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https://doi.org/10.4224/21276133

NATIONAL RESEARCH COUNCIL OF CANADA
DIVISION OF APPLIED CHEMISTRY

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Issued as N.R.C. No. 7941

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ABSTRACT

An experimental study using fixed and fluidized beds in the isomerization of cyclopropane on a silica alumina catalyst is reported for a temperature range 150 to 250°C and of 5 to 150 W/F (g-catalyst, hr/g-mole). The effects of various cylindrical screen packing, 0.2-, 0.4- and 1.0-in. Dia., and 1.0-in. Dia. Pall ring, on final conversion were determined. Reactor scale effects were also considered using reactors of 0.9-, 1.8- and 6.0-in. Dia., and bed heights 1.8- to 11.0-in.

Packing in a fluidized bed improved gas-catalyst contact and resulted in higher over-all conversions when compared to a normal fluidized bed. With properly designed packing, the final conversion closely approached the fixed bed conversion. When catalyst was tested in a fixed bed no scale, film diffusion or significant packing effects were observed. Rate data closely followed first order kinetics. When the same catalyst was tested in a normal fluidized bed, the conversion vs. dimensionless group kW/F was dependent on linear gas velocity and catalyst bed height. With packing present in the fluidized bed this dependency was much less. Packing size and shape, however, had some effect on the x vs. kW/F relationship.

Several models proposed by previous investigators are considered for correlating the data of fluidized bed. Some of these would be useful in designing a fluidized bed reactor containing packing.

INTRODUCTION

The fluidization technique has become a widely used method of solid-gas contact over the last twenty years as a consequence of a large amount of research and development effort in both industry and universities. The quantitative design of fluidized bed reactors is, however, complicated by the by-passing of the gas via bubbles and channels, with the result that in many reactions lower conversions are obtained in a fluidized bed than in a fixed bed at the same space velocity.

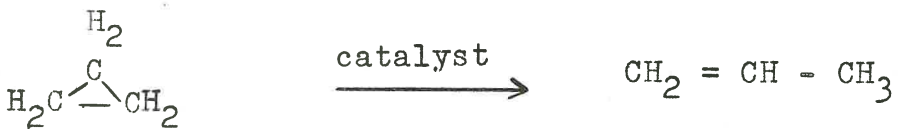
Consequently, many modified reactors have been proposed to control the nature of the gas and the solid flow in the fluidized bed. For example baffles, various packings or stirring devices have been suggested to break a deep bed into a series of shallow beds or to obtain a uniform dispersion of small gas bubbles (1, 2, 3 and 4). Overcashier et al. (5), Lewis et al. (6) and Massimilla et al. (7) have found that horizontal baffles improve the homogeneity of fluidized solid-gas systems and this effect was most pronounced at high gas flow rates. Volk et al. (21) have used vertical surfaces such as tubes or half-rounds successfully. They report that the conversion rates were the same in a 6.5 ft. D. reactor as were obtained in 6-in. D. reactor. Very little quantitative information regarding the effect of fixed packing on conversions in a fluidized bed reactor has been published. Sutherland et al. (8) as a result of a study of the effect of packing on fluidized bed properties such as minimum fluidization velocity, pressure drop and heat transfer suggested that cylindrical screen packing should have useful applications in fluidized beds of free flowing materials.

The work described in this paper is an investigation of the effect of perforated wall packing on a simple first order catalytic reaction. Kinetic data between the fixed bed (with and without packing) and the fluidized bed (with and without packing) were obtained so that direct comparisons could be made.

EXPERIMENTAL PROCEDURE

Reaction Used

The reaction chosen for this study was the catalytic isomerization of cyclopropane to propylene,



The rate of isomerization was measured in the temperature range from 150° to 250°C. Inlet stream composition to the catalyst beds averaged about 2% cyclopropane and 98% air.

