COMPARISON OF A FLUIDIZED BED COMBUSTOR AND ITS SCALE MODEL

by

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ABSTRACT

To determine the validity of a set of proposed scaling laws to be used in the modeling of fluidized beds, a series of experiments were done on a full scale fluidized bed combustor operating at 1500°F and a ¼ scale model fluidized bed operating at 60°F. Minimum fluidization velocity, bed expansion, and pressure fluctuations were measured in both beds. The results were scaled appropriately and compared to determine the validity of the scaling laws.

A comparison of the minimum fluidization velocity and pressure fluctuation amplitude and frequency yielded correlations within 15%. Bed expansion for the hot bed, however, was found to be almost twice that of the cold bed. The cause for the discrepancy is not known. It is possible that the air flow rates in the hot bed were not measured or recorded correctly during the experiment. Since the cold bed does not model the combustion process, the discrepancy might be caused by inaccurate modeling.

The correlation of the minimum fluidization velocity and pressure fluctuation support the validity of the scaling laws, while the bed expansion data seems to contradict the laws. Further tests must be made to reach a final conclusion.

Thesis Supervisor: Leon Robert Glicksman, Ph.D.

Title: Senior Research Scientist

FOREWORD

This project compromises a part of the research of fluidized bed combustors at the Massachusetts Institute of Technology. Its purpose is to verify the validity of a set of scaling laws used to model fluidized beds. The laws were proposed by Dr. Leon Glicksman in September, 1978. This project consists of acquiring data from a fluidized bed combustor which is operated by the Chemical Engineering department, and comparing it to data obtained from a scale model, which was built by David Dalrymple according to the scaling laws of Dr. Glicksman.

The project was carried out in the MIT Mechanical Engineering department under the supervision of Dr. Leon Glicksman. In addition to Dr. Glicksman, I wish to express my gratitude to James Garcia, who provided considerable assistance and consultation. I also wish to thank Pete Walsh, who was extremely helpful and patient in providing me with information on the Chemical Engineering bed, and allowed me to run my experiments on it.

John J. Walsh

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INTRODUCTION

When a gas is passed through a bed of fine particles above a certain minimum velocity, the bed becomes fluidized and behaves in many respects like a liquid.

At very low velocities the gas simply moves through the voids between particles, without displacing the particles themselves. This is termed a fixed bed. As the velocity of the gas is increased the drag on the particles tends to offset their weight, and the bed becomes fluidized. As the velocity is increased further the bed begins to churn more violently, with large voids passing upward through the bed.

The first large-scale commercial use of a fluidized bed was by Fritz Winkler for coal gasification. The bed began operating in Germany in 1926. The next significant development came in 1940, when the United States foresaw a great demand for high octane airplane fuel in the near future. This application, called Thermofor Catalytic Cracking, used two fluidized beds, and enabled large scale production of aviation fuel. In addition, early fluidized beds had applications in carbonization of oil shale, calcination of limestone, and reduction of iron oxide.[1]

The great rise in petroleum prices in the last decade has led to much research aimed at developing new energy sources and finding more efficient ways of using the resources which are already available to us. This research has led to the use of fluidized bed in generating power, specifically as coal combustors. It has been found that when crushed coal is burned in a bed of limestone particles the NO_x and SO_2 emissions from combustion are far less than in conventional combustors. The reduced amount of pollutants eliminates the need for expensive, high maintenance stack scrubbers, which must presently be installed in coal burning plants. A fluidized bed combustor can burn inexpensive, high sulfur coal without exceeding federal emission standards.

Fluidized bed combustors (FBC's) have additional advantages over conventional burners. The rapid mixing of coal and particulate medium leads to nearly isothermal conditions in the bed, making the operation of the bed easily controllable. The heat transfer characteristics of FBC's are significantly better than in conventional combustors. The churning bed of hot solids and gas transfers more heat to an immersed heat exchanger tube than a typical burner would. In the FBC both conduction and convection are modes of heat transfer, whereas in a normal burner, where a flame comes in contact with the heat exchanger tube, only convection is present. The improved heat transfer characteristics of the FBC allow it to be run at about half the temperature of regular coal combustors.

Typically, a fluidized bed combustor consists of three main elements. The plenum is a box which sits below the bed and acts as an air reservoir. It helps to distribute the pressure evenly over the area of the bed. On top of the plenum sits the distributor plate, through which the air passes immediately before it is passed through the bed particles. The distributor plate serves to further distribute the flow. Once the flow has passed through the distributor plate it enters the bed and contacts the particles. Bed geometries vary greatly, but typically a bed is a circular or square column. The height of the column is usually several times higher than the bed of particles.

During operation coal is injected at one or more feed points. The crushed coal ignites in the hot bed, mixing with the particulate medium. The hot churning mixture transfers heat to heat exchanger tubes which are submerged in the bubbling bed. Constant temperature is maintained by controlling the coal and air feed rates.

Although fluidized bed combustors have many advantages over conventional combustors, there is little information available to engineers enabling them to optimize FBC design. Since analytical solutions are too complex to be practical, full size beds must be constructed and tested at normal operating conditions. Making design improvements on full size beds has proved costly and time consuming. The fact that the tests have to be run at high temperatures make the design procedure particularly slow.

Dr. Leon Glicksman of the Massachusetts Institute of Technology has proposed a set of scaling laws that would enable a designer to do his tests on a small model of an FBC at ambient temperature. Changing the configuration of the scale model would be a simple process, and so the design procedure would be greatly improved.

The scale model "cold" bed would primarily aid in optimizing the fluid mechanical behavior of the bed. Bed geometry, particle size and density, and air velocity all greatly influence the fluid mechanics in the bed. Any of these variables could be easily modified using a model bed that has been constructed within the framework of Dr. Glicksman's scaling laws. Once the scaling laws have been verified, they can be used to constuct small beds operating at ambient temperature which accurately model the fluid mechanical behavior of full scale hot beds.

SCALING LAWS

What follows is a brief summary of the scaling laws proposed by Dr. Glicksman. For a more detailed treatment see Reference 3.

The proposes scaling laws are found by non-dimensionalizing the equations which govern the fluid mechanics of fluidized beds. In general these equations are too complex to solve analytically, but the non-dimensionalization provides the necessary information to formulate the scaling laws.

The equations of motion and conservation of mass were written for both fluid and particles. For the gas, only gas-particles forces and pressure forces were considered. Particle-particle forces were neglected. Also, the gas was assumed incompressible.

From these equations the various non-dimensional groups were isolated. Both the viscous and inertial limits were considered and included in determining which groups were important. This insures validity for a wide range of Reynolds' numbers. The non-dimensional groups which must be conserved in the modeling process were found to be:

1) $\dfrac{\rho_s \rho_g d_p^3 g}{\mu^2}$

2) $\dfrac{g d_p}{u_0^2}$

3) $\dfrac{\rho_g}{\rho_s}$

4) $\frac{L}{d_p}$

5) $\frac{D}{d_p}$

6) ϕ_s

7) Particle size distribution

8) Bed geometry

where ρ_s = density of solids

ρ_g = density of gas

d_p = particle diameter

g = acceleration due to gravity

μ = viscosity of gas

u_o = superficial velocity of gas

L = length of bed

D = depth of bed

ϕ_s = sphericity of solids.

