.	3.328	1.38	1.30	3.381	1.20	1.39
9.	3.866	1.32	1.23	3.902	1.17	1.36
10.	4.154	1.20	1.15	4.170	1.15	1.34

TABLE H.3

SOLID FEED = 60.0 G/MIN EACH OF 0.393MM, 0.724MM AND 0.96MM BALLOTINIS
SOLID FEED RATE = 180.0 G/MIN
BED HOLD UP = 90.0 GRAM
STATIC BED HEIGHT= 5.0CM

AVG. d_p MM	τ_S MIN	T_{GA} MIN	T_{GB} MIN	τ_A	τ_B	U/U_{mf}	H(2,1)	H(3,1)	H(3,2)
.613	.500	9.258E-04	4.465E-04	540.05	1119.60	1.780	1.034	1.068	1.033
.639	.500	9.008E-04	3.827E-04	555.05	1306.20	1.926	1.159	1.304	1.125
.650	.500	8.820E-04	3.349E-04	566.86	1492.80	2.134	1.254	1.583	1.261
.662	.500	8.306E-04	2.965E-04	601.92	1686.31	2.334	1.349	1.868	1.385
.667	.500	8.107E-04	2.679E-04	616.72	1866.00	2.551	1.357	1.961	1.444
.674	.500	7.942E-04	2.444E-04	629.48	2045.68	2.741	1.310	2.066	1.576
.678	.500	7.650E-04	2.238E-04	653.52	2233.67	2.962	1.424	2.175	1.527
.679	.500	7.528E-04	2.067E-04	664.16	2418.88	3.197	1.408	2.168	1.539
.680	.500	6.891E-04	1.808E-04	725.56	2764.44	3.644	1.400	2.187	1.562
.677	.500	6.976E-04	1.682E-04	716.73	2971.77	3.948	1.401	2.198	1.568

TABLE H.4

SOLID FEED = 60.0 G/MIN EACH OF 0.393MM, 0.724MM AND 0.96MM BALLOTINIS
SOLID FEED RATE = 180.0 G/MIN
BED HOLD UP = 160.3 GRAM
STATIC BED HEIGHT= 10.0CM

AVG. d_p MM	τ_S MIN	T_{GA} MIN	T_{GB} MIN	τ_A	τ_B	U/U_{mf}	H(2,1)	H(3,1)	H(3,2)
.633	.888	9.258E-04	7.939E-04	960.09	1119.60	1.683	1.260	1.338	1.061
.650	.888	9.008E-04	6.805E-04	986.76	1306.20	1.868	1.350	1.630	1.206
.666	.888	8.820E-04	5.954E-04	1007.75	1492.80	2.045	1.399	1.894	1.354
.678	.888	8.306E-04	5.271E-04	1070.09	1686.31	2.236	1.460	2.194	1.501
.678	.888	8.107E-04	4.763E-04	1096.40	1866.00	2.475	1.465	2.280	1.556
.694	.888	7.942E-04	4.345E-04	1119.08	2045.68	2.600	1.478	2.418	1.635
.697	.888	7.650E-04	3.979E-04	1161.82	2233.67	2.812	1.492	2.557	1.714
.690	.888	7.528E-04	3.674E-04	1180.74	2418.88	3.106	1.429	2.454	1.716
.695	.888	6.891E-04	3.215E-04	1289.88	2764.44	3.504	1.462	2.537	1.734
.692	.888	6.976E-04	2.991E-04	1274.19	2971.77	3.798	1.454	2.543	1.748

TABLE H.5

SOLID FEED = 60.0 G/MIN EACH OF 0.393MM, 0.724MM AND 0.96MM BALLOTINIS
SOLID FEED RATE = 180.0 G/MIN
BED HOLD UP = 320.3 GRAM
STATIC BED HEIGHT= 2.5CM