Roberts (9) studied the isomerization of cyclopropane using acidic solid catalysts. On silica-zirconia-alumina the apparent activation energy of isomerization was 19.0 Kcal. in the temperature range of 100 to 135°C. The most important side reaction was the dimerization of propylene. The material retained by the catalysts was probably largely propylene and polymers of propylene. Kinetics of the reaction was investigated using gas chromatographic methods by Bassett and Habgood (10). The reaction was first order at 370°C on Na-Zeolite (Linde Molecular Sieve 13X). The over-all activation energy was 19.0 Kcal. in the temperature range 263 to 378.5°C. No

hydrocarbons other than propylene were observed in the reaction products. This investigation also suggested that the presence of water or available protons is essential to the formation of a reaction intermediate in the isomerization of cyclopropane. Studies by Davis and Scott (11) have indicated that this isomerization behaves as an ideal homogeneous first-order reaction in contact with borosilicate glass equipment at constant pressure and at temperatures up to 620°C.

Catalyst

The catalyst used in the studies was the 150 to 200 Tyler mesh fraction of an equilibrated silica-alumina cracking catalyst. The bulk density of catalyst containing 2.2% water was 0.87 g/cc. A particle density of catalyst dried at 105°C and 200°C was 2.13 and 2.25 g/cc., respectively. The catalyst particles were generally round but not truly spherical. The catalyst contained 2.2% water when charged to the reactor.

Apparatus

A flow diagram of the experimental system is shown in Figure 1. Cyclopropane of 99.5% purity was employed without further purification. Compressed air from a Nash-Hytor compressor was employed without drying or purification so that air contained some water vapour. Individual gases were metered by flow meter, and mixed to about 2% cyclopropane. The combined feed stream was preheated before entering the catalyst bed. The experiments for both fixed bed and fixed bed with packing were carried out using downward flow of the gas. For the fluidized bed runs upward flow was used. The kinetic data were obtained using three reactors, A, B and C. Details of the reactors are shown in Figure 2.

Reactors

Reactor A consisted of a 16-in. length of 0.89 in. I.D. Pyrex tubing surrounded by a heated oil jacket to maintain the desired reaction temperature. The oil in the jacket was heated electrically, and its temperature was controlled automatically to about 1 to 2°C. A glass porous plate in the bottom of the tube supported the catalyst bed and distributed the gas uniformly. Temperature measurements of the catalyst bed were made with chromel-alumel thermocouple inserted in a small centrally located thermowell. A small glass cyclone separator was connected to the outlet of the reactor

Reactor B consisted of a 30-in. length of 1.81-in. I.D. stainless steel tube surrounded by a sand bath fluidized by air. The reactor temperature was controlled by heating the sand bath with externally mounted electrical resistance heaters. Temperature measurements of the bed were made with a chromel-alumel thermocouple inserted in a small centrally located stainless steel tube of 0.12-in. O.D. A stainless steel porous plate was welded into the lower part of reactor to support the catalyst bed and to distribute the up-flow gas uniformly. Elutriated catalyst particles were collected by a small cyclone connecte to the outlet of the fluidized reactor. The catalyst carry-over was generally negligible under the experimental conditions employed.

Reactor C consisted of 48-in. length of 6.0-in. I.D. stainless steel tube. The lower flange held a stainless steel porous plate and was connected to a conical preheating section. The reactor temperature was automatically controlled by external heaters. The column was wound with four separate sections of heating wires and was insulated with

asbestos insulation. The temperatures within the bed were measured by chromel-alumel thermocouples in stainless steel wells located on the axis and at the inside reactor wall. The temperature distribution of the bed was occasionally checked by moving the thermocouples in the wells. The upper flange supported a porous stainless steel filter which removed suspended catalyst from the upward flowing gas stream. In down-flow experiments, the upper part of reactor was used as a preheater.

Packings

The packings used in the fixed bed and fluidized bed runs (Run No. 4, 5, 7, 9 and 11) consisted of open-ended 14-mesh/in. stainless steel screen cylinders (wire 0.024-in. in diam.) of two sizes (0.2- x 0.2-in. and 0.4- x 0.4-in.). In addition, the 1.0- x 1.0-in. open-ended 14-mesh/in. mild steel screen cylinder and commercially available Pall ring (1.0- x 1.0-in., mild steel) were used in fluidized bed runs as more probable sizes for possible industrial applications (Run No. 12 and 13). In a typical run, the packing was poured into the reactor to a height greater than that of the catalyst bed. A known weight of bed was then added. In general, the ratio of the height of packing to the catalyst bed was maintained at 2.0 to 3.0, so as to make ample allowance for bed expansion in the fluidized bed runs. The catalytic activity of packing in the isomerization reaction was tested. All the packings used in the bed were inactive at temperatures less than 250°C, but some catalytic activity appeared above this temperature.