Since the density and viscosity of air are functions of temperature, any set of scaling factors will also be dependent upon temperature. Table 1 is a summary of scaling factors to convert from a bed operating at $1500^{\circ}F$ to one operating at $60^{\circ}F$. Also, all the cold bed geometries, including any internals, must be scaled down by a factor of four.

Since bubble frequency and pressure fluctuations in the two beds will be compared, it is necessary to know the scaling factors for time and pressure. Using the scaling factors presented in Table 1, the following relations can be arrived at:

$$t_{60} = 2t_{1500}$$
$$P_{60} = .875P_{1500}.$$

where t_n = time @ n^oF

P_n = pressure @ n^oF.

TABLE 1

Scaling Factors for Proper Modeling of Hot Bed Performance

Hot Bed Variable (1500^oF)	Scaled Cold Bed Variable (60^oF)
u_{o1500}	$0.5\ u_{o1500}$
$d_{p\ 1500}$	$0.25\ d_{p\ 1500}$
$\rho_{s\ 1500}$	$3.5\ \rho_{s\ 1500}$
L_{1500}	$0.25\ L_{1500}$
D_{1500}	$0.25\ D_{1500}$

SCALE MODEL BED DESCRIPTION

The scale model cold bed was built by David Dalrymple to replicate the 2' x 2' hot bed operated by the Chemical Engineering department. The cold bed is a geometric copy of the hot bed, with all dimensions scaled down by a factor of four (Figure 1). Its internal cross section measures 6" x 6" and its height is 4 feet. The front wall of the bed is constructed of ½" plexiglass. The other walls are made from ¼" masonite glued to a ½" plywood frame. The joints between walls are made airtight by a bead of glue running the length of the bed. Pressure taps, made from 5/32" nylon tubing, were installed at various heights up the bed. Their exact positions are tabulated in Appendix D.

The distributor plate was made from a 9" square plate of 3/4" aluminum. It was drilled with an 8 by 8 pattern of .154" diameter holes. Each hole was fitted with a tuyere, a piece of nylon tubing with four holes drilled near the sealed end. The air flow enters the bottom of the tuyere and leaves near the top in four equally spaced jets at a horizontal angle.

The plenum is a simple box, constructed of ½" plywood and held together with glue and wood screws. The inside is 6" square and 8" high. The air supply enters the bottom of the plenum through a ½" pipe.

The air is supplied by a shop compressor at 90 psig. The air flow rate is calculated using a square edged orifice plate, which is located in the inlet pipe. The pressure drop across the orifice, measured by a u-tube manometer, is related to the flow rate. For the range of superficial velocities that we will be using (.2 ft/sec-- 2.0 ft/sec) the following equation applies:

$$u_o = .23 \times (\Delta P_o)^{\frac{1}{2}}$$

where u_o = superficial velocity of air (ft/sec)

ΔP_o = pressure drop across orifice (in. H_2O).[2]

The cold bed was constructed to model as closely as possible the hot bed from which data is available. One aspect of the hot bed which was not modeled is the coal feeder. Coal is injected into the hot bed through an air line. The input position is about 2" above the tuyere orifices. Typically, the mass flow rate of the air from the coal feeder compromises less than 10% of the total mass flow rate of air. Thus, the effects of the coal feed air can be neglected.

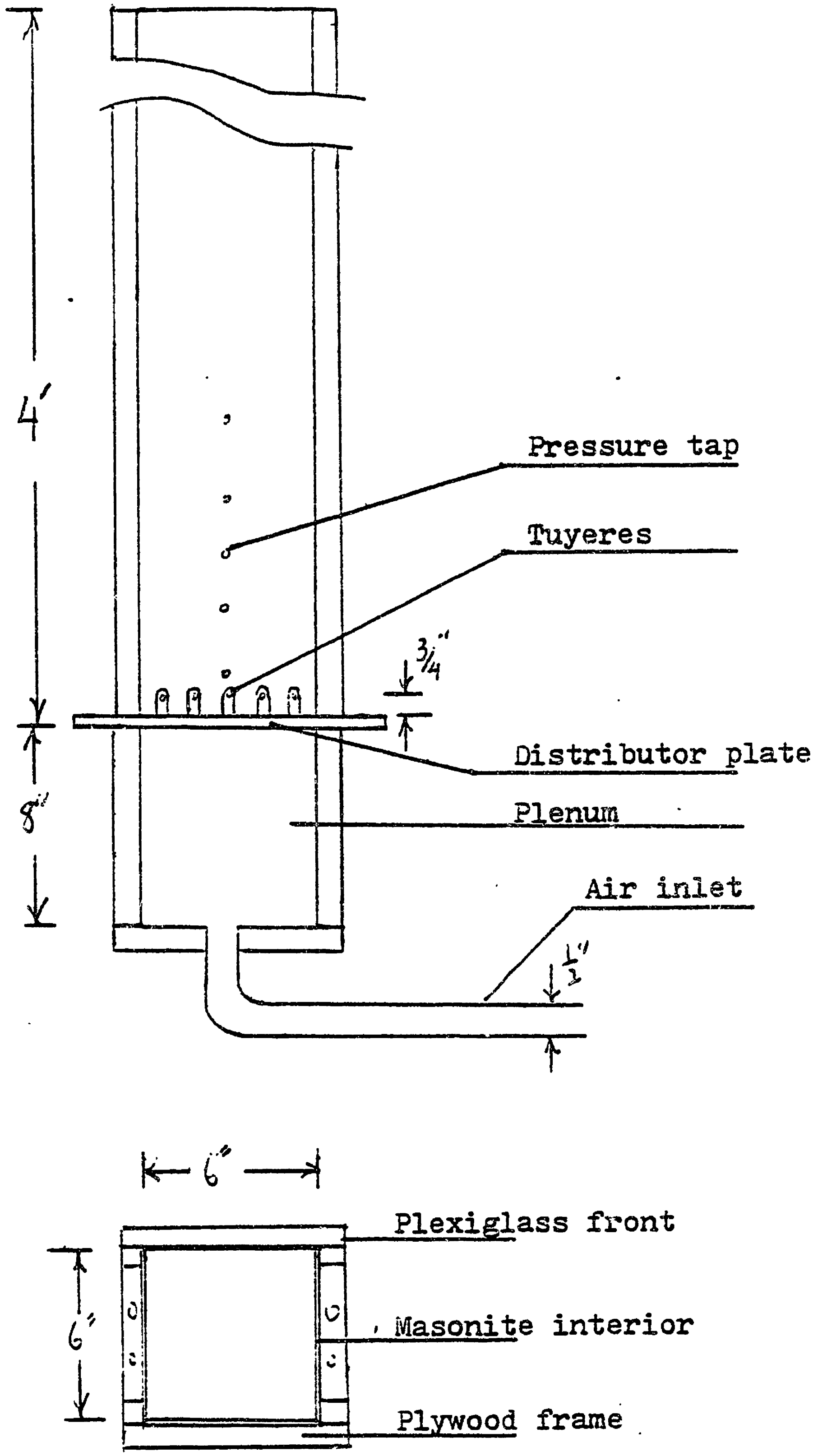


Figure 1. Scaled Bed Dimensions

EXPERIMENTS

In order to verify the proposed scaling laws, measurements of minimum fluidization velocity, bed expansion, and transient pressure fluctuations were made on both the hot and cold beds. For the pressure fluctuations, both frequency and amplitude were considered.

