STUDIES IN MULTISTAGE COUNTER CURRENT FLUIDIZATION

A Dissertation
submitted in partial fulfilment
of the requirements for the award of the degree
of
MASTER OF ENGINEERING
in
PLANT AND EQUIPMENT DESIGN

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CERTIFICATE

Certified that the thesis entitled "STUDIES IN MULTISTAGE COUNTER CURRENT FLUIDIZATION" which is submitted by Sri Ashok Kumar Gupta in partial fulfilment of the requirements for the award of the degree of MASTER OF ENGINEERING IN CHEMICAL ENGINEERING (Equipment & Plant Design) University of Roorkee is a record of 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 award of any other degree.

This is further certified that he has worked for a period of about eight months for preparing this thesis.

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ABSTRACT

Studies in Multistage Fluidization were carried out in a two stage counter current column using air as the fluidizing medium. The column consists of a [illegible] perspex tube of 50 mm internal diameter and each stage is 225 mm long. Down comers made of copper tube of 10 mm I.D. x 230 mm long have been provided on the grid plate of each stage. The solids are fed through the funnels of different discharge openings to the top stage and solids overflow to the bottom stage through the downcomer. The solids from the bottom stage overflow through the downcomer to a receiver kept below the column.

It was found that the solids flow rate increases with air flow rate upto a maximum and then start decreasing till it get entrained. Solid loading ratio, that is the weight ratio of solids to air per unit time will effect the solids downflow. Under limited conditions of operation, the smooth flow region has been identified for different flow rates of air and feed rates of solids. A corelation was found out under smooth zone of operation, which is :

$$\log(W_s/G_f) = 9.9 \log (N_{Re_p}) + C$$

Where, the value of constant C depends upon the density and the particle diameter and it was found for different materials of different sizes as indicated in the table below:

D_p mm	C		
	Calcite	Bauxite	Coal
0.648	-0.0415	-0.0426	-0.108
0.426	-0.0302	-0.01415	-0.170
0.287	-0.188	-0.1425	-

NOMENCLATURES

D_o	Discharge diameter of funnel	mm
D_p	Particle diameter	mm
g	Acceleration due to gravity	cm/sec^2
G_f	Gas flow rate	kgm/M^2-hr
R	Solid loading ratio W_s/G_f	
N_{Re_p}	Reynold number, $D_p G_f/\mu$	
W_s	Solids flow rate	kgm/M^2-hr
ρ_s	Solids density	gm/cc
ρ_f	Air density	gm/cc
ϵ	Void fraction of bed	
μ	Viscosity of air	gm/cm-sec.

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CHAPTER - I

INTRODUCTION :

Fluidization is a technique for contacting solids and fluids and finds wide application in Chemical, Metallurgical and Petroleum industries. In a fluidized bed of solids due to intimate contact between solids and fluid, better heat and mass transfer rates are obtained.

A solid bed is considered to be fluidized when the fluid velocity is such that the solids are suspended. Usually in Gas-Solid systems at low gas velocities the bed is static and is termed as fixed. In this zone an increase in gas velocity leads to an increase in bed pressure drop, accompanied by slight expansion of the bed. An increase in velocity of gas beyond the fixed bed zone leads to loosening of bed and the solids get~~s~~ suspended. This is the onset of fluidization. At the onset the total pressure drop is nearly equal to the total weight of solids. Increase in gas velocity beyond minimum fluidization velocity increases the bed height, increases the particle movement but the bed pressure drop remains essentially same as at minimum fluidization condition.

A dense phase gas fluidized bed looks very much like a boiling liquid and in many ways exhibits liquid like behaviour. For example, a large, light object is easily pushed into a bed, and on release it will pop up and float on the surface. The upper surface of the bed remains horizontal when the container is tipped, and when two beds are connected their levels equilize. Also, the difference in pressure between any two points in a bed is roughly approximated by the static head of bed between these points. The bed also has liquid like flow properties. Solids will gush in a jet from a hole in the side of a container and they can be made to flow like a liquid from vessel to vessel.

If the flow rate of fluid is increased above the minimum required to produce a fluidized bed, one of two things will occur; either the bed will continue to expand so that the average distance between the particles become greater, or the excess fluid will pass through the bed in the form of bubbles, giving rise essentially to a two-phase system. These two types of fluidization are referred to as being respectively "Particulate" and "Aggregative". In general particulate fluidization occurs with liquid solid systems. Aggregative fluidization occurs with all other gas-solid systems.

Harrison et al (1) have suggested that the type of fluidization obtained will be related to the maximum stable size of bubble which can exist within the fluidized bed. If the circulation velocity of gas within the bubble (which is normally approximately equal to the bubble rise velocity) exceeds the terminal falling velocity of the particles, particles will be sucked in at the rear of the bubble which will thus tend be destroyed. Since the bubble rise velocity increases with bubble volume and is substantially independent of the properties of the fluidized bed, the maximum stable bubble size increases with the terminal falling velocity of the particles in the fluid. If this size exceeds say, ten times the particle diameter, the bubble will be obvious and the fluidization will be aggregative, but if it is of same order as the particle diameter, it can no longer be distinguished from the interstitial gas and the fluidization will appear to be particulate. On this basis most gas-solid systems will give rise to aggregative fluidization, and most liquid-solid systems will give rise to particulate fluidization.

Wilhelm and Kwauk (2) have suggested using the Froude group (U_{mf}^2/gd) as a criterion for the type of fluidization obtained; in general aggregative fluidization is obtained at values above unity and particulate fluidization at values below unit. From a consideration of the stability equations for bed-fluid interface originally given by Rice and Wilhelm (3), Romero and Johnson (4) suggested four dimensionless groups to characterise the quality of fluidization.

$$(F_{r,mf})\,(R_{ep,mf})\left(\frac{\rho_s-\rho_f}{\rho_f}\right)\left(\frac{L_{mf}}{d_t}\right) \begin{cases} < 100, & \text{smooth or particulate} \\ > 100, & \text{bubbling or aggregative.} \end{cases}$$

Slug flow will often occur in smaller reactors, because bubbles in a gas-fluidized bed rapidly coalesce above the distributor plate. When the bubble diameter approaches the column diameter, the portion of the bed above the bubble is pushed upward in a piston, like manner, and the fluidized bed is said to be 'slugging'. Particles rain down from the slug and it finally disintegrates. At about this time another slug forms and this unstable oscillatory motion is repeated. Slugging is usually undesirable since it increases the problems of entrainment and lowers the performance potential of the bed for both physical and chemical operations. Slugging is serious in long, narrow fluidized beds.

1.1 Influence of bed support and fluid distributor :

The fluid distributor at the base of the bed can exert a very strong influence on the structure of the whole bed. Ideally it should be porous in structure with a fine texture so that the fluid is introduced through a multitude of small injection points. A coarse distributor gives excessively high injection rates at localized points and increases the tendency for channelling within the bed.

If the bed does not channel readily, more even fluidization is obtained using a distributor with a high resistance to gas flow since the fluid is then introduced fairly evenly to the lower portion of the bed, irrespective of any irregularities in the bed itself. Distribution will be uniform if the pressure drop across the distributor is sufficiently high, generally atleast equal to the pressure drop through the bed. For a few air inlet openings the bed density fluctuates appreciably at all flow rates and varies with height and channelling may be severe. For many air inlet openings the fluctuations in bed density is negligible at low flow rates but again becomes appreciable at high flow rates.

Hiby (5) points out that if a channel forms in a fluidized bed it will offer a low resistance path for the flow of fluid. Furthermore, there will be a tendency for the channel to open up progressively as the velocity is increased. The resistance to flow of the section of the distributor from which fluid enters the channel will, however, increase with flow rate. In order that the channel should not be stabilized, the decrease in pressure drop with flow rate across the channel should be at least compensated by the increase in pressure drop across the section of distributor.

In large beds it is clearly impracticable to use porous distributors, first because of the expense and secondly because of their poor mechanical strength, where good distribution of gas is essential, either a series of nozzles or of bubble caps is frequently used, with the aperture of each capable of adjustment in order to give good uniformity. The minimum fluidizing velocity is found most consistant using a porous plate distributor and the bed expension is then even.

1.2 Initiation of fluidization and minimum fluidizing velocity:

When a fluid at small flow rates is allowed to pass upward through a bed of solids, the fluid enters the void space between the particles and moves upwards, with liquid-solid systems, there are not normally any serious difficulties in initiating fluidization and once the system has been fluidized uniform conditions are usually obtained. With gas-solid systems there is very marked contrast in the behaviour of solid materials, some of which will fluidize extremely readily and others of which are virtually incapable of being fluidized. In general the following properties of the solid and fluid are conducive to well-fluidized systems:

(a) Low particle density.
(b) Small particle size.
(c) Particle shape approaching spherical.
(d) High density fluid.

Unfortunately, those propertie-s of solid which make it capable of giving a well-fluidized system are also those which tend to make the initiation of fluidization difficult. If surface forces between the particles are significant, they will be greater with small particles because of their high specific surface. If the particles are of low density, the gravitational forces tending to pull then apart will be small. Thus low density particles can give rise to serious channelling even though they may fluidize well once they have been brought into suspension.

At low velocities of flow through a bed, the particles maintain a fixed orientation with respect to one another. However, there is evidence that at velocity considerably below the minimum fluidizing velocity, some movement of the loosely held particles take place. As the fluid velocity is increased fluidization of small localized pockets probably occurs. Then, over a fairly small velocity range, most of the bed becomes fluidized, but it may often contain small pockets of unfluidized solids at velocities considerably in excess of those at which the bed gives the appearance of being well fluidized.

In an idealized system, the minimum fluidizing velocity is that velocity at which the bed suddenly changes from a fixed to a fluidized state. At this point the frictional force between a particle and fluid counter balances the weight of the particles, the vertical component of the compressive force between adjecent particles disappeares and pressure drop through any section of the bed about equals the weight of fluid and particles in that section. In practice, however, there may be a large transitional region and the minimum fluidizing velocity will then have no absolute ~~dkm~~ significance.

The minimum fluidizing velocity can be predicted using a number of different correlations given by different workers. Some of which are given below:

$$G_{mf} = \frac{688\ D_p^{1.82}\left[\rho_f(\rho_s-\rho_f)\right]^{0.94}}{\mu^{0.18}}$$

where D_p is in inches, ρ_s, ρ_f is in lbm/ft^3 and μ is in c.p. and G_{mf} is in lbm/ft^2.hr.

$$U_{mf} = \frac{0.00059\ d^2(\rho_s-\rho_f)g}{\mu}$$

Here, D_p is in cm, g the acceleration due to gravity is in cm/sec^2, μ is in gm/cm-sec and ρ_s, ρ_f are in gm/cc, and U_{mf} is in cm/sec.

1.3 Types of fluidization operations:

Fluidized beds are operated as batch or continuous systems. In batch fluidized beds the particles are out of contact with each other and the particle position is defined only as a function of time. The batch fluidizer is a perfect example of backmixing and the properties at any point in the bed are same. In a batch process the only drawback is that it cannot be used for large scale operation.

1.3.1 Continuous single stage fluidization:

Batch fluidizers can be adopted for continuous operation by continually feeding solid material from one end and discharging from the other end with proper control in its hold-up time or time of contacting. Fluidized bed being an effective solid mixer, the feed, as soon as it enters the fluidizer, gets distributed. The characteristics of the fluidized bed are essentially independent of feed condition. The feed as soon as it enters the bed, is instantaneously distributed in the bed. Hence there is a possibility of a portion of the feed coming out of the discharge immediately after entering the bed. To avoid this short circuiting, residence time has to be increased. This is achieved by increasing bed heights or holding in fluid-solid contacting systems.

When a fluidized bed is used it is necessary to compare its advantage and disadvantage of which the most important are:

- Catalyst attrition and entrainment of catalyst dust with the gas stream.
- Reduction of average concentration of the reacting gases, because of intermixing of the gas over the height of the bed.
- The possibility of gas passing in the form of large bubbles, which leads to a decrease in the degree of conversion.
- In some cases the use of a fluidized bed leads to an increase in side-reactions, especially in case of homogeneous reactions.
- The rapid mixing of solid particles leads to uneven residence time of solids in the reactor, resulting in a product quality which is not uniform.

However the following advantages should be noted:

- The system allows for continuous automatically controlled operations with ease of solid handling for large scale operation.

- The rapid mixing of solids gives nearly isothermal operating conditions in the bed leading to easy control.
- Large quantities of heat addition or removal can be done with ease.
- Since mass and heat transfer rates between gas and solids are high, smaller equipments serve the purpose.
- Due to shorter time of contact, larger thermal gradients can be used without detrimental effect on product quality.
- It has no moving part.

The disadvantages of a fluidizer can be partly eliminated by various methods in which residence time distribution of solids and gases are narrowed to give uniform product. This is achieved in practice by multistage operation in which the fluidizer is devided into segments which act as individual units.

Due to increased contact time in multi-stage unit either with respect to gas or solid or both, higher efficiency is achieved in a transfer process or in a chemical reaction. Multi-stage units are operated as batch units or continuous units.

1.3.2 Multi-stage batch systems:

In these systems solids are handled in batches. The single stage may be devided into individual stages either fitted one over the other stage or, each stage is a seperate unit and kept on the same floor. In the first scheme solids are fed to each stage seperately and discharged seperately. Same gas is allowed to flow from one to other stages. Drawback is to feed solids as batches to each stage at different heights. This difficulty is removed by using second scheme, however large floor area is required per unit solids handled.

In batch systems high efficiency of performance is obtained with respect to gas. But the quality of product is different in each stage. Each stage being batch with respect to solid, large scale operations will be difficult as considerable time is lost in charging and discharging of solids.