d_P MM	τ_S MIN	TGA MIN	TGB MIN	τ_A	τ_B	U/U_{mf}	H(2,1)	H(3,1)	H(3,2)
.642	1.777	9.258E-04	1.587E-03	1920.18	1119.60	1.638	1.321	1.455	1.101
.664	1.777	9.008E-04	1.361E-03	1973.52	1306.20	1.797	1.450	1.815	1.252
.682	1.777	8.820E-04	1.190E-03	2015.51	1492.80	1.956	1.555	2.196	1.411
.703	1.777	8.306E-04	1.054E-03	2140.18	1686.31	2.095	1.699	2.711	1.595
.708	1.777	8.107E-04	9.527E-04	2192.80	1865.99	2.284	1.754	2.980	1.698
.719	1.777	7.942E-04	8.690E-04	2238.17	2045.68	2.439	1.796	3.150	1.753
.724	1.777	7.650E-04	7.958E-04	2323.65	2233.67	2.628	1.833	3.327	1.814
.719	1.777	7.528E-04	7.349E-04	2361.48	2418.88	2.879	1.743	3.227	1.851
.716	1.777	6.891E-04	6.430E-04	2579.77	2764.44	3.320	1.655	3.064	1.851
.710	1.777	6.976E-04	5.982E-04	2548.39	2971.77	3.625	1.581	2.981	1.885

TABLE H.6

SOLID FEED = 100.0 G/MIN EACH OF 0.393MM, 0.724MM AND 0.96MM BALLOTINIS
SOLID FEED RATE = 300.0 G/MIN
BED HOLD UP = 90.0 GRAM
STATIC BED HEIGHT= 5.0CM

AVG. d_p MM	τ_S MIN*	τ_{GA} MIN	τ_{GB} MIN	τ_A	τ_B	U/U_{mf}	H(2,1)	H(3,1)	H(3,2)
.608	.300	9.258E-04	4.465E-04	324.03	671.76	1.811	1.033	1.052	1.018
.627	.300	9.008E-04	3.827E-04	333.03	783.72	1.998	1.119	1.269	1.134
.641	.300	8.820E-04	3.349E-04	340.11	895.68	2.193	1.192	1.445	1.212
.650	.300	8.306E-04	2.965E-04	361.15	1011.78	2.415	1.245	1.650	1.324
.656	.300	8.107E-04	2.679E-04	370.03	1119.60	2.623	1.281	1.709	1.333
.662	.300	7.942E-04	2.444E-04	377.69	1227.41	2.831	1.295	1.778	1.372
.664	.300	7.650E-04	2.238E-04	392.11	1340.20	3.078	1.310	1.817	1.386
.663	.300	7.528E-04	2.067E-04	398.50	1451.33	3.336	1.302	1.851	1.421
.660	.300	6.891E-04	1.808E-04	435.33	1658.66	3.852	1.106	1.869	1.690
.659	.300	6.976E-04	1.682E-04	430.04	1783.06	4.146	1.287	1.822	1.415

TABLE H.7

SOLID FEED = 166.7 G/MIN EACH OF 0.393MM, 0.724MM AND 0.96MM BALLOTINIS
SOLID FEED RATE = 500.0 G/MIN
BED HOLD UP = 90.0 GRAM
STATIC BED HEIGHT= 5.0CM

AVG. d_P MM	τ_S MIN	τ_{GA} MIN	τ_{GB} MIN	τ_A	τ_B	U/U_{mf}	H(2,1)	H(3,1)	H(3,2)
.611	.180	9.258E-04	4.465E-04	194.41	403.05	1.794	1.016	1.049	1.032
.622	.180	9.008E-04	3.827E-04	199.81	470.23	2.024	1.087	1.200	1.103
.628	.180	8.820E-04	3.349E-04	204.07	537.40	2.277	1.134	1.311	1.155
.638	.180	8.306E-04	2.965E-04	216.69	607.07	2.496	1.163	1.399	1.203
.644	.180	8.107E-04	2.679E-04	222.02	671.76	2.719	1.200	1.513	1.261
.647	.180	7.942E-04	2.444E-04	226.61	736.44	2.949	1.195	1.528	1.278
.644	.180	7.650E-04	2.238E-04	235.27	804.12	3.254	1.197	1.549	1.293
.650	.180	7.528E-04	2.067E-04	239.10	870.80	3.459	1.203	1.570	1.305
.646	.180	6.891E-04	1.808E-04	261.20	995.20	4.003	1.187	1.583	1.333
.642	.180	6.976E-04	1.682E-04	258.02	1069.84	4.354	1.180	1.573	1.333