Each of these variables plays an important role in the fluid mechanics of a fluidized bed. Each of them is dependent upon other properties of the bed, and therefore the group as a whole represents a rigorous testing of the proposed scaling laws. In addition, each of these properties can be measured in both the hot and cold beds.

Once the data has been collected, the values of the cold bed are scaled to the hot bed data for comparison. If the values for the two beds correlate, the proposed scaling laws can be regarded as valid.

Minimum fluidization velocity

The pressure at the base of the solids rises linearly with superficial velocity up to a certain point. The pressure at this point reaches a maximum and thereafter remains constant. This maximum pressure is equal to the weight of the solids divided by the cross-sectional area of the bed. In other words, the drag forces acting on the particles are just balanced by their weight. A bed in this condition is said to be fluidized. The minimum superficial velocity for which the bed is fluidized is called the minimum fluidization velocity, u_{mf}.

The minimum fluidization velocity can be calculated using a modified form of Ergun's relation:

$$\frac{d_p\, u_{mf}\, g}{\mu} = \left[33.7^2 + 0.0408\, \frac{d_p^3 \rho_g (\rho_s - \rho_g)\, g}{2}\right]^{\frac{1}{2}} - 33.7 \quad (1)$$

To measure the minimum fluidization velocity, the maximum static pressure in the bed is measured as a function of the superficial velocity. This curve will be linear with a positive slope until u_{mf} is reached. Thereafter the maximum static pressure will remain constant. The value of u_{mf} can be read from the point where the pressure curve first becomes flat.

Bed expansion

The percent increase in bed height over the bed height at minimum fluidization is called the bed expansion.

$$\%\ \text{Expansion} = \frac{h_o - h_{mf}}{h_{mf}} \quad (2)$$

where h_o = bed height for a certain superficial velocity

h_{mf} = bed height at minimum fluidization.

As the superficial velocity is increased the bed expands in height as the voids between the suspended particles grow. At low velocities, when little bubbling occurs, the bed height can simply be measured with a ruler. At higher velocities, however, large bubbles form which burst upon reaching the top of the bed. This causes large displacements of the bed surface, making it impossible to measure the height of the bed.

The bed height of a bubbling bed can be accurately measured by observing the pressure distribution within the bed. Normally, the pressure is a linear (or nearly linear) function of height. The pressure is greatest at the bottom of the bed and rises linearly until it reaches the bed surface (freeboard). Above the freeboard the pressure remains constant. Thus, if the pressure distribution in the bed is known, the height of the bed is simply the level where the pressure first reaches freeboard pressure. In general a pressure tap will not be located at the exact bed height. In this case, the pressure curve must be extrapolated to the freeboard pressure in order to obtain the height of the bed.

All static pressure readings taken from the bed were made with a set of nine water manometers, each having a range of 50 inches.

Pressure fluctuations

In a fully fluidized bed, bubbles of gas form and rise to the bed surface where they burst resulting in a large displacement of particles. A knowledge of bubble behavior is essential to predict accurately the mixing properties of the bed.

The bubble frequency can be measured using a pressure transducer. As a bubble passes upward through any point in the bed, the pressure at that point rises as the bubble goes by. When a pressure transducer is used to convert the pressure signal into an electronic signal, the pressure fluctuations can be viewed on an oscilloscope. The result is a series of peaks and valleys, caused by a stream of bubbles passing by the point of measurement. An oscilloscope camera is used to take photographs of the fluctuations. From these photographs the bubble frequency and fluctuation amplitude are determined.

For these experiments measurements of the pressure fluctuations were taken at the bed wall at two different heights (12 in. and 27.5 in. for the hot bed) at a single superficial velocity (2.7 ft/sec for the hot bed). The parameter values for the cold bed were scaled accordingly.

A linear pressure transducer (Statham UC3) was used in conjunction with a transducer readout (Statham UR5) and an oscilloscope (Hewlett-Packard 130C). The system was calibrated such that 1 volt on the oscilloscope represented a pressure of 13.33 in. H_2O.

Particle density and size distribution

To achieve the 3.5 ratio of particle densities, steel particles were used in the cold bed. The density of steel is slightly higher than $3\frac{1}{2}$ times the density of sand, the medium for the hot bed.

The average particle diameter, $\bar{d}_p$, for the cold bed should be $\frac{1}{4}$ of that of the hot bed particles. Since the hot bed was being run with different particle sizes each time, the particle size distribution could not be met exactly, because it was not known ahead of time what the particle size in the hot bed would be when the data was obtained. A typical hot bed particle has a diameter of .8 mm, so the cold bed tests were run with a distribution of particles averaging .200 mm in diameter. This mix was prepared using the formula

$$\bar{d}_p = \frac{1}{\sum x_i/d_{pi}} \tag{3}$$

where x_i = mass fraction of ith particle group

d_{pi} = diameter of particles in the ith group.

The hot bed data was actually obtained with particles averaging .74 mm in diameter, thus the particle sizes were not scaled exactly as they should have. Particle distributions for both beds are shown in Table 2.

TABLE 2

PARTICLE SIZE DISTRIBUTIONS

Hot Bed	d_{pi} (mm)	x_i (%)
	2.200	0.7
	1.705	0.5
	1.205	21.4
	0.855	48.1
	0.605	19.6
	0.427	6.0
	0.250	3.7

$\bar{d}_p = .736$ mm

Cold Bed		
	0.213	21.0
	0.163	79.0

$\bar{d}_p = .200$ mm

RESULTS

Minimum Fluidization Velocity

Cold Bed

The minimum fluidization velocity for the cold bed was found by constructing a plot of pressure at tap #1 vs. superficial velocity (Figure 2). From this graph u_{mf} is seen to be .36 ft/sec for the cold bed.

Equation 1 predicts a value of .33 ft/sec. The 9% error is well within the 34% standard deviation of Equation 1.

Hot Bed

Obtaining the minimum fluidization velocity for the hot bed was difficult. Typically, the hot bed is run at one superficial velocity. When this velocity is changed, the bed can take several hours to reach a steady state. Changing u_o also changes the temperature in the bed, thus altering the results. To obtain the necessary data it was decided that the best method would be to lower the air flow rate quickly, and then take the pressure readings before the temperature changed very much.