Gas Analysis

Samples of inlet and outlet gas were analyzed for cyclopropane and propylene by gas chromatography using a Beckman GC-2A Chromatograph equipped with a Hydrogen Flame Detector. Separations were made by a 10-ft. column of propylene carbonate on fire brick using helium as the carrier gas. The temperature of the column was 40°C and the pressure within the column 30 psig. The sample volume was approximately 1.5 ml.

Preliminary experiments

Prior to establishing a standard testing procedure, preliminary experiments on catalyst stability were carried out in a 0.9-in. I.D. x 25-in. glass tube fitted with a porous plate and heated by an electrical resistance wire. Bed temperature was regulated with a temperature controller.

In a typical test, the catalyst activity decreased gradually during the first 70 hrs., then became steady for the period 70 hr. to 400 hr. for the temperature range 200 to 250°C. This reactor also served to evaluate the effects of inlet cyclopropane concentration, the effect of air as compared with nitrogen as the diluent and the catalyst. At the low cyclopropane concentration (about 1 to 2%) used in these experiments, no significant effects on conversion were observed for concentration fluctuations within ± 0.2%. Using 1% cyclopropane concentration the conversions were the same in air as in nitrogen, and so the oxygen in the air had no noticeable effect on the isomerization reaction. The selectivity of the reaction to propylene formation was studied at temperatures from 300 to 450°C. At 300°C the selectivity

was 92%, and it decreased to 18.7% at 450°C. Although propylene was relatively stable over the whole temperature range investigated, a small portion of propylene was consumed at temperatures above 250°C.

From these preliminary experiments the test reaction condition were chosen to be in the temperature range from 150 to 250°C with an inlet gas mixture of 2 ± 0.2% cyclopropane in air.

Procedure

Prior to a set of runs the required amount of catalyst was charged into the reactor and the various heater settings adjusted to give the desired line-out temperature. Before each run, the catalyst bed was dried for about 15-hr. in air at 200°C. Following this pre-treatment, the reaction mixture was fed at the prescribed rate. For fixed bed operation, the gas flowed down through the bed, but up through the bed for fluidized bed runs. The catalyst was allowed to reach a steady activity which took from 50 to 70 hours before a run was begun at the test conditions. In order to compare the conversions at the same experimental conditions between the fixed and fluidized beds, the up-flow gas stream was switched to down-flow for the fixed bed operation at the end of the sampling for the fluidized bed runs.

Temperature distribution within the bed was occasionally checked by moving the thermocouple in the wells. Experimental results were obtained at a constant bed temperature (generally 150 - 250°C ± 1.0°C) to eliminate the need for temperature corrections. The pressur drop across the catalyst bed compared with the operating pressure, which was atmospheric, was assumed negligible. At the maximum flow rates, e.g. in run No. 11 to 13, the outlet reactor pressure was 70-in.

of water, but the greater part of the experimental data were obtained with the reactor pressure less than 20-in. of water. No corrections for pressure drop across the bed or for total reactor pressure were made to the reaction rate calculations. The weights of catalyst collected in the cyclone separator were negligible compared with the bed weights. The catalyst bed height, L, was calculated by dividing the catalyst bed weight by the cross-sectional area of the reactor, and by an average settled catalyst density of 0.87 g/cm^3. With packing present, L was estimated from the total volume of the bed which was the sum of the packing volume and the settled bed volume. Separate reproducibility studies demonstrated that no noticeable change in catalyst activity took place during the runs. The activity was checked during each run by returning to the reference conditions of 200°C and W/F ratios within the range 28.6 to 57.6. The activity was found to be reproducible to within ± 0.5 to ± 2.0%.