TABLE H.9

SOLID FEED = 166.7 G/MIN EACH OF 0.393MM, 0.724MM AND 0.96MM BALLO39N9
SOLID FEED RATE = 500.0 G/MIN
BED HOLD UP = 160.3 GRAM
STATIC BED HEIGHT= 10.0CM

AVG. d_P MM	τ_S MIN	T_{GA} MIN	T_{GB} MIN	τ_A	τ_B	U/U_{mf}	H(2,1)	H(3,1)	H(3,2)
.617	.533	9.258E-04	7.939E-04	576.05	671.76	1.759	1.121	1.172	1.046
.631	.533	9.008E-04	6.805E-04	592.05	783.72	1.970	1.198	1.400	1.168
.639	.533	8.820E-04	5.954E-04	604.65	895.68	2.204	1.271	1.543	1.214
.654	.533	8.306E-04	5.271E-04	642.05	1011.78	2.385	1.310	1.664	1.270
.660	.533	8.107E-04	4.763E-04	657.84	1119.60	2.594	1.351	1.780	1.317
.666	.533	7.942E-04	4.345E-04	671.45	1227.41	2.802	1.369	1.852	1.352
.668	.533	7.650E-04	3.979E-04	697.09	1340.20	3.041	1.381	1.922	1.391
.666	.533	7.528E-04	3.674E-04	708.44	1451.33	3.309	1.375	1.932	1.405
.666	.533	6.891E-04	3.215E-04	773.93	1658.66	3.782	1.356	1.931	1.424
.665	.533	6.976E-04	2.991E-04	764.51	1783.06	4.080	1.338	1.902	1.422

TABLE H.10

SOLID FEED = 100.0 G/MIN EACH OF 0.393MM, 0.724MM AND 0.96MM B+LLO
SOLID FEED RATE = 300.0 G/MIN
BED HOLD UP = 320.3 GRAM
STATIC BED HEIGHT= 2.5CM

d_p MM	τ_S MIN	τ_{GA} MIN	τ_{GB} MIN	τ_A	τ_B	U/U_{mf}	H(2,1)	H(3,1)	H(3,2)
.625	.640	9.258E-04	1.587E-03	691.26	403.05	1.721	1.179	1.241	1.052
.634	.640	9.008E-04	1.361E-03	710.46	470.23	1.953	1.261	1.289	1.021
.662	.640	8.820E-04	1.190E-03	725.58	537.40	2.068	1.329	1.768	1.330
.660	.640	8.306E-04	1.054E-03	770.46	607.07	2.345	1.380	1.814	1.314
.670	.640	8.107E-04	9.527E-04	789.41	671.76	2.529	1.397	1.966	1.406
.676	.640	7.942E-04	8.690E-04	805.74	736.44	2.728	1.432	2.076	1.449
.678	.640	7.650E-04	7.958E-04	836.51	804.12	2.961	1.447	2.226	1.537
.681	.640	7.528E-04	7.349E-04	850.13	870.80	3.181	1.450	2.176	1.500
.678	.640	6.891E-04	6.430E-04	928.71	995.20	3.661	1.433	2.173	1.516
.680	.640	6.976E-04	5.982E-04	917.42	1069.84	3.919	1.418	2.153	1.518

TABLE H.11

SOLID FEED = 166.7 G/MIN EACH OF 0.393MM, 0.724MM AND 0.96MM BALLO39N9
SOLID FEED RATE = 500.0 G/MIN
BED HOLD UP = 320.3 GRAM
STATIC BED HEIGHT= 2.5CM

AVG. d_P MM	τ_S MIN	T_{GA} MIN	T_{GB} MIN	τ_A	τ_B	U/U_{mf}	H(2,1)	H(3,1)	H(3,2)
.630	1.066	9.258E-04	1.587E-03	1152.11	671.75	1.694	1.258	1.331	1.058
.643	1.066	9.008E-04	1.361E-03	1184.11	783.72	1.907	1.351	1.460	1.080
.661	1.066	8.820E-04	1.190E-03	1209.31	895.68	2.073	1.423	1.849	1.299
.672	1.066	8.306E-04	1.054E-03	1284.10	1011.78	2.269	1.488	2.112	1.419
.682	1.066	8.107E-04	9.527E-04	1315.68	1119.59	2.450	1.529	2.315	1.513
.684	1.066	7.942E-04	8.690E-04	1342.90	1227.41	2.671	1.459	2.283	1.565
.687	1.066	7.650E-04	7.958E-04	1394.19	1340.20	2.887	1.480	2.379	1.606
.688	1.066	7.528E-04	7.349E-04	1416.89	1451.33	3.121	1.478	2.408	1.629
.689	1.066	6.891E-04	6.430E-04	1547.86	1658.66	3.563	1.457	2.387	1.638
.688	1.066	6.976E-04	5.982E-04	1529.03	1783.06	3.841	1.420	2.359	1.660