The data for this experiment appears in Appendix B. Observing the overall pressure drop across the bed, one can see that the bed never defluidized. According to the air flow rates recorded, a minimum value for the superficial velocity of .3 ft/sec was reached. Both Ergun's relation and the proposed scaling laws indicate that at

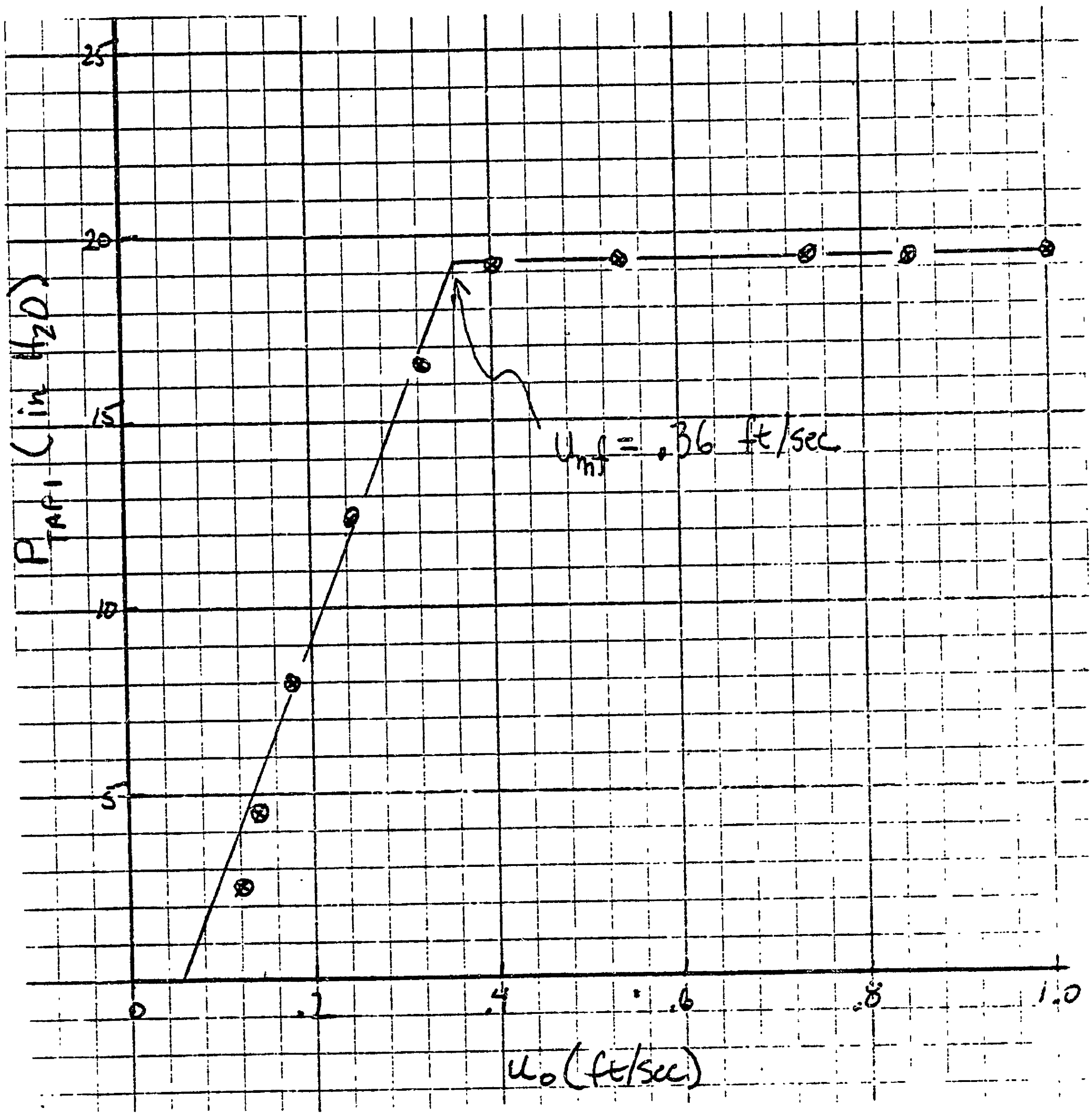


Figure 2. Pressure vs. Superficial Velocity--Cold Bed

.3 ft/sec the bed should not be fluidized, and there should be a significant drop in pressure across the bed at that velocity. Equation 1 predicts a minimum fluidization velocity of .66 ft/sec, while the cold bed performance indicates that u_{mf} for the hot bed should equal .72 ft/sec.

Since additional hot bed data was not available in time for this report, it was decided to calculate the minimum fluidization velocity for the large bed at ambient temperature. The value for u_{mf} for this experiment was found to be 1.25 ft/sec. It is possible, using Equation 1, to adjust this value to correspond to the hot conditions. At low Reynolds' numbers, Equation 1 can be written as follows:

$$u_{mf} = \frac{d_p^2\,(\rho_s - \rho_g)\,g}{1650\,\mu} \qquad (4)$$

Since the density of the gas is negligible compared to that of the solid, the only parameter which changes with temperature is the gas viscosity. At 1500°F the viscosity of air is twice that of air at 60°F. Thus, u_{mf} at 1500°F will be ½ of the value at 60°F. This yields a value of .63 ft/sec for the minimum fluidization velocity of the hot bed. This value correlates closely with the value obtained in the cold bed. The values would have been even closer had the particle diameters been scaled more precisely.

The cause for the error in the data taken from the hot bed is thought to be improper reading of the air flow rates.

This would cause the superficial velocities to be calculated incorrectly, making it impossible to locate the minimum fluidization velocity.

Bed Expansion

Cold Bed

The bed expansion for the cold bed is plotted against the ratio of the superficial velocity to the minimum fluidization velocity, u_o/u_{mf}, in Figure 4. The values for bed height were obtained by drawing the pressure distribution in the bed, and extrapolating the curve to the freeboard pressure, (zero, in the case of the cold bed). Three typical pressure distributions are shown in Figure 5.

Hot Bed

The problems with acquiring hot bed data were discussed above. To construct a curve for bed expansion, the pressure distributions at several superficial velocities are needed. For the data taken from the hot bed, the pressure readings indicate a fully fluidized bed. We are assuming that the first few values of u_o in the hot bed data are accurate, since the air flow rates for these runs were still fairly high. So we have several distributions at relatively high superficial velocities. Two of these pressure distributions are shown in Figure 6.