Velocity at minimum fluidization

The minimum fluidization velocity was taken as the point at which the pressure drop across the reactor (not the catalyst bed) vs. air flow curve showed a discontinuity. In cases where this sharp discontinuity did not occur, the intersection of the two parts of the curve was selected as the point defining u_f. No large effects of reactor size, bed height, packing size except for Pall ring or temperature on u_f were observed. The average value of u_f over the experimental range used was 0.0108 $\pm$ 0.0007 ft./sec. (u_f for Pall ring was 0.0075 ft./sec.).

RESULTS

A summary of the runs carried out in this investigation is given in Table 1. In run No. 1 the preliminary experiments were carried out using a fluidized bed without packing. In run No. 2 the direct comparisons between the fixed bed without packing and fluidized bed without packing were carried out in the small diameter (0.89-in. I.D.) glass tube. In run No. 3 to No. 11 the effect of reactor size was studied for the same catalyst bed height as in run No. 2 and for the bed held at the same linear scale ratio, L/D, of run No. 2 in the two stainless steel reactors, 1.81-in. and 6.0-in. I.D. Also in these runs the effects of 0.2-in. and 0.4-in. screen packing on conversion were studied. In run No. 12 and No. 13 the effects of 1.0-in. screen packing and 1.0-in. Pall ring were ascertained in the 6.0-in. reactor.

Determination of Reaction Order and Rate Constants

The conversions as a function of W/F and temperature for fixed beds without packing are shown in Figure 3. The conversions of high W/F and high temperature approached 100%, and for this reason it may be concluded that the reaction is irreversible. The pressure drop across the catalyst bed was negligible, because a relatively small bed height and a low flow velocity were used for these fixed bed runs. All the data except a few points fall on single isothermal curves. The data represent three different diameters of reactor, three different L/D ratios, various flow velocities and five reaction temperatures. Although the data on the isothermal curves were a little scattered, no scale effects and no film diffusion effects were observed in these fixed bed runs.

In order to compare the effects of packing on the conversions in a fixed bed without packing, a series of runs with packing present were performed. The experimental conditions were the same as for the fixed bed without packing. The data for these runs are shown in Figure 4. No large effects due to packing were found, although the data for the fixed bed with packing were slightly higher than for the runs without packing. In both fixed bed and fixed bed with packing runs nearly isothermal conditions (± 1.0°C) were obtained within the beds.

The basic rate equation for a catalytic reaction in a flow system is:

$$r_A = aF \frac{dx}{dW} \qquad (1)$$

which can be rearranged as

$$r_A = \frac{adx}{d(\frac{W}{F})}$$

Thus, the rate of conversion corresponding to any given value of W/F in Figure 3 and 4 has been taken as the slope of the tangent to the curve at that point. Conversion rate as a function of conversions for both fixed bed and fixed bed with packing are plotted in Figure 5. A linear relation between the conversion rate and the fractional conversion was found for both series. On the assumption that the total moles of gas undergoes no change during the reaction, the reaction follows approximately a first-order process, for which the equation is:

$$r_A = +k_A a(1-x) \qquad (2)$$

plus sign instead of minus sign

The values of first order over-all rate constant as defined by equation (2) are given in Table 2. The rate constant, k_p, for fixed bed with packing runs is slightly larger than the rate constant, k_o, for runs without packing. The mean value of k_p/k_o ratio within the temperature range 150 to 250°C is 1.09. It is not clear why this ratio is larger. The mean value, k_m, between k_o and k_p was, however, used in this paper as a measure of the activity of the catalyst bed.

Temperature dependency of rate constants was evaluated from an Arrhenius' plot over the temperature range from 150 to 250°C. The data are presented in Figure 6 as a plot of log k_m vs. 1/T. An over-all activation energy of 15.7 Kcal was calculated from the slope of this curve.

Fluidized Bed Runs

Fluidized bed runs without packing were made in the 0.89-, 1.81-, and 6.0-in. reactors at temperatures ranging from 150 to 250°C and with various gas velocities and ratios of L/D. The fractional conversions as a function of W/F obtained in run No. 2, 3, 6, 8 and 10 are given in Table 3. A typical run No. 10-FL-0 is shown in Figure 7. In all cases the conversions from the fluidized bed runs were equal to or less than those from the fixed bed runs. The data indicate that the L/D ratio, gas velocity and reactor diameter influence the final conversion. Furthermore these effects on conversion were much more evident at the higher temperatures.