1.3.3 Multi-stage continuous systems:

Continuous contacting of solids can be achieved by the following methods:

(i) Simple multiple contact system (fig. 1-a)
(ii) Cross-flow contact system (fig. 1-b)
(iii) Continuous counter current system (fig.1-c)

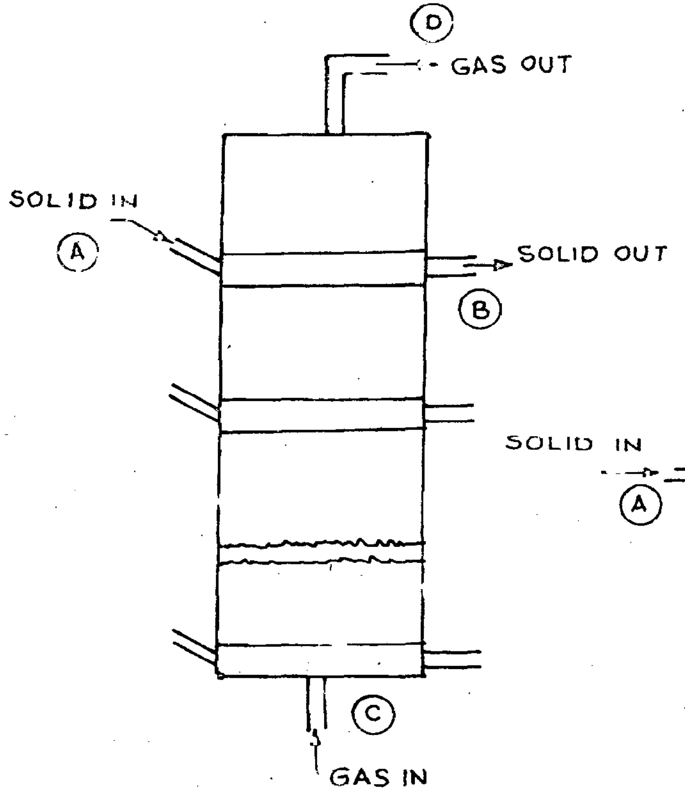

FIG.1a CONTINUOUS SIMPLE MULTIPLE CONTACTING UNIT.

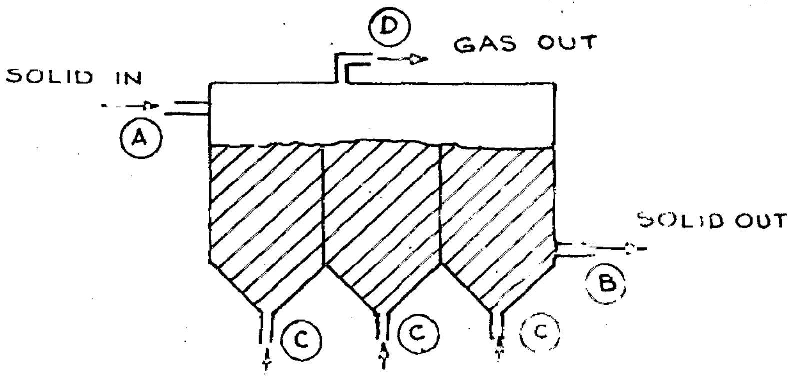

FIG1-b MULTI-STAGE CONTINUOUS CROSS FLOW UNIT.

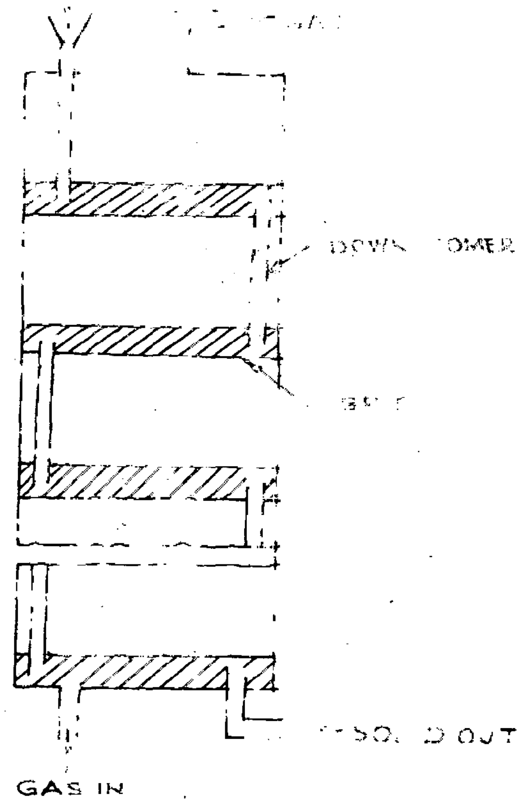

FIG1-c MULTI-STAGE COUNTER CURRENT CONTINUOUS UNIT

In these units solids are continuously fed at one end and product is taken out at the other end. In simple multiple contact system the solids are fed continuously to each stage. The product is similarly withdrawn continuously. The gas condition differs substantially from stage to stage because the same gas contacts the solids in different stages.

Better product distribution can be achieved by increasing residence time of solids as in cross flow system and counter-current systems. In the cross flow system, the solids are fed at point A. The solids flow from chamber to chamber and finally get discharged at point B. As regards gas, it is fed seperately to each chamber at point C and has only once through contact. The requirement of fluidizing medium will be large due to its multiple entry. Due to multiple contacting of solids, the residence time distribution is more uniform. This gives better solid product quality. As far as the gas is concerned, due to single once through contact, process has lower thermal and conversion efficiency.

The counter current multistage system shown in fig. 1-c is an improvement over the cross-flow system. Here the solids enter at the top at point A and move down from section to section through down comer and finally leaves at the ~~top~~ bottom. These units resemble distillation columns. Due to its multiple contacting the solids product is uniform. The residence time distribution of solids has a definite improvement over single stage unit. The gas requirement for fluidization are less than in cross flow systems. This gives better thermal and conversion efficiency for the gas.

Hence the use of multistage counter current contacting unit is advisable where solids are free flowing. An understanding of the behaviour of the multistage counter current system like fluid velocity, Pressure drop, slugging, residence time etc. will help in the careful selection and use of the equipment keeping in mind its limitations, like solid down flow. Three principal problems are encountered when attempts are made to multistage fluidized beds as follows:

(a) Solid recirculation :

Bursting of bubbles, causes the solids to be entrained. However it can be decreased by increasing free board, but with multistage beds there is a limit to the amount of free board that can be allowed.

(b) Distributor blockage :

Fine solids entrained in the gas stream may also give trouble by depositing in the distributor of an adjacent stage. This gradually increases the pressure drop across the distributor.

(c) Interstage solid transfer:

The solids are transferred from stage to stage using downcomers. This technique posses problems when a unit is started up and such devices may pass gas upward rather than solids down-ward.

Levison et al (6) had shown a number of arrangements to prevent short circuiting of air through the downcomer which enables the solids downflow. Rhombic type of baffles are installed on the lower ends of the downcomers. Downcomers with slurry inlet beveled at a 35° - 45° angle, do not require any special baffle. Contactors with this type of downcomers operates very satisfactorily and are stable under practically any conditions of countercurrent adsorption.

Multistage fluidization for contacting gases and solids seems to have received fairly wide attention of the industry. A glance at the practice of this technique shows that it can give better performance in operations like drying, absorption, heating and cooling of solids. Chemical reactions, like calcination, roasting, reduction and the like envolving complicated heat recovery systems. It can be better appreciated from a short summary provided in the following pages of the recent uses of multistage gas-solid contacting technique.

As has been mentioned above multistage counter current fluidization technique seems to have been accepted in industries. The practice seems to be way ahead of theory. There is very little information available regarding design of multistage fluidizers, especially in fixing the number of stages, apparatus dimensions or predicting performance and energy requirements. The present practice seems to be a recourse to expensive bench scale and pilot plant experimentation to arrive at suitable equipment size and operating conditions for specific application.

1.4 Applications of countercurrent multistage fluidization units

Operation with a staged counter-current flow plan improves the contacting conditions, eliminates channel break-throughs of unreacted components and stagnent zones in the contactor, and increases the operating reability of the system. The high reability and efficiency of sectionalized contactors show up clearly in severe forms of adsorptive seperation where the decisive factors becomes complete elimi-nationof channel break-throughs of feed-stock, which lead to off specfinished product. This phenomenon is not eliminated in shaft contactors but is avoided completely in adsorbers of the staged countercurrent type, owing to the repeated redistribution of streams of liquid feed stock during adsorbent contact by means of the sectionalizing perforated trays.

The advantage of sectionalized contactors are especially important as the capacity of adsorption unit is increased, since with high through-puts it is difficult to obtain uniform distribution of feed-stock and adsorbent and conditions arise that favour the formation of stagnent zones and the spontaneous appearance of channel formation with even the slight disruptions of hydrodynamic conditions that can hardly be avoided in practice.

In operation like absorption, single stage continuous units have low efficiency though the product is of uniform quality. In any operation total mass transfer is directly dependent on time of contact. Longer contact or residence time in single stage unit is obtained by deeper beds. But deeper beds have the inherent disadvantage of slugging. The Wide Residence Time Distribution of a single bed gives a non-uniform solid product and is inefficient for the high conversion of solids. The RTD can be narrowed and greatly improved by multistage operations. Here each stage acts almost like a theoritical stage. The gas and solids leaving each stage are nearby in equilibrium.

Multistage units provide a unique system of handling heat sensitive materials by providing different operating conditions at different stages. In practice multistage units are being used as pre-heating, reaction and cooling zones in the same unit. Thus the external material handling is avoided besides obtaining higher efficiencies.

Fluidization finds its wide applications in Drying Processes. Fluidized bed drying has been applied not only to grannular materials but also to pastes, suspensions, solutions and molten materials. As a result many batch processes have been replaced by continuous counter-current multistage processes which are much faster and economic.

Grain Dryer (7) :

The drier contains three sections. Two upper sections are used for removing moisture content and lower section is used for cooling grains. The closed rectangular chamber has following dimensions:

Cross section	2.5 x 1.25 m
Height	3.8 m
Grid slope	2 to 3°
Grid's round hole dia	1.4 mm
Grid area in heating zone	3.12 m^2
Grid area in cooling zone.	2.6 m^2

Material is delivered to first (top) chamber, it is gradually dried and passes to second chamber and then to the cooling zone whence it is discharged through a flap valve.

Sometimes mixers are used to ensure uniform fluidization and more through drying of very wet or finely devided materials.

Preheater for cement Rotary Kiln (8)(9) :

To efficiently recover heat of the exit gases from a Rotary Kiln for cement clinker. Russian Engineers employed counter-current heat exchange fresh cool solids with hot exit gases in a multistage fluidized bed. The operating data for the unit are:

Pallet feed rate	7 kg/hr-m^2
Superficial gas velocity	1 m/sec
Heat consumption	1100 Kcal/Kg. of Clinker
Cross section of the unit	2.5 m x 2.5 m
Height of the unit	8 m

Drying air in silica gel bed (10):

A five stage fluidized absorber of 1.2 m^2 cross section and 6.1 mm height was used to reduce the moisture content of air from 0.000919 Kg/Kg to 0.0015 Kg/Kg. of air by Cox. Silica gel beads of 3.2 to 4.8 mm size were used.

Recovery of carbon disulfide(11) :

Rawson described the first commercial process for the recovery of CS_2 from exhaust gases from a viscose plant using activated carbon in a multistage fluidized bed. The tower diameter was 11.6 m and the material depth on each stage was 5 to 8 cm.

Activation of charcoal (12) :

Charcoal is activated by the gassification with steam and CO_2 of wood in a two stage counter-current unit. Multistaging gives a more uniform residence time distribution for the solids and also helps recover heat for the endothermic gassification reaction by secondary combustion of CO and H_2 produced from the solids.

Calcination of limestone (13) :

A five stage counter-current fluidization column is used for the calcination of lime-stone. Large amount of heat is recovered using five stages. The column diameter is 4 m and height of the column is 14 m. This unit can handle 100 tons of limestone per day. The reaction takes place is

$$CaCO_3 \xrightarrow{1000^\circ C} CaO + CO_2 + 42.9 \text{ Kcal/mole}$$

Reduction of iron oxide :

A three stage counter-current unit was used to reduce iron oxide by Alan Wood Steel Company in Conshohoken. The reactor vessel for the 50 tones/day plant is about 1.7 m I.D. and 29 m high.

$$\underset{\text{magnetite}}{Fe_3O_4} + 4H_2 \longrightarrow Fe + 4H_2O$$

Roughly 1.4 tons of high grade magnetite ore, 0.051 - 0.056 tons of hydrogen and 0.25 ton of oxygen are needed to produce 1 ton of iron by this process.

Adsorption of Hydrocarbons (14) :

Continuous multistage fluid char adsorption process for seperation and recovery of gaseous component was developed by Esso Laboratories. The main unit, which is similar to a bubble cap-distillation column, has 20 bubble cap trays with heating and cooling coil. Various products draw-offs with auxilliary scrubbers and regenerators. Based on pilot plant data the above unit was found suitable for removal and concentration of components like C_3 & C_4 from effluent gases in petroleum and petro-chemical plants.

Pyrolysis of Western Bituminous coal (15) :

In order to prevent agglomeration of coal multistage fluidized beds are used for the Pyrolysis of Coal. Coal is heated to successively higher temperatures in different stages. Each successive temperature is selected to be just below the agglomerating temperature of coal. As the volatile products are released at one temperature, the agglomeration temperature indeases. The heated coal can then be transferred to a higher temperature stage, but again just below the agglomeration temperature in that stage. In this way the coal could progress through the temperature range of devolatilization without agglomeration.