APPENDIX I

COMPUTER PROGRAM FOR PARAMETER ESTIMATION BY THE METHOD OF WEIGHTED MOMENTS

PROGRAM I.1 COMBINATION OF FIRST AND SECOND MOMENTS

```
//DKB      JOB      ,RU209J34,05,15,0
/*
/+
//DKB     EXEC FORTRAN)BCD,MAP*
C      RESEARCH WORK D K BHARADWAJ
C
C      PARAMETER CALCULATION BY WEIGHTED MOMENTS USING COMBINATION OF
C      MOMENTS
C
       DIMENSION C1(50),C2(50)
       READ 135, NN
   135 FORMAT (I5)
       DO 1000 L = 1,NN
       READ 10,N1,N2,T1,T2,DT,H
    10 FORMAT (2I5,4F10.3)
C
C      READING RESPONSE CURVES
C
       READ 11, (C1(I), I = 1,N1),(C2(I), I = 1,N2)
    11 FORMAT(8E10.4)
       AREA1 = TRANS(0.0,C1,T1,DT,N1,0,0.0)
       AREA2 = TRANS (0.0,C2,T2,DT,N2,0,0.0)
       PRINT 5,AREA1,AREA2
     5 FORMAT (///22H  AREA UNDER CURVE 1 =E12.4,11H  CURVE 2 =E12.4///)
C
C      NORMALIZING RESPONSE CURVES
C
       DO 20 I = 1,N1
    20 C1(I) = C1(I)/AREA1
       DO 21 I = 1,N2
    21 C2(I) = C2(I)/AREA2
       PRINT 65
    65 FORMAT (26H THE NORMALIZED CURVES ARE/)
       PRINT 111, (C1(I), I = 1,N1), (C2(I), I = 1,N2)
   111 FORMAT(8E12.4)
C
C      T IS TIME IN SECONDS AND DT IS TIME INTERVAL
C
C      CALCULATION OF TRANSFORMS AT DIFFERENT VALUES OF S
C
       READ 1, KPS
     1 FORMAT (I5)
       PRINT 1,KPS
       EX = 2.0/3.0
       DO 1000 K = 1,KPS
       READ 12, S
    12 FORMAT (F10.3)
```

```
      PRINT 12,S
C
C     ZEROTH MOMENTS
C
      TZ1    = TRANS(S,C1,T1,DT,N1,0,0.0)
      TZ2    = TRANS(S,C2,T2,DT,N2,0,0.0)
C
C     FIRST MOMENTS
C
      TF1 = TRANS(S,C1,T1,DT,N1,1,0.0)
      TF2 = TRANS(S,C2,T2,DT,N2,1,0.0)
C
C     SECOND MOMENTS
C
      TF1 = TF1/TZ1
      TF2 = TF2/TZ2
      TS1 = TRANS(S,C1,T1,DT,N1,2,TF1)
      TS2 = TRANS(S,C2,T2,DT,N2,2,TF2)
      TS1 = TS1/TZ1
      TS2 = TS2/TZ2
      Z = ALOG(TZ2/TZ1)
      B = TS2 - TS1
      A = TF2 - TF1
   50 CONTINUE
C
C     PARAMETER CALCULATION FROM ZEROETH AND FIRST MOMENTS.
C
      TAU = -1.0/(1.0/A + 2.0*S/Z)
      DN = (1.0-TAU/A)/2.0/Z
      PRINT 51
   51 FORMAT(8X,52HPARAMETERS CALCULATED FROM ZEROETH AND FIRST MOMENTS/
     114X,3HTAU 9X,2HDN,1HU 11X,1HD)
      U = H/TAU
      D = U*H*DN
      PRINT 60,TAU,DN,U,D
   60 FORMAT(8X,F10.4,4X,F8.6,4X,F7.4,F10.4)
C
C     PARAMETER CALCULATION FROM FIRST AND SECOND MOMENTS
C
      TAU = 1.0/SQRT(1.0/A**2-2.0*S*B/A**3)
      DN  = B*TAU/2.0/A**3
      U = H/TAU
      D = U*H*DN
      PRINT 52
   52 FORMAT(8X,51HPARAMETERS CALCULATED FROM FIRST AND SECOND MOMENTS/
     114X,3HTAU 9X,2HDN 10X,1HU 11X,1HD)
      PRINT 60,TAU,DN,U,D
 1000 CONTINUE
      STOP
```

```
      END
      FUNCTION TRANS(S,C,TIME,DT,N,NX,TS)
      DIMENSION Y(50), C(50)
C
C     TRANS IS A LAPLACE TRANSFORM
C
      T = TIME
      IF (T) 3,3,4
    3 Y(1) = 0.0
      J = 2
      T = DT
      GO TO 5
    4 J = 1
    5 DO 1 I = J,N
      P = S*T
      IF (ABS(P) - 100.0)13,14,14
   14 EP = 0.0
      GO TO 15
   13 EP = EXP(-P)
   15 Y(I) = C(I)*EP*(T-TS)**NX
    1 T = T+DT
C
C     SIMPSONS RULE
C
      TRANS = Y(1) - Y(N)
      N1 = N-1
      DO 2 I = 2,N1,2
    2 TRANS = TRANS + 4.0*Y(I) + 2.0*Y(I+1)
      TRANS = TRANS*DT/3.0
      RETURN
      END
```