The bed height at minimum fluidization was taken from the pressure distibution at u_{mf} for the large bed run at

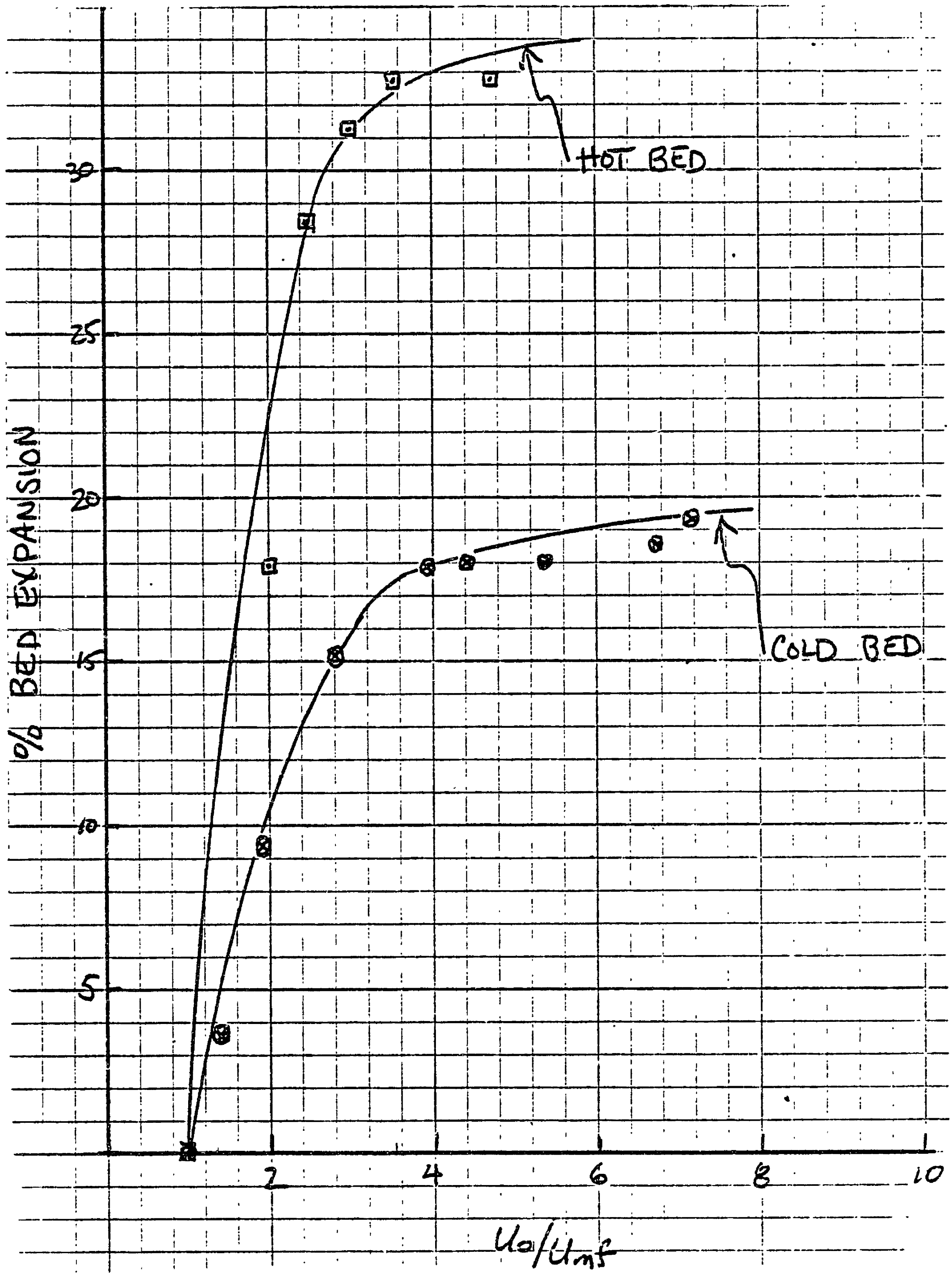


Figure 4. Plots of Bed Expansion for Cold and Hot Beds

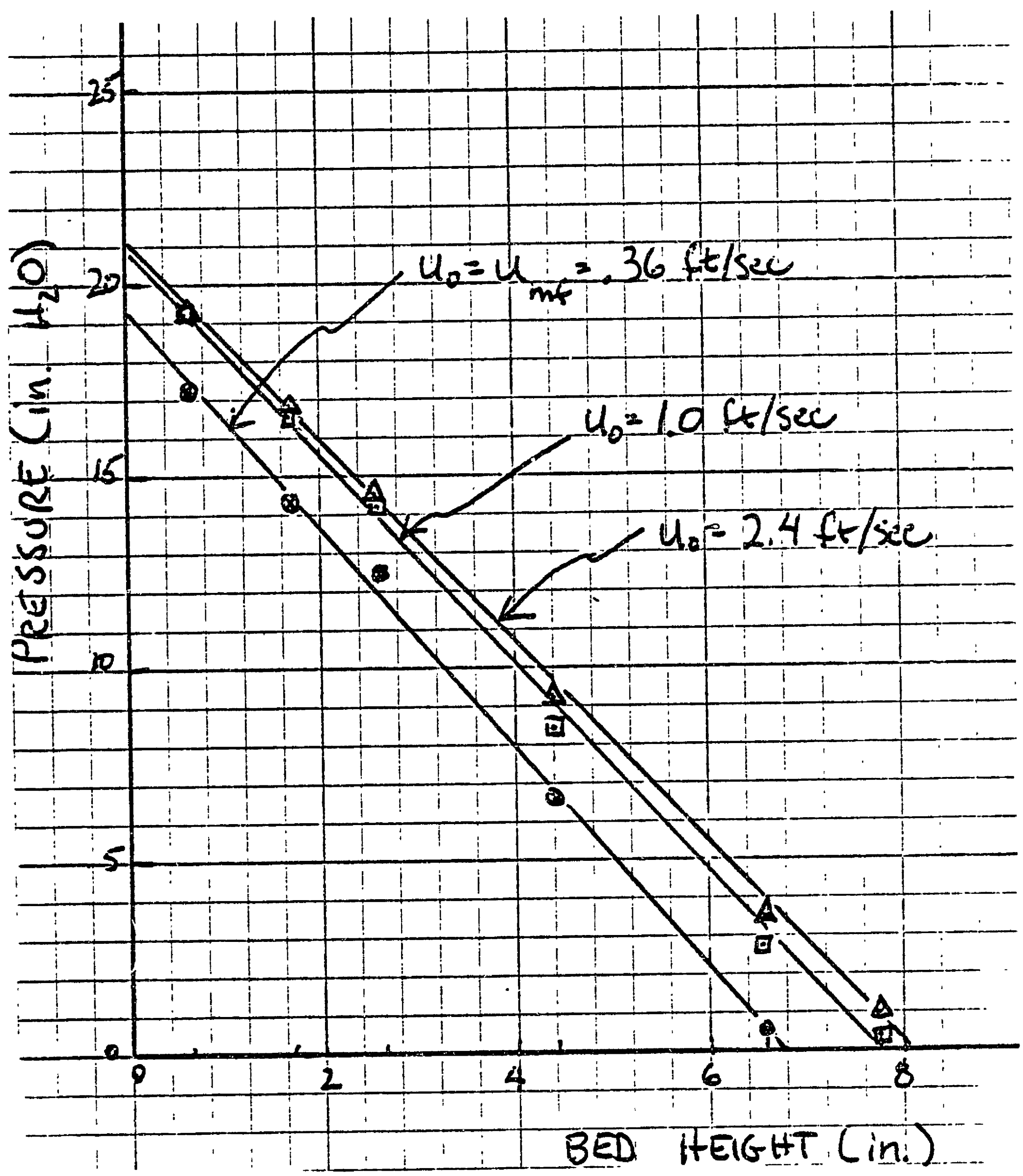


Figure 5. Pressure Distributions from Cold Bed

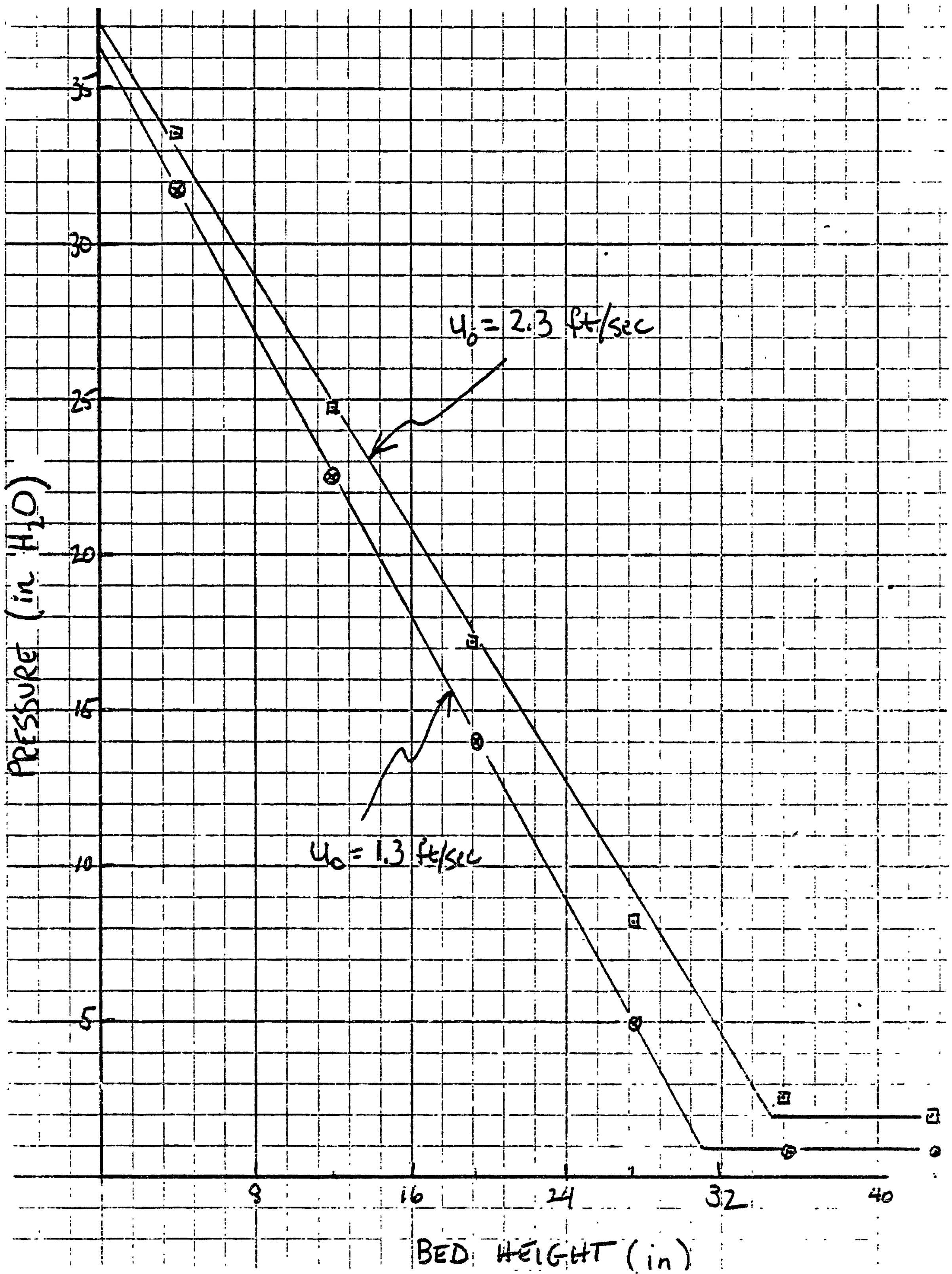


Figure 6. Pressure Distributions from Hot Bed

ambient temperature. Figure 7 shows that the bed height at minimum fluidization is 26.4 inches. From this data, the bed expansion was plotted and appears in Figure 4 with the cold bed expansion curve.

As can be seen there is a large difference between the two beds. The hot bed expanded almost twice as much at similar velocities. The reason for the large discrepancy is not known. It is possible that the data that was used to calculate the values of expansion for the hot bed was not valid. Also, the expansion in the cold bed, which never exceeded 20%,seems too low. Further tests must be run, and accurate data from the hot bed must be obtained, to either eliminate the difference between the two curves, or to explain why the difference exists.

Pressure Fluctuations

Photographs of the pressure fluctuations are shown in Figures 8 and 9. These measurements were taken at heights of 12" and 27.5" in the hot bed, and correspondingly at 3 3/4" and 6" in the cold bed (exact scaling could not be made for the heights due to the location of the pressure taps). The superficial velocity in the hot bed was 2.7 ft/sec, and in the cold bed u_o was 1.3 ft/sec.

Qualitatively, the signals look very similar. A steady stream of peaks and valleys was observed in both beds. There were few cases of flat spots of constant pressure. The continuous fluctuations indicate large number of bubbles passing by the point of measurement.