CHAPTER - II

FLOW PROPERTIES OF SOLIDS :

Multistage fluidization process involve flow of solids from stage to stage. The nature of flow of solids from feed hopper into the system and flow through the downcomers between the stages is essentially slip-stick flow (gravity flow) whereas the flow of solids from the downcomer to the overflow weir across the tray is a case of aerated flow. Thus it is important to know the flow characteristics of solids before conducting the experiments on multistage fluidization units. The material used in the experiment included crushed Bauxite, Calcite and Coal of different sizes and for all these material the following flow properties have been determined.

Particle size :

Crushed material is analysed in a sieve shaker using B.S. shieves ranging from -20 to +60 mesh number. Average opening diameter of the sieve through which it passes and the sieve on which it is retained, is the diameter of the particle. The intermediate sieves were 30, 40 and 50 mesh number.

Density :

Density is determined by usual liquid displacement method. The two liquids i.e. water and carbon-tetrachloride are used to ensure that the solid material is not internally porous. The weighed quantity of solid material is poured in a measuring cylinder into which the liquid is kept before the solid material is poured. The liquid displaced gives the actual volume of solids. Thus weight of solids dividing by its volume gives density.

Void-Fraction :

Void fraction is determined by knowing the volume of bed and the volume of solids. The difference of the two gives the void volume. The void volume dividing by the volume of bed gives void fraction.

Angle of Repose :

The solid material is first filled in a cylinder and then the cylinder is lifted to a small height. The material into it starts coming out and makes a cone. The base angle of this cone is the Angle of Repose, which may be calculated by knowing the dimensions of the cone.

Table 2.1 shows the various properties of the materials used for investigation.

Table 2.1

Material	Size		Density	Porosity	G_{mf}
	Mesh No.	D_p mm	ρ gm/cc	ϵ	kgm/m^2-hr.
	-20+30	0.648	2.4	0.43	800
Calcite	-30+40	0.426	2.4	0.40	347
	-50+60	0.287	2.4	0.32	157
	-20+30	0.648	2.17	0.526	725
Bauxite	-30+40	0.426	2.17	0.464	314
	-50+60	0.287	2.17	0.30	142
Coal	-20+30	0.648	1.6	0.55	535
	-30+40	0.426	1.6	0.45	231

CHAPTER - III

EXPERIMENTAL SETUP :

A two stage counter-current fluidizer have been built of prospex 50 mm I.D tube each stage is 225 mm long. The two stages are seperated by Aluminimum grid, 3 mm thick (details are shown in fig. 3.2). The Column was tightened by tie rods as shown in sketch (fig. 3.1 and 3.3). The Aluminium grid plates were drilled with 1.5 mm holes on a traingular pitch to act as air distributor. The total opening provided for air flow is about 8-9% of the total tower cross-section. The grids were covered with fine copper mesh of 100 size to ensure that the solids supported above the grid do not fall through or chock the air holes in the distributor grid.

Each tray was provided with copper down comer of 10 mm I.D. (12 mm O.D.) and about 230 mm long. The downcomers fitted on grid plate by a nut arrangement. For the down-flow of solids from the bottom most stage a copper downcomer of same size was provided with a gland arrangement to adjust the height of the overflow. For collection of solids from bottom a narrow long perspex tube of 39 mm diameter was provided.

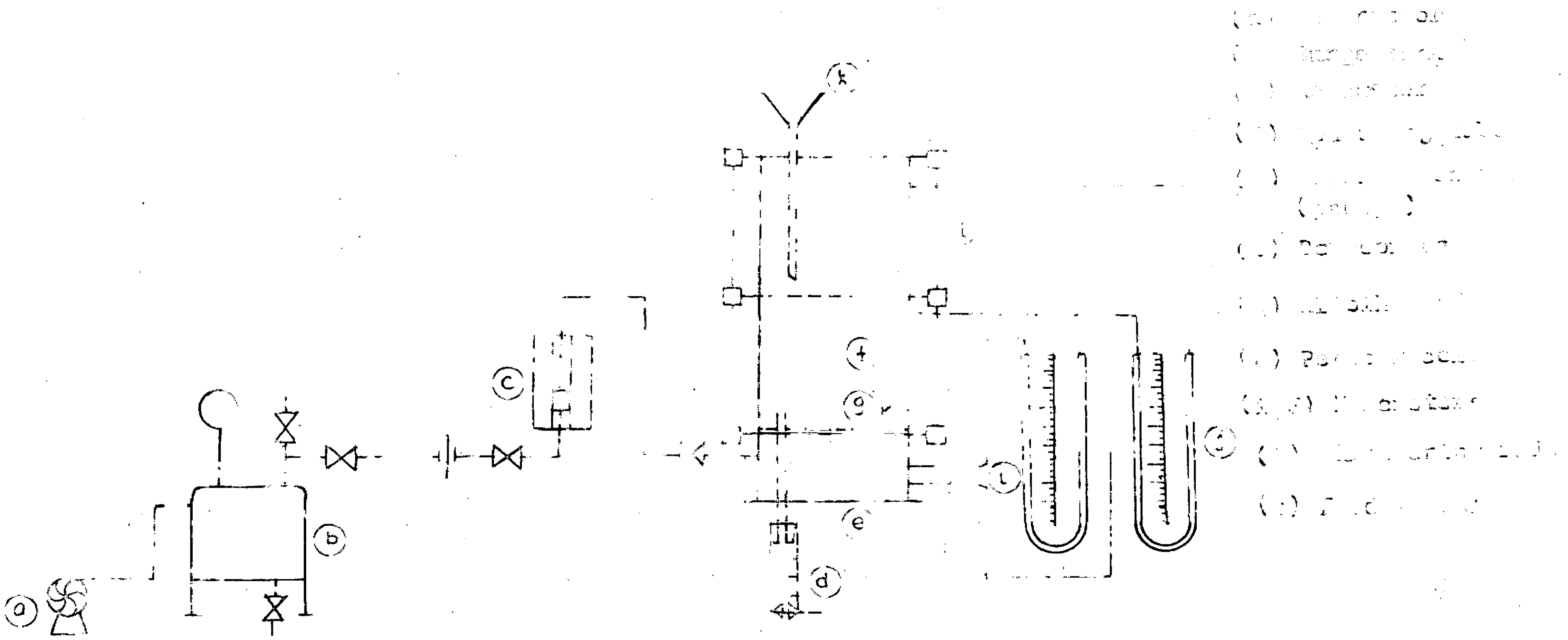

FIG 3.1 SCHEMATIC DIAGRAM OF THE APPARATUS

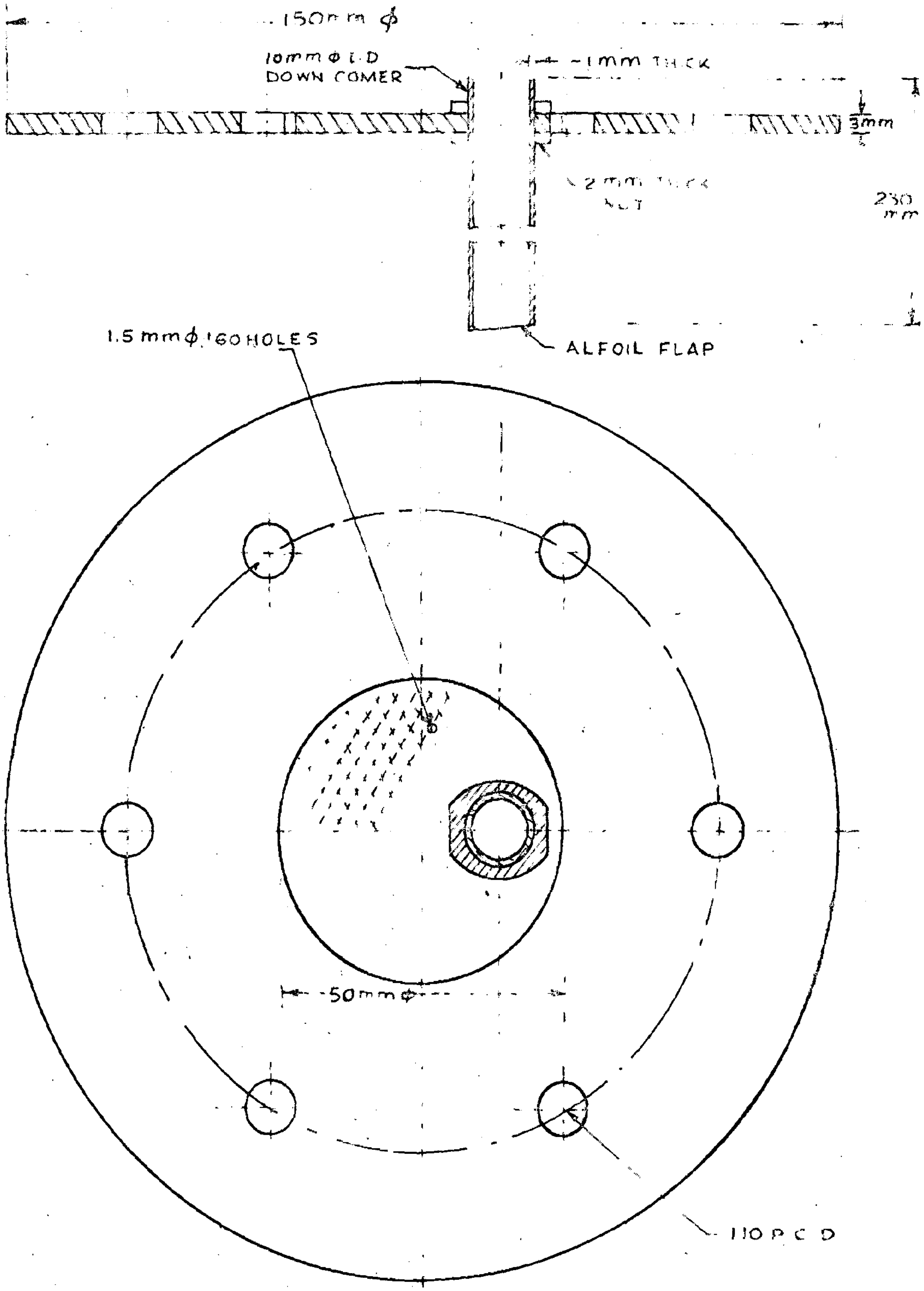

FIG. 3.2 DETAILS OF GRID PLATE

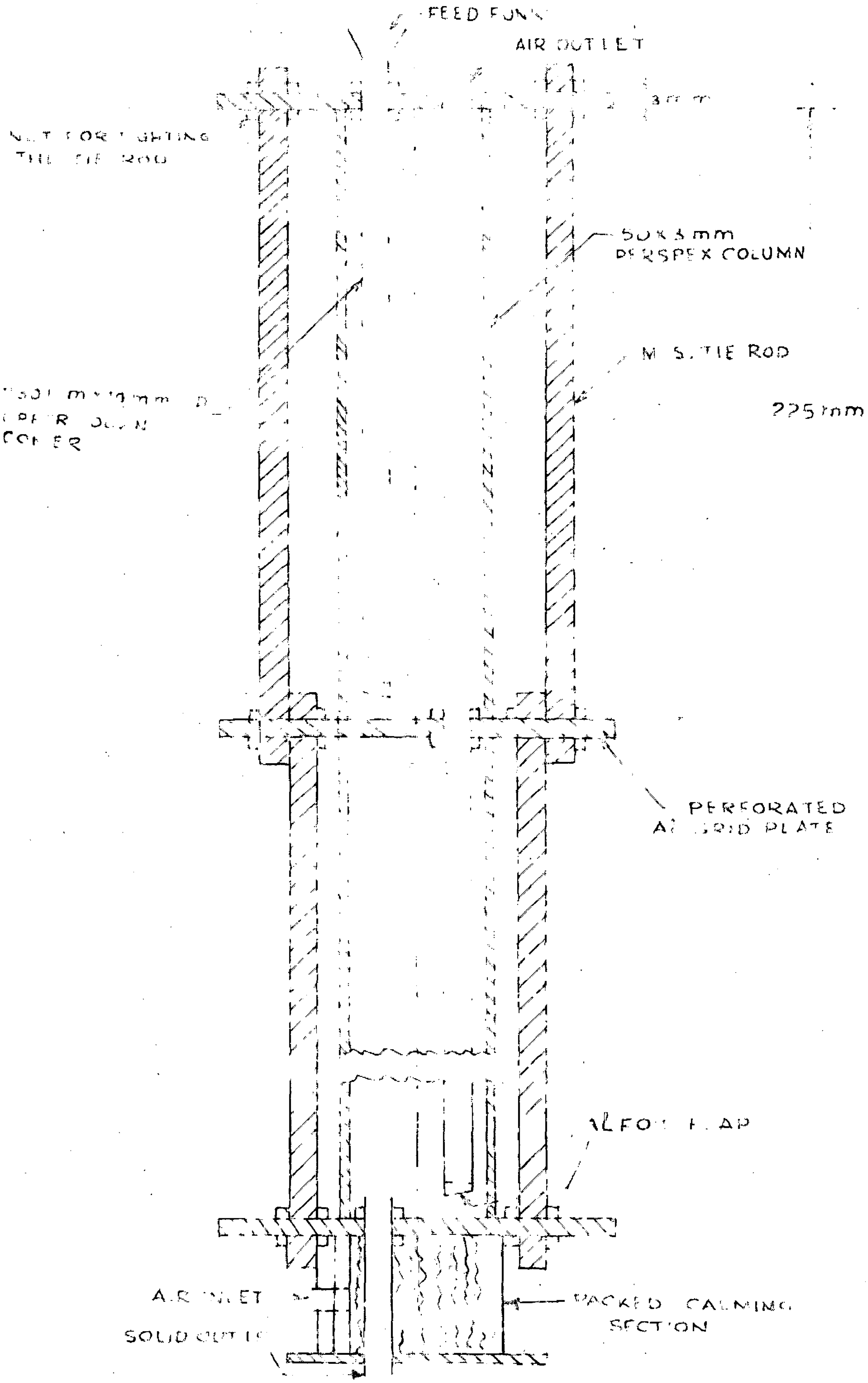

FIG 3.3 TWO STAGE COLUMN DETAILS

The solids fed to the top most stage by gravity through openings of different sizes. This was achieved by utilizing glass funnels of different throat openings as feed hoppers.