PROGRAM I.2 SINGLE MOMENT AT TWO S VALUES

```
//DKB    EXEC FORTRAN)BCD,MAP*
C      RESEARCH WORK D K BHARADWAJ
C      PARAMETER ESTIMATION BY WEIGHTED MOMENTS METHOD
C
C      MAIN PROGRAMMM
       DIMENSION C1(50),C2(50),S(20),TZ1(30),TZ2(30),TF1(30),TF2(30),
      1TS1(30),TS2(30),A(20),B(20),FZ(20)
       READ 135, NN
   135 FORMAT (I5)
       DO 1000 L = 1,NN
       READ 10,N1,N2,T1,T2,DT,H
    10 FORMAT (2I5,4F10.3)
C
C      READING RESPONSE CURVES
C
       READ 11, (C1(I), I = 1,N1),(C2(I), I = 1,N2)
    11 FORMAT(8E10.4)
       AREA1 = TRANS(0.0,C1,T1,DT,N1,0,0.0)
       AREA2 = TRANS (0.0,C2,T2,DT,N2,0,0.0)
       PRINT  5,AREA1,AREA2
     5 FORMAT (///22H  AREA UNDER CURVE 1 =E12.4,11H  CURVE 2 =E12.4///)
C
C      NORMALIZING RESPONSE CURVES
C
       DO 20 I = 1,N1
    20 C1(I) = C1(I)/AREA1
       DO 21 I = 1,N2
    21 C2(I) = C2(I)/AREA2
       PRINT 65
    65 FORMAT (26H THE NORMALIZED CURVES ARE/)
       PRINT 111, (C1(I), I = 1,N1), (C2(I), I = 1,N2)
   111 FORMAT(8E12.4)
C
C      T IS TIME IN SECONDS AND DT IS TIME INTERVAL
C
C      CALCULATION OF TRANSFORMS AT DIFFERENT VALUES OF S
C
       READ 1, KPS,NS
     1 FORMAT (2I5)
       PRINT 1, KPS,NS
       EX = 2.0/3.0
       DO 1000 K = 1,KPS
       READ 12,(S(I), I = 1,NS)
```

```
   12 FORMAT (8F10.3)
      PRINT 12,(S(I), I = 1,NS)
      DO 50 I = 1,NS
      P = S(I)
C     ZEROTH MOMENTS
C
      TZ1(I) = TRANS(P,C1,T1,DT,N1,0,0.0)
      TZ2(I) = TRANS(P,C2,T2,DT,N2,0,0.0)
C
C     FIRST MOMENTS
C
      TF1(I) = TRANS(P,C1,T1,DT,N1,1,0.0)
      TF2(I) = TRANS(P,C2,T2,DT,N2,1,0.0)
C
C     SECOND MOMENTS
C
      TF1(I) = TF1(I)/TZ1(I)
      TF2(I) = TF2(I)/TZ2(I)
      TS = TF1(I)
      TS1(I) = TRANS(P,C1,T1,DT,N1,2,TS)
      TS = TF2(I)
      TS2(I) = TRANS(P,C2,T2,DT,N2,2,TS)
      TS1(I) = TS1(I)/TZ1(I)
      TS2(I) = TS2(I)/TZ2(I)
      FZ(I) = TZ2(I)/TZ1(I)
      B(I) = TS2(I) - TS1(I)
   50 A(I) = TF2(I) - TF1(I)
C
C