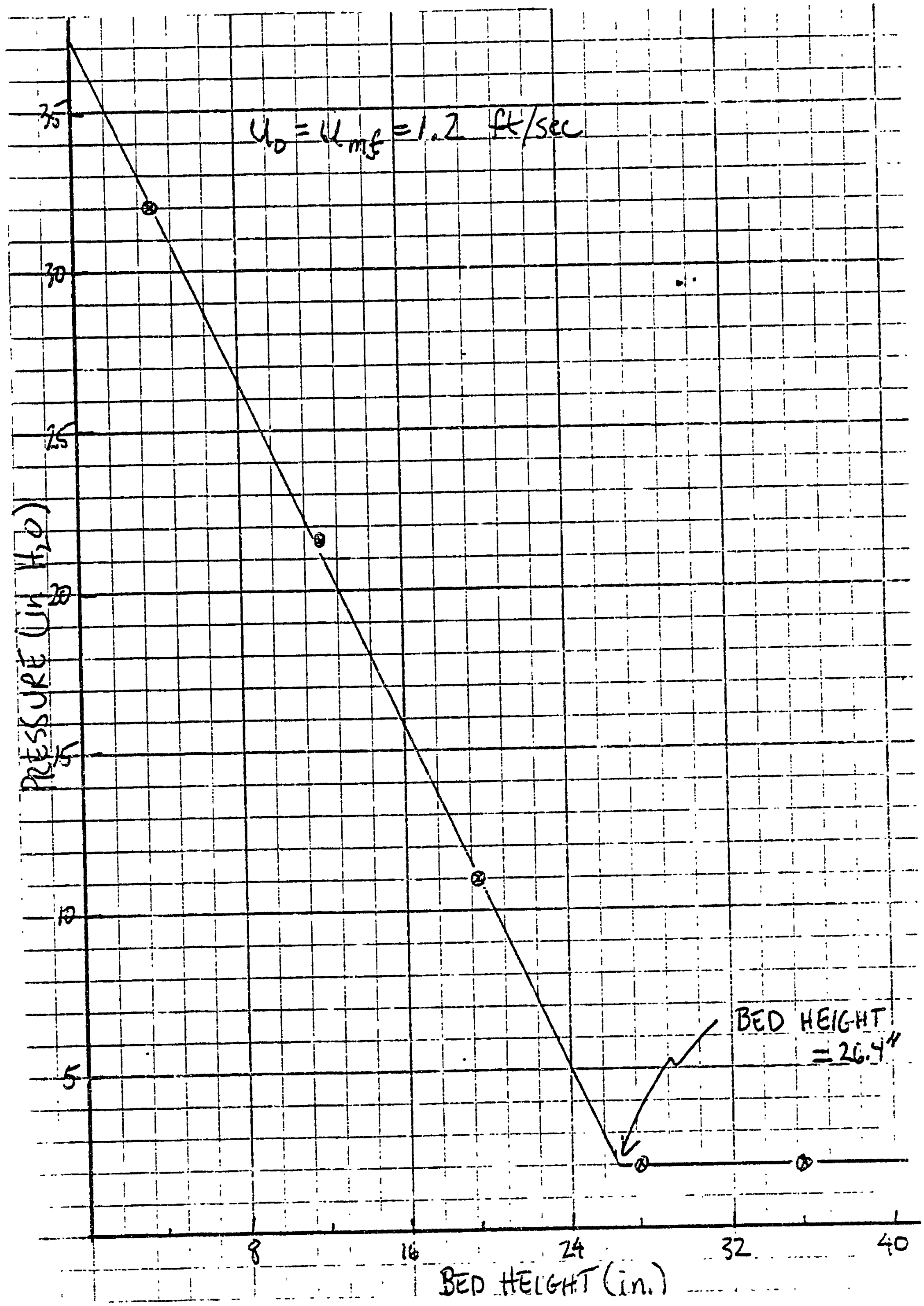
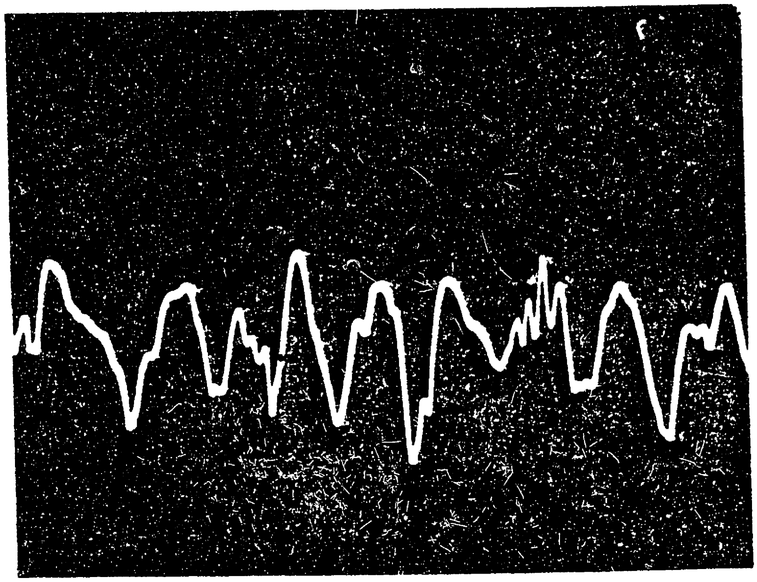
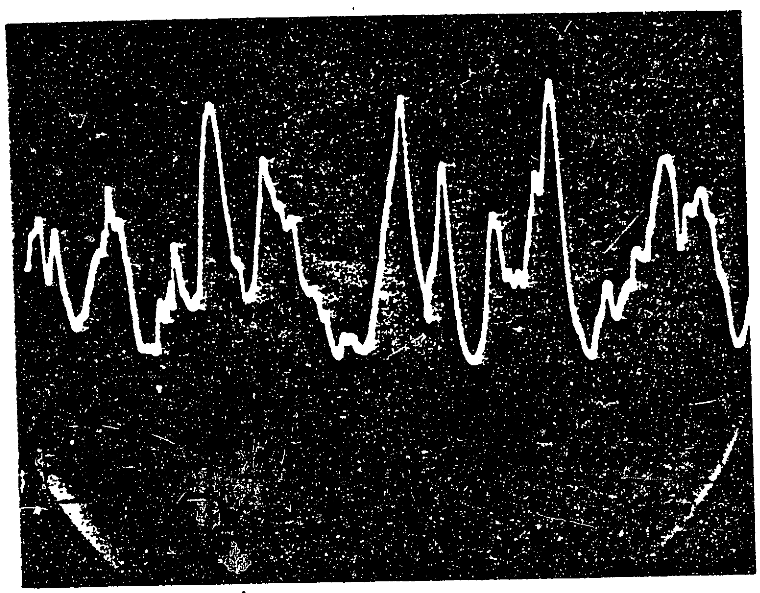


Figure 7. Pressure Distribution @ u_{mf}--Large Bed @ Amb. Temp.

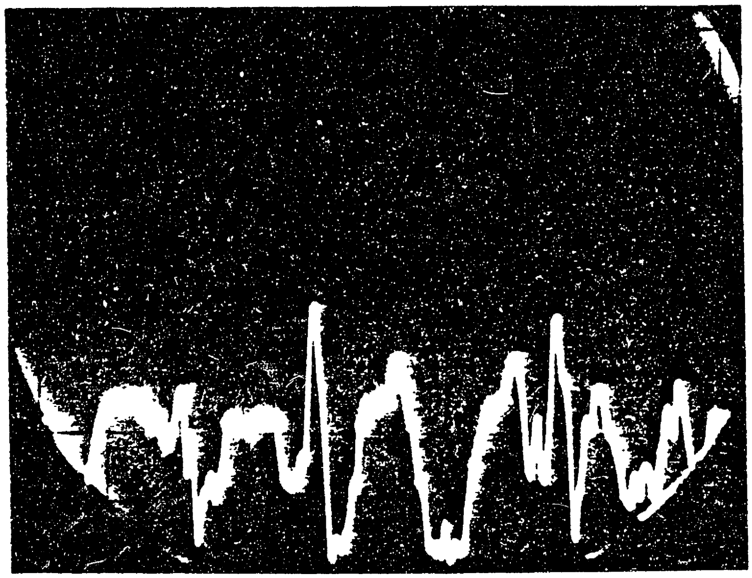


8a. Cold Bed; @ 3 3/4"; Scale: 6.7 in H_2O/cm, .2 sec/cm.