Air was drawn from a compressor via a surge tank and fed it to the column through a calibrated Rotameter. A cloth bag is provided in the exit air line to ensure that any fine solid entrained is collected.

Air is introduced in the column through a packed cylindrical calming section 50 mm I.D. and 40 mm height, to ensure proper distribution of air. The packing used were rashigring of 7 mm size.

The data ~~was~~ obtained on the flow rate of solids of different material and of different sizes for a given apparatus setting as a function of air velocity. The solids studied included crushed bauxite, calcite and coal in the sizes ranging from -20+60. Different orifice openings used were 0.45 cm, 0.6 cm and 0.7 cm.

Before conducting experiment the entire unit was tested for air leakages by soap solution.

Operation :

Air drawn from the ~~pum~~ compressor is introduced at the bottom of the column and air flow rates varied from zero to 2,500 L.P.H. Solids are fed at the top and the feed rate is varied by changing funnels of different ~~dim~~ discharge openings.

Solids used were bauxite, calcite and coal of different mesh sizes.

The solids flow down via the downcomer on/to the top stage and the flow stopped as soon as the material formed a cone below the bottom of downcomer. At this stage air flow was started and it was observed that the air flowed through the tray free of solids and the downcomer of the stage below. To make the solids flow on the entire tray air flow was successively started and stopped so that the solids below the feeding downcomer started moving cross-wise. After some affort it was usually observed that the top stage get loaded more and more as long as the gas flow continues. This happened due to the air by-passing the loaded grid and flowing through the downcomer of the stage below, which offered lesser resistance. When the air flow is stopped the solids moves from top to bottom stage by small quantity untill the bottom of the downcomer got covered by the solids. The solids were made to flow across the stage by successive air starting and stopping. In the operation when both the stages had material it is observed that the column performance was rather erratic. Often it is noticed that solids accumulated in one of the stages (usually top stage) or the solids got completely depleted in one of the stages (usually bottom stage). An analysis of this problem gave the clue that air had a tendency to by-pass the grids loaded with material and flow via the downcomers. Another problem that can happen at high velocities of air is the non flow of feed material through the feed funnel. To avoid these problems, small weightless Al foil flaps were made and attached to the bottom of downcomers. This improved the column performance considerably as the flap tended to close the downcomer opening whenever air had a tendency to by-pass the grid. Such situation usually occured at the start up when downcomer was empty of solids. The foil being weightless offered practically no resistance to solids downflow. Even with flaps attached to the downcomers, the column performance was smooth over a limited range of gas velocities. The data were taken on solid flow rate under steady state conditions as the function of gas flow rate in the smooth zone of operation.

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CHAPTER - IV

RESULTS AND DISCUSSIONS :

Data were taken for solids flow rate at different air flow rate and are reported from table 4.1 to table 4.8. The three systems on which the data were taken are Calcite Air, Bauxite Air and Coal Air. The data were taken for a fixed column geometry.

Table 4.1

System : Calcite-Air
Particle size: -20+30
Density : 2.4 Kgm/lit.
Porosity : 0.43

D_o = 6 mm

Sl. No.	Gas flow rate(G_f)		Solids flow rate W_s, kgm/M^2-hr.	W_s/G_f
	Lit/Hr	Kgm/M^2-hr.		
1.	2140	1270	487	0.383
2.	2210	1312	786	0.600
3.	2250	1335	1018	0.762
4.	2300	1362	1180	0.862
5.	2340	1387	1320	0.952
6.	2380	1410	1438	1.02

D_o = 7 mm

Sl. No.	Lit/Hr	Kgm/M^2-hr.	W_s, kgm/M^2-hr.	W_s/G_f
1.	2000	1187	279	0.235
2.	2050	1218	842	0.692
3.	2140	1270	1172	0.924
4.	2210	1312	1445	1.100
5.	2250	1335	1590	1.19
6.	2300	1362	1680	1.235
7.	2340	1387	1690	1.22
8.	2380	1410	1565	1.11

Table : 4.2

System : Calcite-Air
Particle size: -30+40
Density : 2.4 kgm/lit
Porosity : 0.40 .

D_o = 4.5 mm

Sl. No.	Gas flow rate (G_f) Lit/Hr	Gas flow rate (G_f) Kgm/M^2.Hr.	Solids flow rate, W_s, Kgm/M^2.Hr.	W_s/G_f
1.	1735	1030	479	0.465
2.	1825	1085	578	0.533
3.	1900	1128	635	0.563
4.	2000	1187	880	0.742
5.	2050	1218	1030	0.845
6.	2140	1270	1320	1.04
7.	2210	1312	890	0.678

D_o = 6 mm

Sl. No.	Gas flow rate (G_f) Lit/Hr	Gas flow rate (G_f) Kgm/M^2.Hr.	Solids flow rate, W_s, Kgm/M^2.Hr.	W_s/G_f
1.	1735	1030	923	0.896
2.	1825	1085	1360	1.253
3.	1900	1128	1380	1.225
4.	2000	1187	1730	1.458
5.	2050	1218	1900	1.56
6.	2140	1270	2030	1.60
7.	2180	1295	2170	1.678
8.	2210	1312	2255	1.718
9.	2250	1335	2180	1.633
10.	2300	1362	1790	1.315

		D_o = 7 mm		
1.	1735	1030	1670	1.62
2.	1825	1085	2175	2.0
3.	1900	1128	2620	2.33
4.	2000	1187	2765	2.33
5.	2050	1218	2830	2.32
6.	2140	1270	2890	2.28
7.	2180	1295	2540	1.935

Table 4.3

System : Calcite-Air
Particle size : -50+60
Density : 2.4 kgm/lit
Porosity : 0.32

$D_o = 4.5$ mm

Sl. No.	Gas flow rate (G_f) Lit/Hr.	Gas flow rate (G_f) Kgm/M^2-Hr	Solids flow rate W_s, Kgm/M^2-Hr.	W_s/G_f
1.	1735	1030	286	0.278
2.	1780	1055	452	0.428
3.	1825	1085	602	0.555
4.	1900	1128	701	0.612
5.	2000	1187	659	0.555

$D_o = 6$ mm

1.	1735	1030	381	0.370
2.	1780	1055	554	0.525
3.	1825	1085	665	0.613
4.	1850	1099	754	0.685
5.	1900	1128	775	0.688
6.	2000	1187	745	0.628

		D_o = 7 mm		
1.	1735	1030	474	0.460
2.	1780	1055	614	0.582
3.	1825	1085	744	0.685
4.	1850	1099	780	0.710
5.	1900	1128	807	0.715
6.	2000	1187	763	0.644

Table 4.4

System : Bauxite-Air

Particle size : -20+30

Density : 2.17 kgm/lit

Porosity : 0.526

$D_o = 6$ mm

Sl. No.	Gas flow rate(G_f)		Solids flow rate W_s, Kgm/M^2-hr.	W_s/G_f
	Lit/hr.	Kgm/M^2-hr		
1.	1935	1149	296	0.258
2.	2000	1187	370	0.312
3.	2050	1218	450	0.370
4.	2100	1245	465	0.374
5.	2140	1270	514	0.404
6.	2180	1295	590	0.456
7.	2210	1312	696	0.530
8.	2300	1362	864	0.634
9.	2380	1410	980	0.695

$D_o = 7$ mm

1.	1900	1128	464	0.402
2.	1935	1149	531	0.462
3.	2000	1187	696	0.587
4.	2050	1218	745	0.612
5.	2140	1270	800	0.630
6.	2180	1295	864	0.666
7.	2210	1312	940	0.715
8.	2250	1335	980	0.734
9.	2300	1362	1020	0.748
10.	2340	1387	950	0.685
11.	2380	1410	865	0.614

Table 4.5

System : Bauxite-Air

Particle size: $-30+40$

Density : 2.17 kgm/lit

Porosity : 0.464

$D_o = 4.5$ mm

Sl. No.	Gas flow rate(G_f)		Solids flow rate W_s, Kgm/M^2-hr.	W_s/G_f
	Lit/Hr.	Kgm/M^2-hr		
1.	1780	1035	437	0.414
2.	1825	1085	944	0.870
3.	1850	1099	1065	0.97
4.	1900	1128	1325	1.175
5.	1935	1149	1448	1.260
6.	2000	1187	1633	1.377
7.	2020	1197	1642	1.373
8.	2050	1218	1665	1.370
9.	2100	1245	1810	1.455
10.	2140	1270	1990	1.568
11.	2180	1295	1885	1.455
12.	2210	1312	1865	1.42

$D_o = 6$ mm

1.	1735	1030	940	0.912
2.	1825	1085	1040	0.958
3.	1900	1128	1400	1.24

		D_o = 6 mm		
4.	2000	1187	1815	1.53
5.	2020	1197	2005	1.675
6.	2050	1218	2170	1.782
7.	2100	1245	2330	1.873
8.	2140	1270	2270	1.788
9.	2210	1312	2660	2.03
10.	2300	1362	2490	1.828
		D_o = 7 mm		
1.	1780	1055	441	0.418
2.	1825	1085	1052	0.97
3.	1850	1099	1690	1.44
4.	1900	1128	1787	1.585
5.	1935	1149	1885	1.642
6.	2000	1187	2270	1.915
7.	2020	1197	2330	1.95
8.	2050	1218	2525	2.075
9.	2100	1245	2630	2.11
10.	2140	1270	2720	2.14
11.	2210	1312	2780	2.12
12.	2300	1362	2815	2.065
13.	2340	1387	2815	2.03

Table : 4.6

System : Bauxite-Air

Particle size : -50+60

Density : 2.17 kgm/lit

Porosity : 0.3

$D_o = 4.5$ mm

Sl. No.	Gas flow rate (G_f)		Solids flow rate, W_s, kgm/M^2-hr.	W_s/G_f
	Lit/hr.	Kgm/M^2-hr.		
1.	1735	1030	173	0.168
2.	1780	1055	263	0.249
3.	1825	1085	314	0.289
4.	1850	1099	349.5	0.318
5.	1900	1128	377	0.335
6.	1935	1149	338	0.294
7.	2000	1187	516	0.435
8.	2020	1197	562	0.470
9.	2050	1218	764	0.627
10.	2100	1245	690	0.602

$D_o = 6$ mm

Sl. No.	Lit/hr.	Kgm/M^2-hr.	W_s, kgm/M^2-hr.	W_s/G_f
1.	1735	1030	250	0.243
2.	1780	1055	534	0.506
3.	1825	1085	640	0.590
4.	1850	1099	720	0.655

5.	1900	1128	842	0.746
6.	1935	1149	915	0.797
7.	2000	1187	980	0.825
8.	2020	1197	1010	0.844
9.	2050	1218	1030	0.845
10.	2100	1245	940	0.765
$D_o = 7$ mm				
1.	1735	1030	292	0.283
2.	1780	1055	550	0.521
3.	1825	1085	668	0.616
4.	1850	1099	763	0.695
5.	1900	1128	1000	0.887
6.	1935	1149	1120	0.975
7.	2000	1187	866	0.730
8.	2020	1197	1142	0.955
9.	2050	1218	1162	0.955
10.	2100	1245	1142	0.917

Table 4.7

System : Coal-Air
Particle size: -20+30
Density : 1.6 Kgm/lit
Porosity : 0.55

D_o = 6 mm

Sl. No.	Gas flow rate(G_f)		Solids flow rate, W_s, Kgm/M^2-hr.	W_s /G_f
	Lit/hr.	Kgm/M^2-hr		
1.	1900	1128	29.9	0.02655
2.	1935	1149	38.9	0.0339
3.	2000	1187	73.9	0.0622
4.	2020	1197	131.8	0.110
5.	2050	1218	189.2	0.1555
6.	2100	1245	189.2	0.1520
7.	2140	1270	224	0.1762
8.	2180	1295	263	0.203
9.	2210	1312	311	0.237
10.	2250	1335	292.5	0.219
11.	2300	1362	342	0.251
12.	2380	1410	364	0.258

$D_o = 7$ mm

1.	1900	1128	33.5	0.0297
2.	1935	1149	56	0.0488
3.	2000	1187	97	0.0817
4.	2020	1197	126	0.1052
5.	2050	1218	162	0.133
6.	2100	1245	174.7	0.1402
7.	2140	1270	197.5	0.1555
8.	2180	1295	316	0.244
9.	2210	1312	293	0.223
10.	2300	1362	404	0.297
11.	2380	1410	430	0.305