C     PARAMETER CALCULATION FROM ZEROETH MOMENTS AT TWO VALUES OF S
C
      PRINT 55
   55 FORMAT (8X,57HPARAMETERS CALCULATED FROM ZEROETH MOMENT AT TWO S V
     1ALUES/14X,3HTAU 9X,2HDN 10X,1HU 11X,1HD)
      DN=(S(1)*ALOG(FZ(2))-S(2)*ALOG(FZ(1)))/((S(1)*(ALOG(FZ(2)))**2
     1-S(2)*(ALOG(FZ(1)))**2))
      TAU=((1.0-2.0*ALOG(FZ(1))*DN)**2-1.0)/(4.0*S(1)*DN)
      U = H/TAU
      D = U*H*DN
      PRINT 60,TAU,DN,U,D
   60 FORMAT (8X,F10.4,4X,F8.6,4X,F7.4,F10.4)
C
C     PARAMETER CALCULATION FROM FIRST MOMENTS AT TWO VALUES OF S
C
      PRINT 51
   51 FORMAT(8X,55HPARAMETERS CALCULATED FROM FIRST MOMENT AT TWO S VALU
     1ES/14X,3HTAU 9X,2HDN 10X,1HU 11X,1HD)
      TAU=A(1)*A(2)*SQRT((S(2)-S(1))/(A(2)**2*S(2)-A(1)**2*S(1)))
```

```
      DN=((TAU/A(1))**2-1.0)/(4.0*S(1)*TAU)
      U=H/TAU
      D= U*H*DN
      PRINT 60,TAU,DN,U,D
C
C     PARAMETER CALCULATION FROM SECOND MOMENTS
C
      PRINT 61
   61 FORMAT(8X,56HPARAMETERS CALCULATED FROM SECOND MOMENT AT TWO S VAL
     1UES/14X,3HTAU 9X,2HDN 10X,1HU 11X,1HD)
      TAU=2.0*SQRT((S(2)-S(1))**3/(S(2)/B(1)**EX-S(1)/B(2)**EX))/(1.0/B
     1(2)**EX-1.0/B(1)**EX)
      DN=(1.0/B(2)**EX-1.0/B(1)**EX)/(4.0*TAU*(S(2)/B(1)**EX-S(1)/B(2)**
     1EX))
      U=H/TAU
      D = U*H*DN
      PRINT 60,TAU,DN,U,D
 1000 CONTINUE
      STOP
      END
      FUNCTION TRANS(S,C,TIME,DT,N,NX,TS)
      DIMENSION Y(50) , C(50)
C
C     TRANS IS A LAPLACE TRANSFORM
C
      T = TIME
      IF (T) 3,3,4
    3 Y(1) = 0.0
      J = 2
      T = DT
      GO TO 5
    4 J = 1
    5 DO 1 I = J,N
      P = S*T
      IF (ABS(P)  - 100.0) 13,14,14
   14 EP = 0.0
      GO TO 15
   13 EP = EXP(-P)
   15 Y(I) = C(I)*EP*(T-TS)**NX
    1 T = T+DT
C
C     SIMPSONS RULE
C
      TRANS = (Y(1)-Y(N))
      N1 = N-1
      DO 2 I = 2,N1,2
    2 TRANS = TRANS + 4.0*Y(I) + 2.0*Y(I+1)
      TRANS = TRANS*DT/3.0
      RETURN
      END
```

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