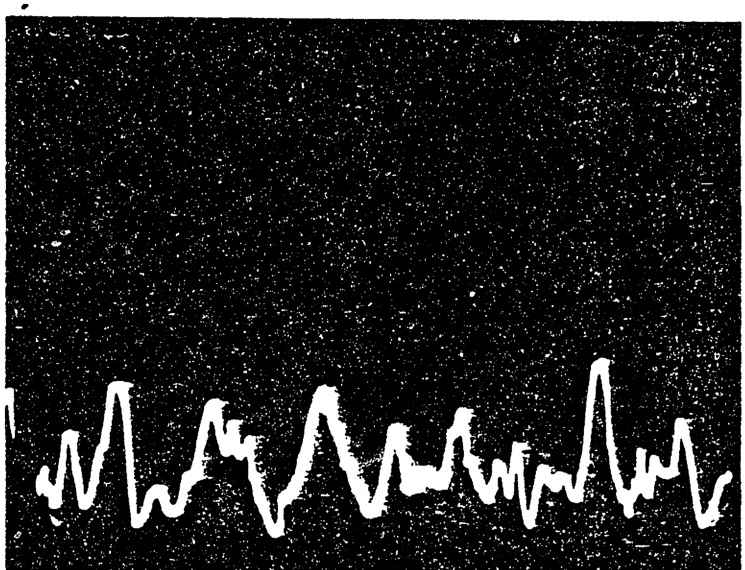


8b. Hot Bed; @ 12"; Scale: 6.7 in H_2O/cm, .5sec/cm.

Figure 8. Pressure Fluctuations at Lower Level.



9a. Cold Bed; @ 6"; Scale: 2.7 in H_2O/cm, .2 sec/cm.



9b. Hot Bed; @ 27.5"; Scale: 6.7 in H_2O/cm, .5 sec/cm.

Figure 9. Pressure Fluctuations at Upper Level

The scaling laws predict the bubble frequency in the hot bed to be $\frac{1}{2}$ of the bubble frequency in the cold bed. Counting the major peaks of pressure fluctuation, we find the frequency of the bubbles in the cold bed to be $4\frac{1}{2}$ hz. For the hot bed the frequency is slightly greater than 2 hz. This correlates quite closely with the scaling law prediction.

The amplitude of the pressure fluctuations in the cold bed should be 87% of the magnitude of the hot bed fluctuations. At a height of 12" in the hot bed the maximum fluctuations have a magnitude of 26.7 in. H_2O. The cold bed, at the correponding height, showed a maximum fluctuation of approximately 20 in. H_2O, or 75% of the hot bed value. At the higher level, the magnitude of fluctuations in the hot bed was 13.3 in. H_2O. The cold bed at the upper level showed maximum fluctuation magnitudes of 10.6 in. H_2O, which is 80% of the hot bed value. The values for the magnitude of the pressure fluctuations in the cold bed are slightly lower than predicted, but the values are close enough to be considered significant.

CONCLUSION

Comparisons of hot and cold bed results for minimum fluidization velocity and pressure fluctuations indicate that the proposed scaling laws are valid for the modeling of fluidized beds. The large discrepancy in the bed expansion results, however, suggest either some error in the scaling laws or an error in collecting the data, particularly the hot bed data.

Further experiments need to be done on the hot bed to obtain an accurate value for the minimum fluidization velocity, and accurate measurements of the bed expansion. If possible, the tests should be repeated several times under the same conditions. Varying the conditions in the hot bed, and correspondingly in the cold bed, and then comparing the results would be a good test of the scaling laws.

On the basis of the experiments performed thus far, the scaling laws cannot be regarded as absolutely valid, although enough correlation between hot bed and cold bed data exists to justify further testing of both beds.

APPENDIX A

Pressure Data for Cold Bed

Height of fixed bed--6 13/16"

	Pressure @ Tap # (in H_2O)					
u_o (ft/sec)	1	2	3	4	5	6
.36	17.8	14.9	12.7	6.7	0.6	0.0
.42	19.3	16.2	13.1	6.8	0.5	0.0
.50	19.3	16.1	14.0	7.3	1.1	0.0
.70	19.3	16.5	13.2	8.3	2.2	0.0
1.0	19.3	16.7	14.2	8.5	2.8	0.3
1.4	19.3	16.6	14.3	8.6	3.4	0.7
1.6	19.3	16.7	14.5	8.7	3.4	0.7
1.9	19.2	16.7	14.3	8.7	3.3	0.8
2.4	19.2	16.8	14.4	8.7	3.4	0.8
2.5	19.3	16.8	14.3	8.7	3.6	1.0

APPENDIX B

Pressure Data for Hot Bed @ 750°C.

Height of fixed bed--unknown

	Pressure @ Tap # (in H_2O)					
u_o(ft/sec)	1	2	3	4	5	6
2.3	33.6	24.8	17.3	8.2	2.6	2.0
2.0	33.0	23.6	16.4	7.0	1.6	1.4
1.6	32.0	22.6	14.6	5.2	0.8	0.6
1.3	31.8	22.6	14.0	5.0	0.8	0.8
1.0	30.8	21.6	13.0	3.6	0.4	0.4
0.7	29.0	19.8	11.2	1.8	0.2	0.2
0.5	29.0	19.8	10.8	1.0	0.0	0.0
0.3	28.6	18.8	9.8	0.0	0.0	0.0

Pressure Data for Hot Bed @ Ambient Temperature

0.46	11.6	7.8	4.8	0.4	0.4	0.4
0.62	15.4	10.4	5.4	0.4	0.4	0.4
0.95	24.6	16.8	8.8	1.2	1.2	1.2
1.21	32.0	21.6	11.0	2.0	2.0	2.0
1.48	33.8	23.6	13.0	4.2	3.0	3.0
1.84	36.4	26.8	15.8	8.6	5.0	5.0
2.30	38.4	28.8	20.6	11.4	6.8	6.8
2.62	40.8	32.0	22.8	14.6	9.4	9.4
2.95	44.0	35.2	26.0	18.6	12.6	12.6

APPENDIX C

Cold Bed Data--Maximum Static Pressure for Various Superficial Velocities

u_o (ft/sec)	Max. Pressure (in H_2O)
0.73	19.3
0.55	19.2
0.40	19.1
0.36	17.8
0.34	16.5
0.31	15.0
0.25	12.3
0.18	7.9
0.15	4.5
0.13	2.5

APPENDIX D

Height of Pressure Taps in Cold and Hot Beds

Tap #	Hot Bed	Cold Bed
1	3.94 in	5/8 in
2	11.81	1 11/16
3	19.69	2 5/8
4	27.56	4 7/16
5	35.43	6 5/8
6	43.31	7 3/4
7	51.18	8 5/8

REFERENCES

1. Dalrymple, D., "Testing of a Scale Model of a Fluidized Bed Combustor," Bachelor's Thesis, MIT, 1979.

2. Kunii, D. and Levenspiel, O., Fluidized Engineering, New York, 1969.

3. Glicksman, L., "Proposal: Scale-Up of Fluidized Beds," MIT, 1978.

NOTES

1. Kunii and Levenspiel, pp. 16-59.
2. Dalrymple, pp. 13-17.