Table 4.8

System : Coal-Air
Particle size : -30+40
Density : 1.6 Kgm/lit
Porosity : 0.45

$D_o = 4.5$ mm

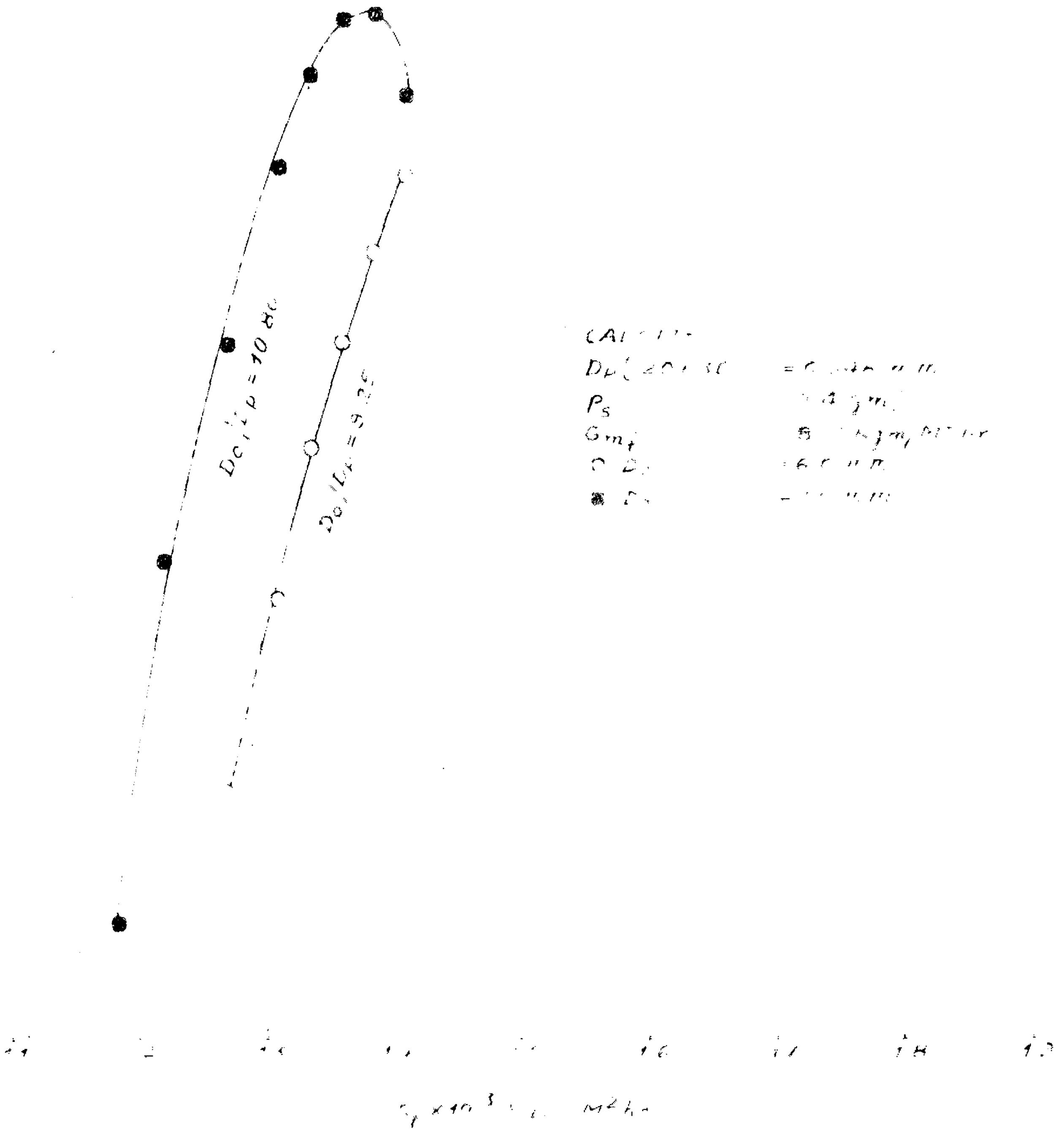
Sl. No.	Gas flow rate(G_f)		Solids flow rate W_s, Kgm/M^2-hr.	W_s /G_f
	Lit/hr.	Kgm/M^2-hr		
1.	1735	1030	50.2	0.0487
2.	1780	1055	154	0.146
3.	1825	1085	330	0.304
4.	1850	1099	392.5	0.358
5.	1900	1128	484	0.423
6.	1935	1149	770	0.670
7.	2000	1187	923	0.779
8.	2020	1197	880	0.810
9.				

$D_o = 6$ mm

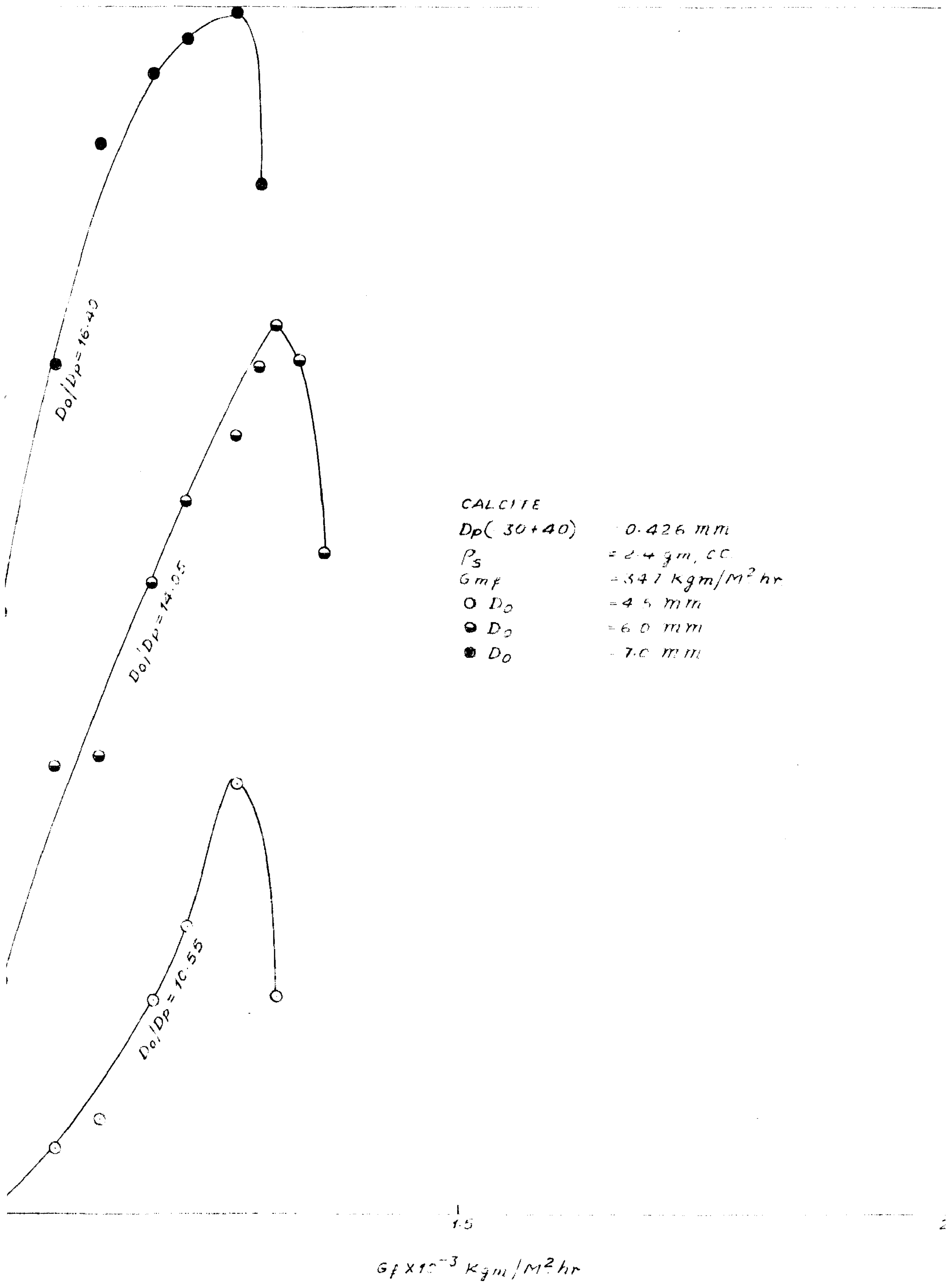
1.	1735	1030	294	0.285
2.	1780	1055	411	0.390
3.	1825	1085	560	0.516
4.	1850	1099	514	0.468
5.	1900	1128	587	0.521
6.	1935	1149	659	0.574
7.	2000	1187	710	0.598
8.	2020	1197	679	0.567

$D_o = 7$ mm

1.	1735	1030	486	0.472
2.	1780	1055	528	0.500
3.	1825	1085	636	0.586
4.	1850	1099	711	0.648
5.	1900	1128	755	0.670
6.	1935	1149	803	0.700
7.	2000	1187	1030	0.869



. 4.1 SOLIDS FLOW RATE Vs GAS FLOW RATE



.4·2 SOLIDS FLOW RATE V_S GAS FLOW RATE.

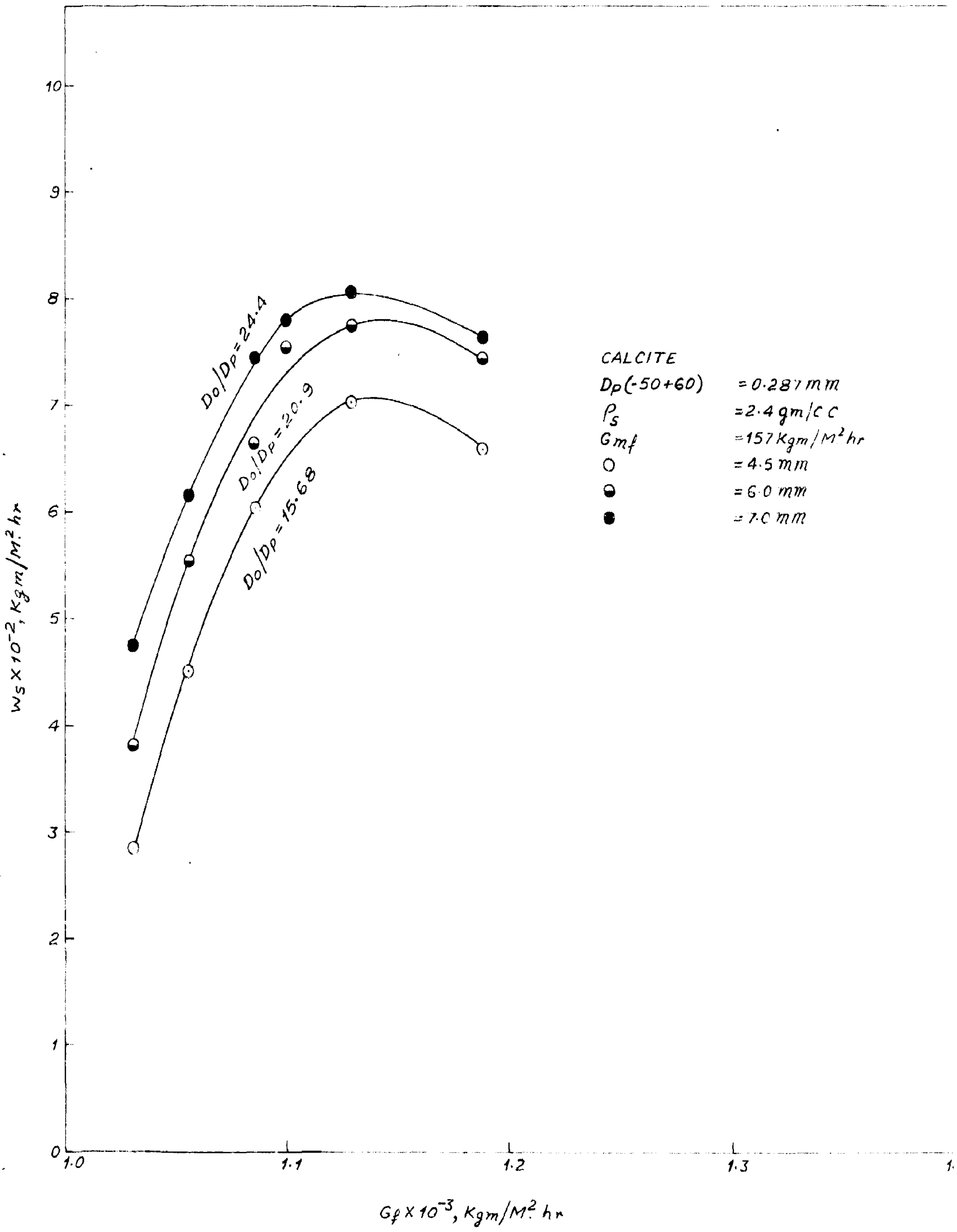


FIG. 4.3 SOLIDS FLOW RATE VS. GAS FLOW RATE.

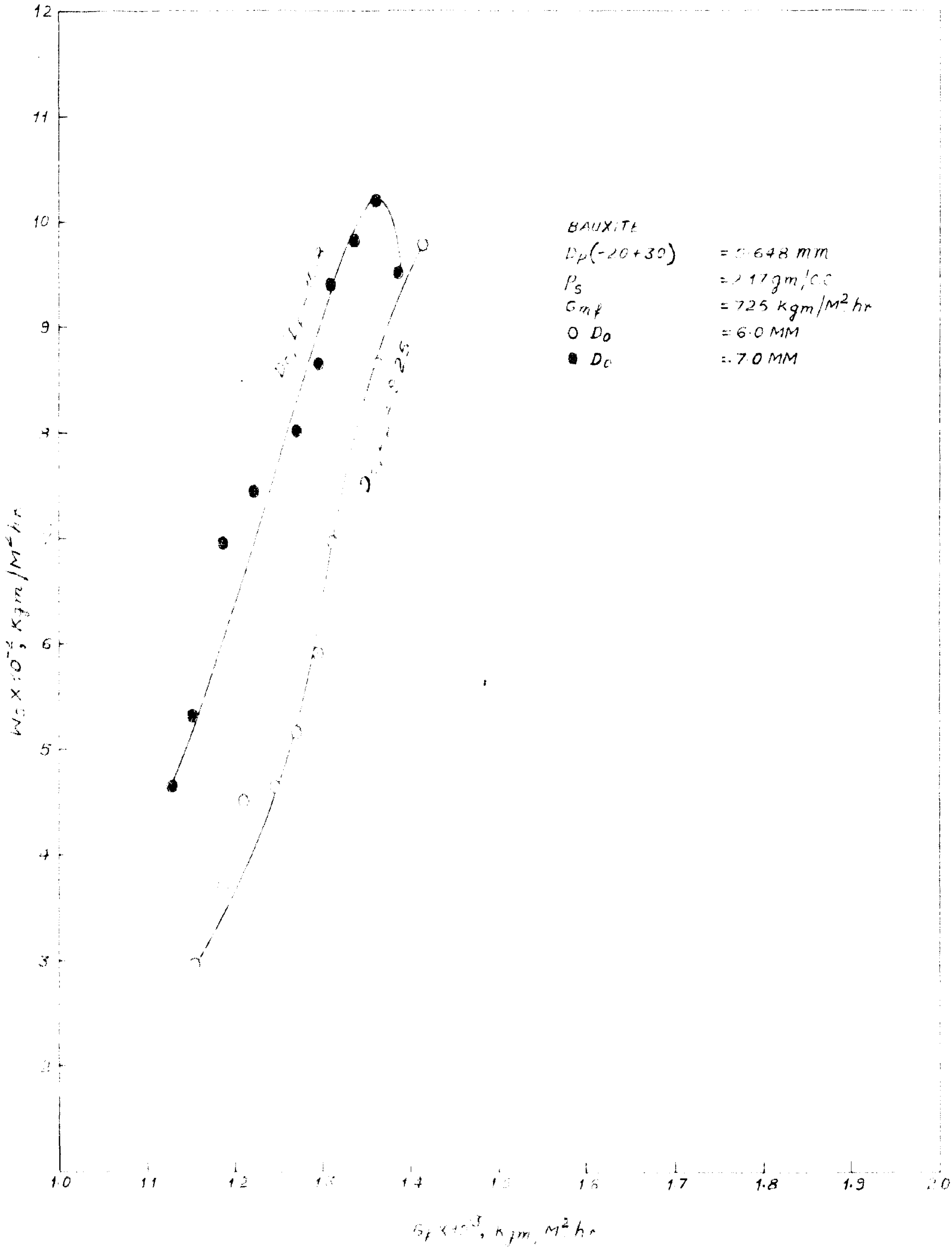


FIG. 4.4 SOLIDS FLOW RATE VS. GAS FLOW RATE.

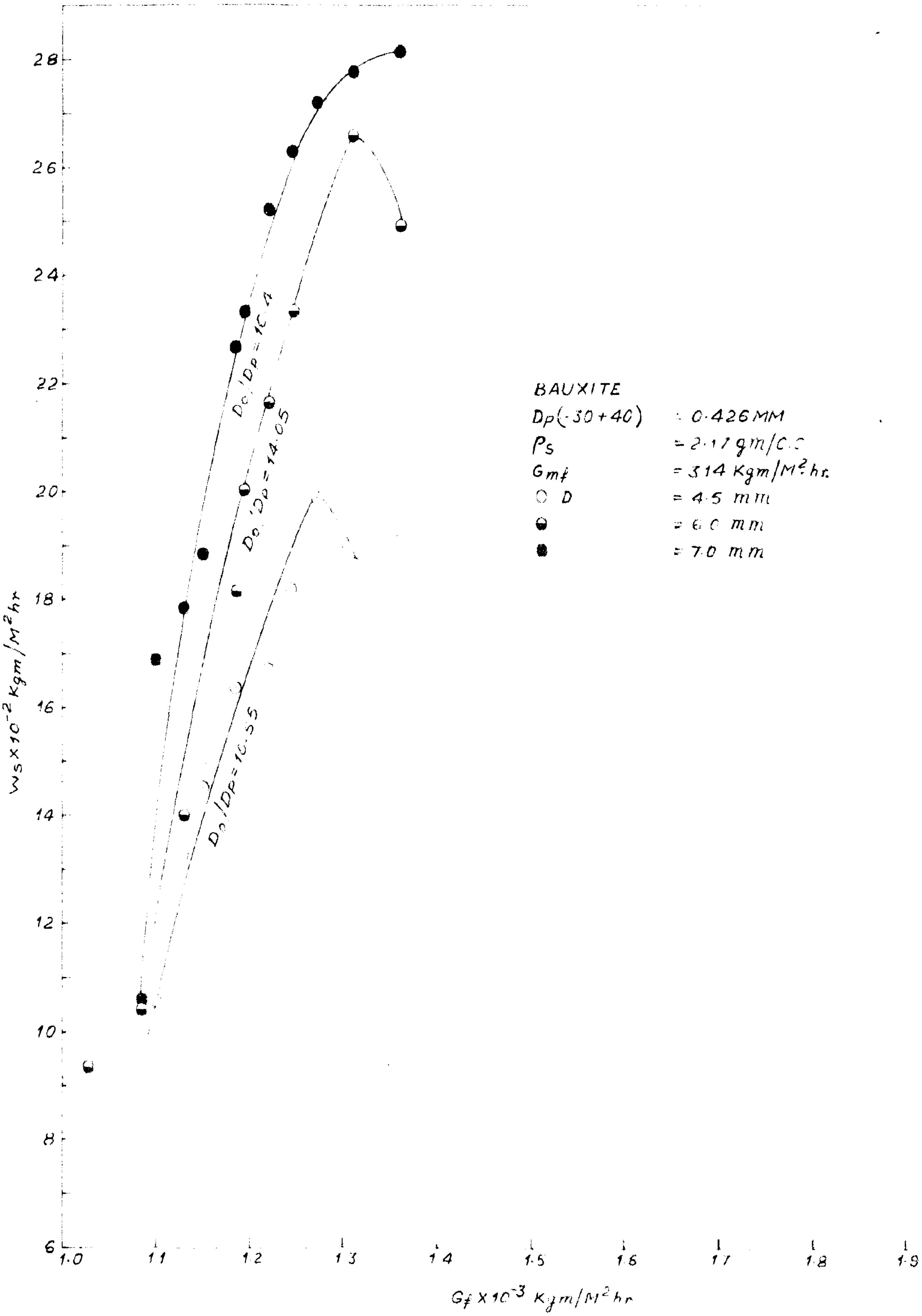


FIG.4.5 SOLIDS FL[illegible] RATE VS. GAS FLOW RATE.

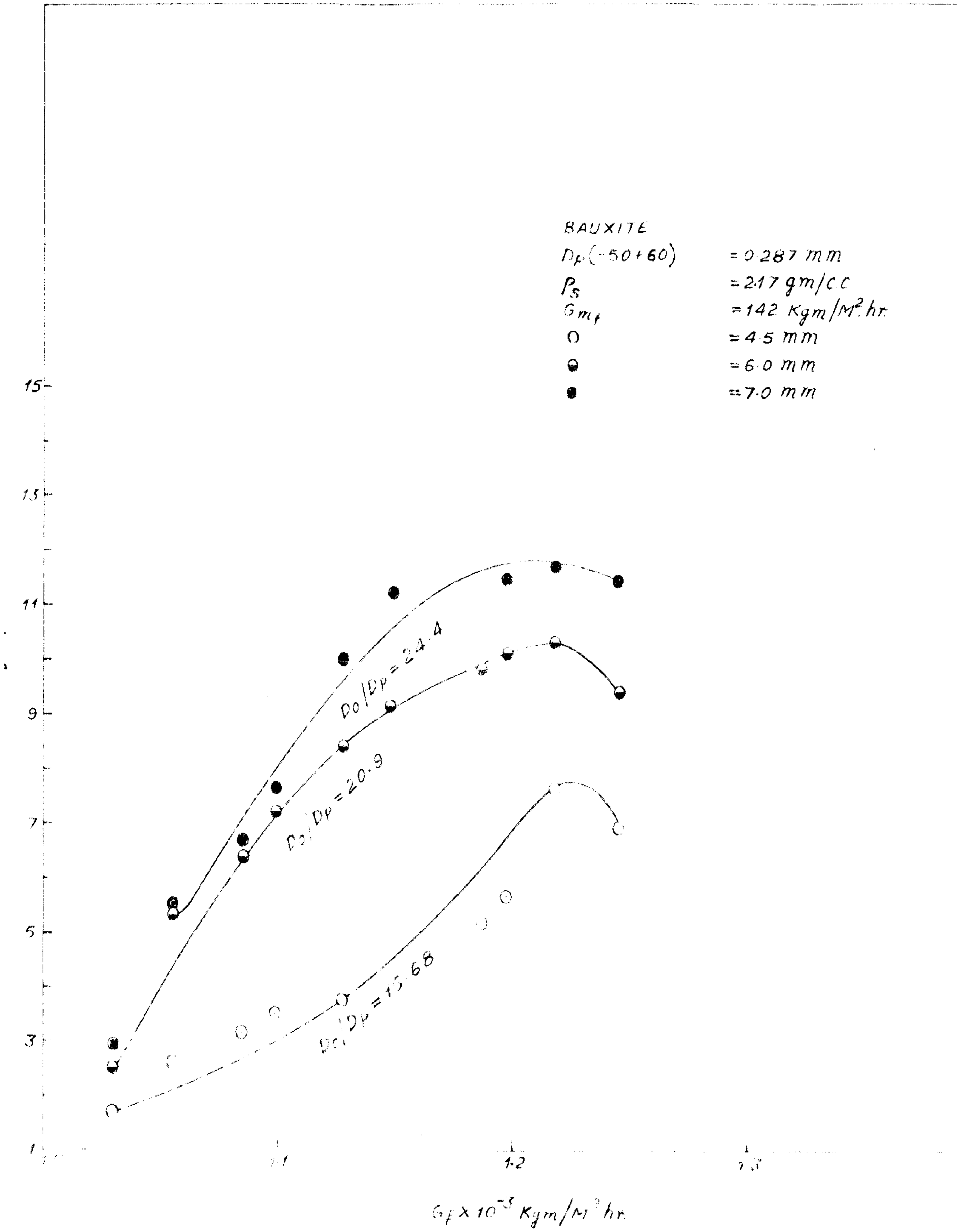


FIG. 4.6 SOLIDS FLOW RATE VS GAS FLOW RATE.

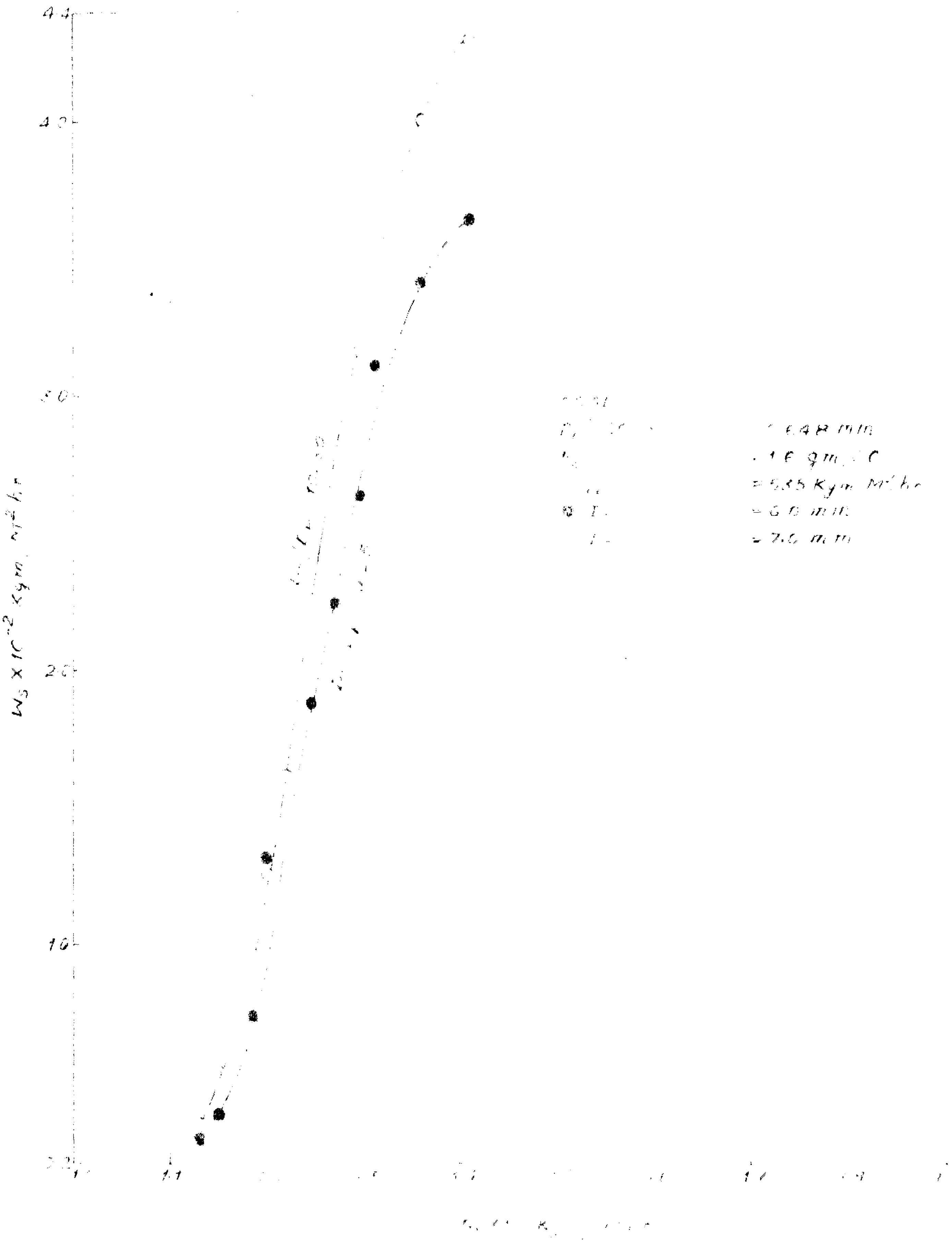


FIG. 4.7 SOLIDS FLOW RATE VS GAS FLOW RATE.

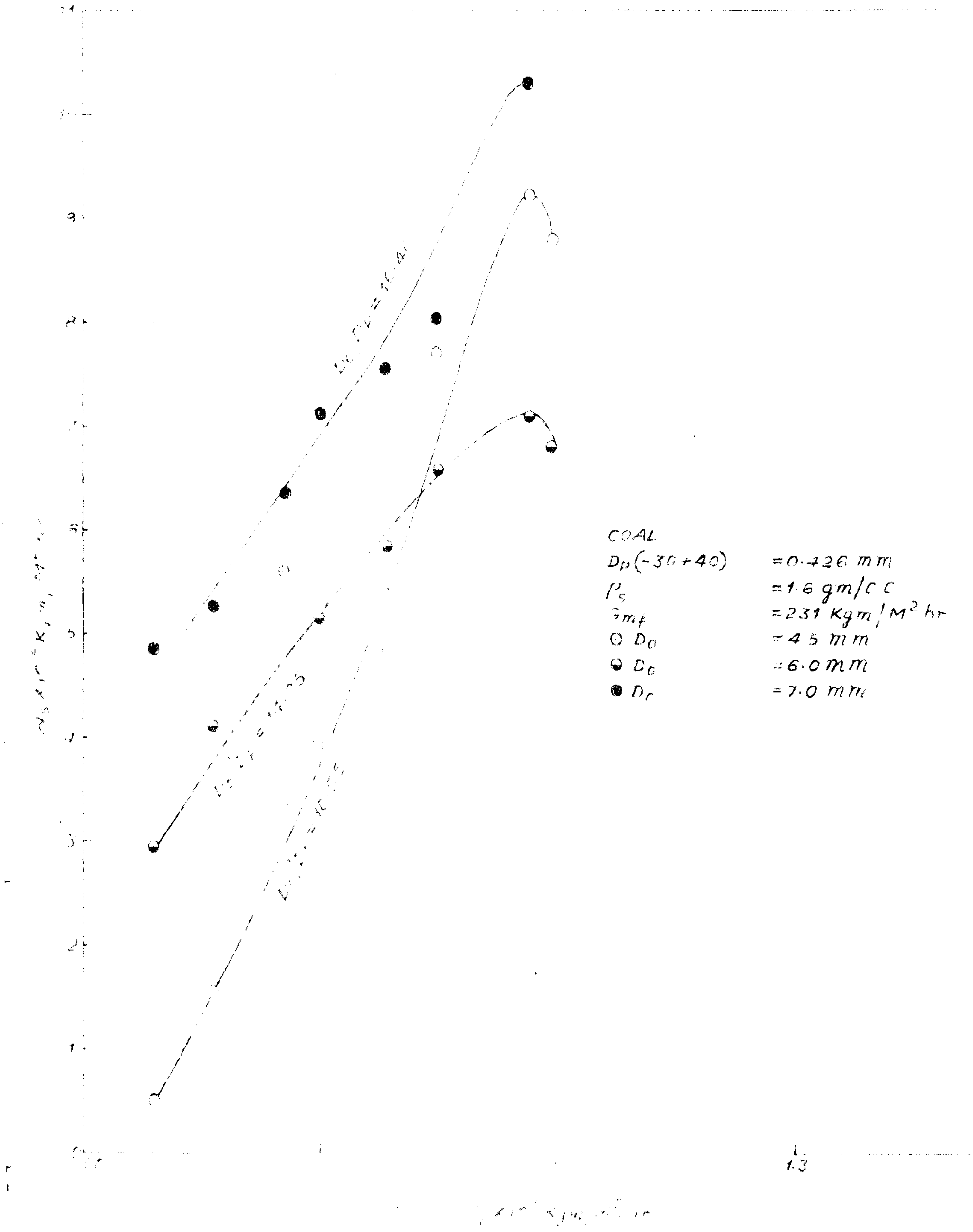


FIG. 4.8 SOLIDS FLOW RATE VS. GAS FLOW RATE.

CALCULATION OF PARTICLE REYNOLDS NUMBER

$$N_{Rep} = \frac{D_p \, G_f}{\mu_{air}} \; ; \qquad \mu_{air} = 0.018 \text{ c.p.}$$

Sl. No.	G_f, Kgm/M^2-hr.	N_{Rep}		
		D_p=0.648 mm (-20+30)	D_p=0.426 mm (-30+40)	D_p=0.287 mm (-50+60)
1.	1030	10.3	6.75	2.72
2.	1055	10.55	6.91	2.80
3.	1085	10.83	7.10	2.87
4.	1099	10.99	7.2	2.91
5.	1128	11.3	7.4	2.965
6.	1149	11.4	7.53	3.04
7.	1187	11.85	7.78	3.14
8.	1197	12.0	7.85	3.17
9.	1218	12.12	7.96	3.22
10.	1245	12.40	8.17	3.300
11.	1270	12.70	8.32	3.36
12.	1295	12.92	8.47	3.42
13.	1312	13.10	8.58	3.47
14.	1335	13.32	8.75	3.53
15.	1362	13.60	8.94	3.61
16.	1387	13.9	9.1	3.67
17.	1410	14.08	9.25	3.74

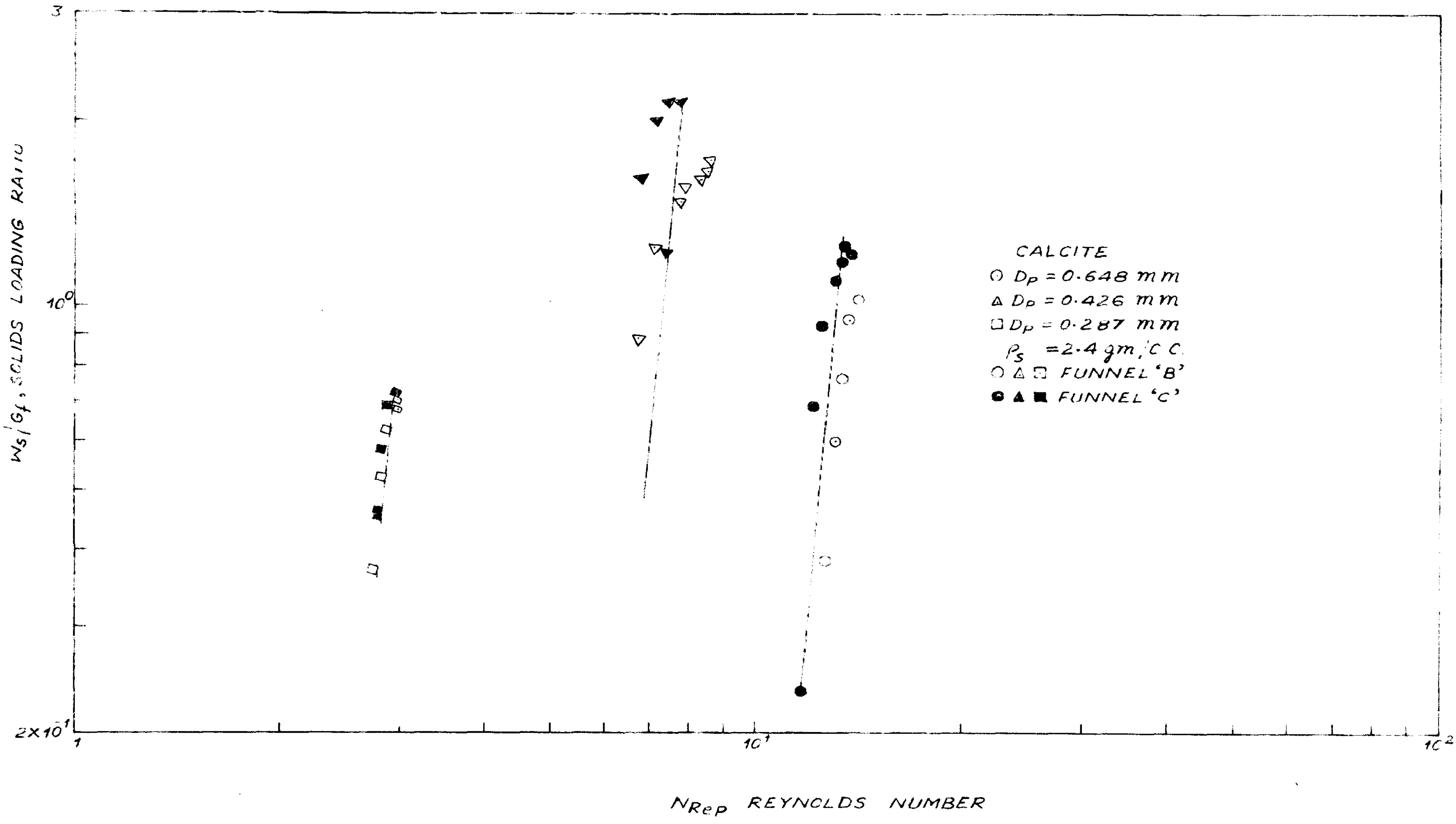


FIG.4.9 VARIATION OF SOLIDS LOADING RATIO WITH REYNOLDS NUMBER.

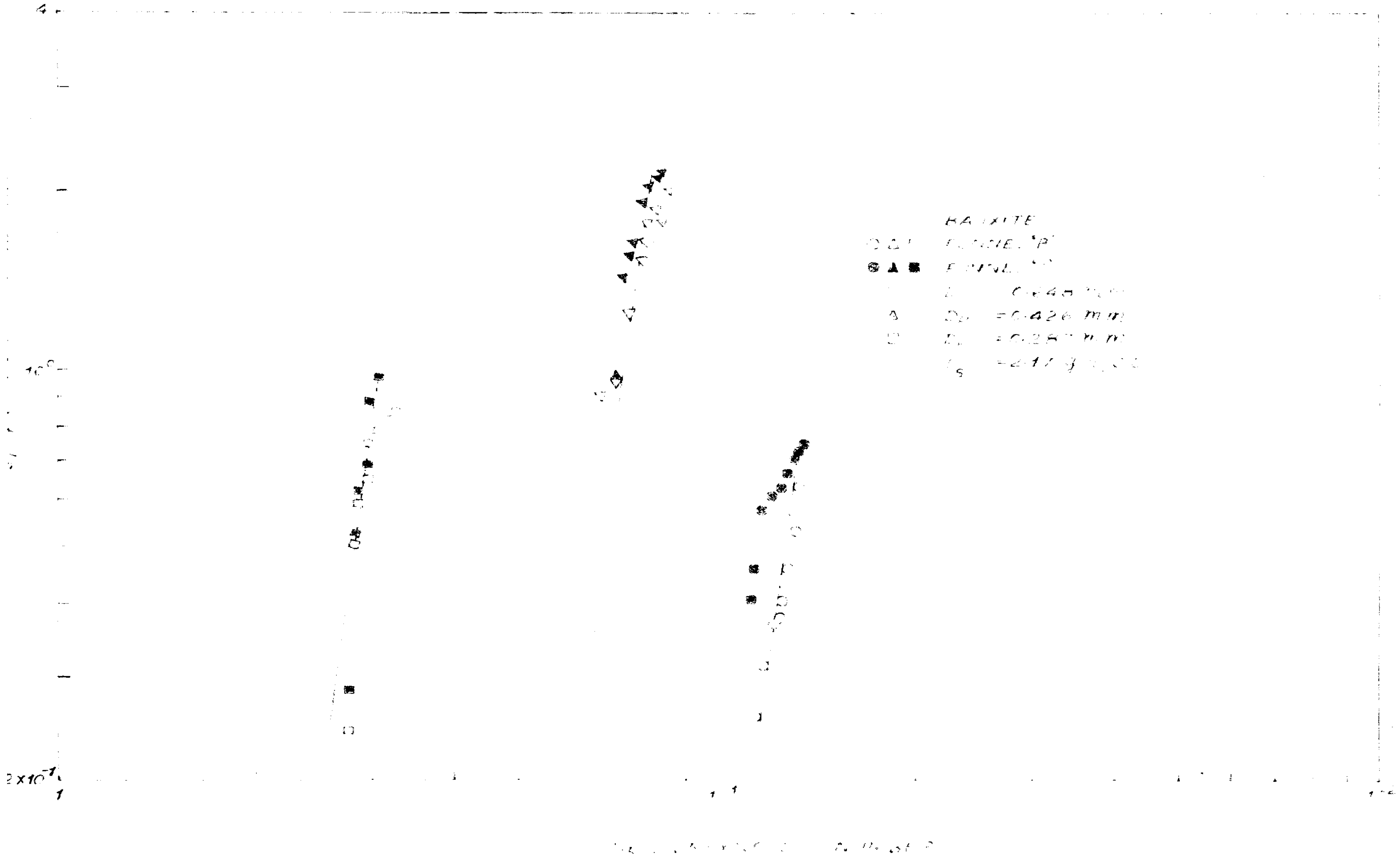


FIG. 4·10 VARIATION OF SOLIDS LOADING RATIO WITH REYNOLDS NUMBER

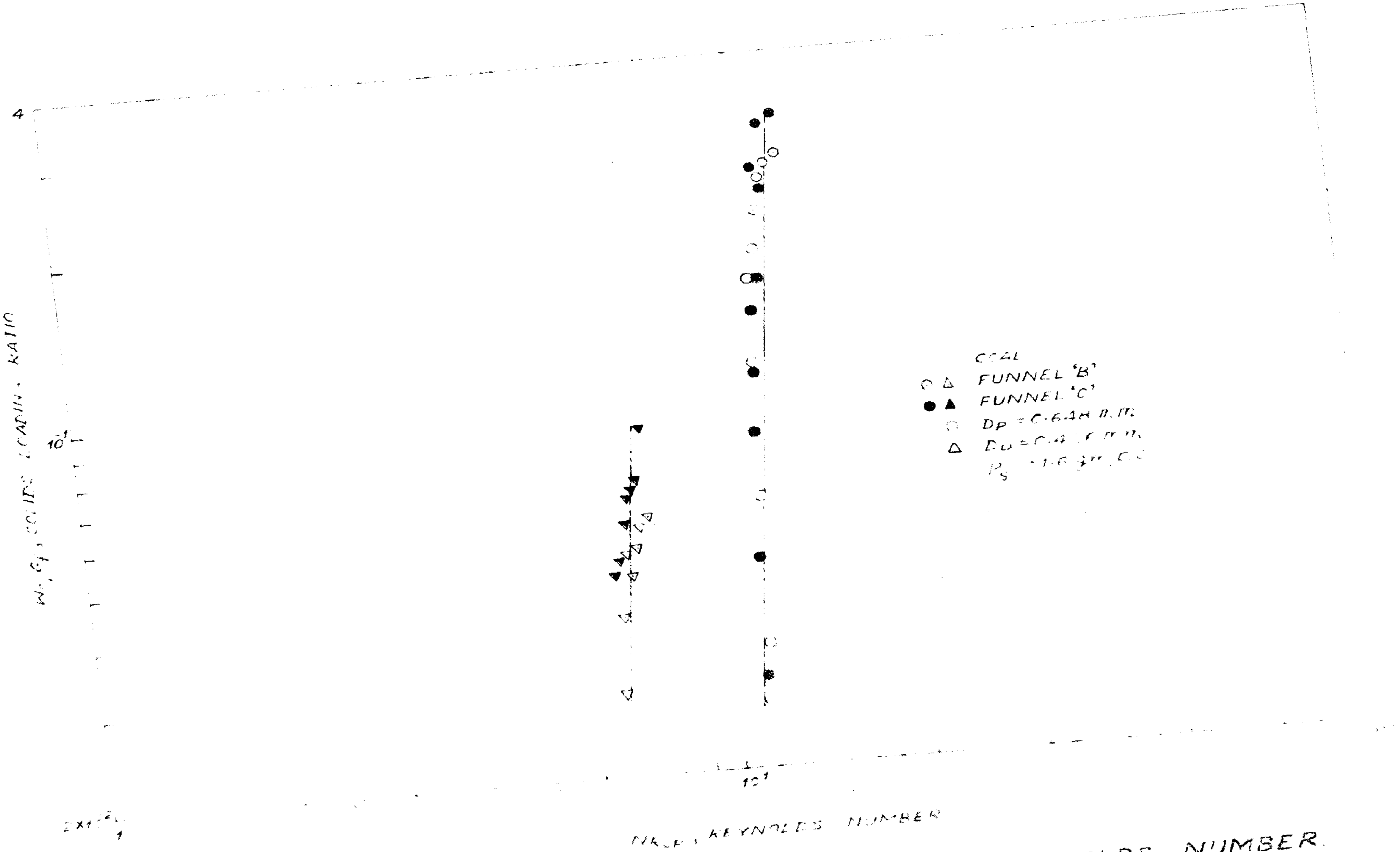


FIG. 4.11 VARIATION OF SOLIDS LOADING RATIO WITH REYNOLDS NUMBER.

DISCUSSIONS :

Graphs were plotted with solid mass flow rate based on empty tower cross-section (W_s, Kgm/M^2-hr) vs. mass flow rate of air also based on empty tower cross-section. These are shown for different materials and particle sizes in figures 4.1 to 4.8. As can be seen from fig. 4.1 for Calcite of -20+30 mesh size (D_p = 0.648 mm) the solids flow rate increases with gas flow rate for a given throat opening of a funnel upto a certain peak value after which increase in gas rate has lead to a decrease in solid rate.

Looking at fig. 4.2 for Calcite of -30+40 mesh size and funnel A(D_o=4.5 mm) corresponding to a D_o/D_p=10.55 the solids flow initiated at a certain gas velocity whose value is nearly 1.03x10^3 kgm/m^2-hr below this gas velocity the flow of gas was not sufficient to move the particle away from the downcomer bottom which would have allowed more fresh solids to enter. Thus it will be observed that a certain minimum velocity of air is essential to start the flow of solids. An increase in gas rate is accompanied by an increase in solid flow, upto a certain point. This is expected because at higher velocities the particle movement becomes more vigorous leading to clearing of area around the downcomer bottom. This increase the possibility of fresh solids being loaded into the column. As the gas velocity is increased a new force which will slowly become~~s~~ more predominent is the upward thrust of gas to entrain the particles on one hand and to close the flap below the downcomer. Thus a combination of these two lead to a reduction in the chances of solids to leave a stage and enter the downcomer via the overflow weir. Thus solids rate start decreasing after the peak value and large increase in gas velocity beyond this point resulting heavy intrainment and fluctuate column performance. Because of these reasons much data could not be taken on the flow of solids at high gas velocities approaching the terminal velocity.

In graph 4.2 it can also be seen that when throat diameter is increased (funnel B, D_o=6 mm), the flow of solids initiated around the same velocity of gas as was observed in case of funnel A. But increase in gas velocity for funnel B, higher solids flow rate than for funnel A at the same air rate. This only means that the ease of flow of solids through an orifice increases with increase in orifice diameter for the same particle diameter, a fact which is well established in case of gravity flow of bulk solids. The same trend is observed for funnel C (D_o=7mm) having a still larger throat diameter. Similar trends are observed for Calcite of size -20+30 fig. 4.1, -50+60 fig. 4.3.

It would be observed that an increase in particle size for the same material and same throat opening usually leads to lowering in solid flow rate for a given gas rate. This is due to a decrease in the value of D_o/D_p.

Similar graphs of solids flow rate versus gas flow rate were plotted for crushed bauxite in the size ranges -20+30, -30+40, -50+60 for the three funnels A.B,C as shown in fig. 4.4, 4.5, 4.6. Identical plots for coal are shown in fig. 4.7 and fig. 4.8. The general trend of solid flow as a function of gas rate will be summerized as follows:

- Smooth solid flow rate for a given condition at a certain air velocity whose value is higher than the corrosponding minimum fluidization velocity.
- For a given throat opening and particle size, solids flow rate increases with increase in gas rate upto a certain peak value after which it starts decreasing.
- At high gas velocity beyond the peak solid rate the column operation becomes erratic with large entrainment of solids.

- For a given particle size and gas flow rate of a material, the solids flow rate increases with increase in throat diameter. Similarly, it may be noted that for a given throat diameter and gas rate this solid flow rate decreases with increase in particle diameter.
- It seems that for a given apparatus setting particle size, throat diameter, solid flow rate seems to increase with increase in particle density for a given gas rate. However, this needs more ellaborate study before drawing certain conclusion.

To evaluate a functional relation between G_f and W_s, graphs were plotted with different material, particle sizes, between particle Reynolds number in x - axis $(D_p G_f/\mu)$ versus the solids loading ratio R which is W_s/G_f as shown in fig. 4.9 to 4.11 on log-log graph papers.

Looking at fig. 4.9 for Calcite it will be observed that in the zone of smooth flow of solids upto the peak value the value of R (W_s/G_f) increases almost linearly with increase in particle Reynolds number for a given particle size and throat opening. Identical lines were obtained for different throat openings. It was observed for most of materials that R is a strong function of N_{Rep} with an average slope of 9.9 for all the materials and all funnel under study. The results of these graphs may be summerised as under:

- The exact value of R for a given N_{Rep} is strongly dependent upon -
  - (i) particle size and
  - (ii) the throat diameter to particle diameter ratio.

The exact nature of this dependence could be determined as extensive data is required. However, if the nature of material, particle size and throat opening is known the value of R can be corelated by a general equation shown below.

$$\log(W_s/G_g) = 9.9 \log N_{Rep} + C$$

the value of constant C being dependent upon the bed geometry, D_p, D_o and is calculated and shown in the following table.

D_p mm.	C		
	Calcite	Bauxite	Coal
0.648	-0.0415	-0.0426	-0.108
0.426	-0.0302	-0.01415	-0.170
0.287	-0.188	-0.1425	-

Note: The value of constant C seems to be independent of throat diameter for a given particle size and material.

CHAPTER - V

CONCLUSIONS & RECOMMENDATIONS :

The experimental studies on multistage counter-current fluidisation unit indicates that the solid flow is strongly dependent upon the gas flow rate, the flow properties of solids, the feeding mechanism and the apparatus geometry. As can be seen from the graphs relating to solid flow vs. gas flow rate, the smooth zone of solids flow is limited to a narrow zone and the initiation of solid flow takes place at a gas velocity slightly higher than minimum fluidization velocity. At very high gas velocities when the entrainment of solid becomes slowly predominent the column behaviour is usually erratic. Thus the designer has to necessarily choose gas velocity in this narrow zone where smooth column performance can be expected.

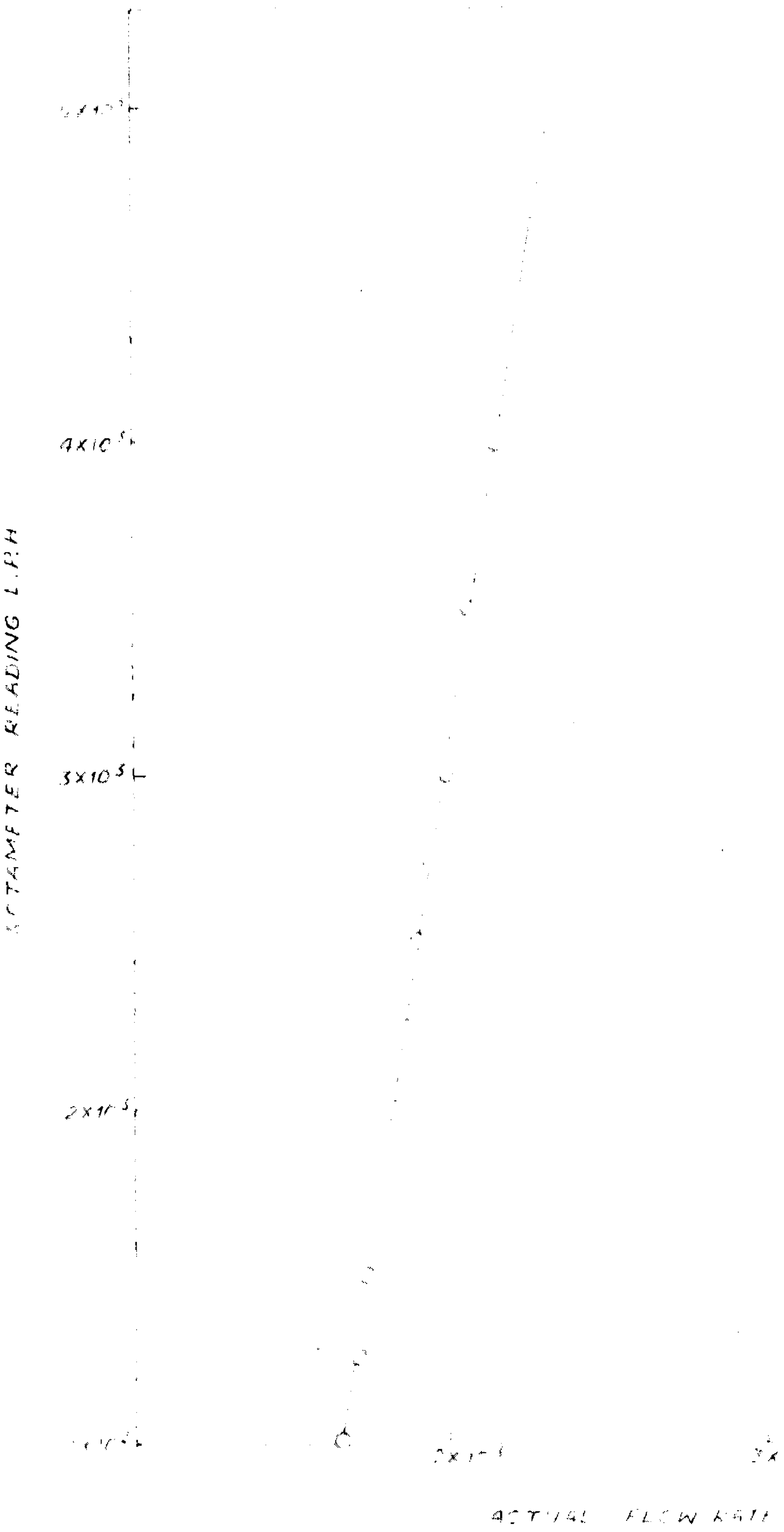
The column operation particularly at start up is rather tricky. In large scale industrial units this can be a major problem unless some care is taken. Based on limited experiments that have been conducted it can be safely predicted that the rate of solid flow in smooth zone of operation for a given material and system geometry is strong function of particle Reynolds number. However, to predict the flow rate for any material and any column's geometry large scale experimentation is necessary.

The results of present investigation are very much encouraging and suggests the need of more work.

Appendix - 1

ROTAMETER CALIBRATION :

Rotameter reading L.P.H.	Actual gas flow rate L.P.H.
1000	1650
1250	1710
1500	1735
1750	1780
2000	1825
2250	1850
2500	1900
2750	1935
3000	2000
3250	2020
3500	2050
3750	2100
4000	2140
4250	2180
4500	2210
4750	2250
5000	2300
5250	2340
5500	2380


5x10⁵
4x10⁵
3x10⁵
2x10⁵
1x10⁵
2x10⁵
3x
ROTAMETER READING L.P.H
ACTUAL FLOW RATE

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