The effects of packing on the final conversions for fluidized bed were also studied with run No. 4, 5, 7, 9, 11, 12 and 13. The fractional conversions as a function of W/F are given in Table 4. A

typical run No. 11-FL-P is shown in Figure 8. When these curves are compared with the fixed bed curves, it is clear that the conversions approach the fixed bed values, but are lower especially at the higher temperatures. Thus, a simple relationship between conversion and W/F can not be expected in correlating these fluidized bed data. The apparent rate constant was used, however for considering the effects of L/D ratio, gas velocity, reactor diameter and bed temperature on the final conversions.

Apparent Rate Constants for a Fluidized Bed

Equation (2) can be substituted into equation (1) to become

$$aF\frac{dx}{dW} = +k_A a(1-x) \qquad (3)$$

+ sign here

Equation (3) is readily integrated for an idealized piston-flow reactor model to give on rearrangement the equation (4),

$$k_A = \frac{F}{W} \ln \left(\frac{1}{1-x}\right) \qquad (4)$$

For an idealized piston-flow reactor, k_A calculated from equation (4) is numerically equal to catalytic activity and is independent of the conversion, the flow rate and the amount of catalyst. k' estimated from equation (4) for other types of reactor such as fluidized beds is a function of the catalyst activity, the gas-solid contact and the temperature. When the catalyst activity and the temperature are held constant, k' becomes a direct measure of the effectiveness of the gas-solid contact and will generally be smaller than k for an idealized piston-flow reactor, because a fluidized bed reactor, for example, will usually not make as efficient use of the catalyst. Apparent rate

constants as a function of the L/D ratio, gas velocity and temperature for the fluidized bed alone in the 1.8- and 6.0-in. reactor are plotted in Figure 9. When the temperature and the reactor diameter are held constant, k' is a direct measure of the effects of the catalyst bed height and the gas velocity on gas-solid contact in a fluidized reactor. Also, in this figure, a comparison of the value between the fixed and fluidized beds is shown. According to this figure k' appears to decrease rapidly at the high temperature regions with increasing bed height and decreasing gas velocity, but no pronounced effects of these parameters on k' are observed at low temperatures.

Figure 10 shows the effects of the reactor diameter on k' as a function of the gas velocity and the bed temperature at a constant bed height (L = 1.75-in.) for the fluidized bed. At small reactor diameters, the value of k' increased with increasing reactor diameter, but when the diameter is greater than 4-in. this effect appears to become negligible. These results suggest that at a constant bed temperature and gas velocity the effectiveness of a gas-solid contact in a fluidized bed should depend only on the catalyst bed height for reactor diameters greater than 4-in. diameter. Another very interesting result shown in Figure 10 is that k' for the 0.89-in. reactor is independent of the gas velocity, whereas the effect of the gas velocity on k' is large for 1.8-in. and 6.0-in. reactor. Also, this figure illustrates for the fluidized bed the effect of the reactor diameter on k' as a function of the gas velocity and the temperature at constant L/D ratio (1.7 - 1.9), i.e. the so-called geometrical similarity relationships. The plot shows that at small reactor diameters k'

increases with increasing reactor diameter because of the dominating effect of diameter, but that at a reactor diameter above 4-in. k' decreases with increasing reactor diameter because of the diminishing effect of the reactor diameter and the increasing effect of the bed height.

The effects of the L/D ratio on apparent rate constant as a function of the gas velocity and the temperature were also considered for the fluidized bed with packing. The data for the fluidized bed with 0.4-in. screen packing in a 1.8-in. reactor and for the 0.4-in. screen packing in a 6.0-in. reactor are shown in Figure 11. The relationship between k' and L/D is similar in shape to that shown in Figure 10 for the fluidized bed alone, but the values of k' are appreciably higher, and the effect of reactor diameter is very small.

The effect of packing size on k' as a function of the ratio of the reactor diameter, D, to the packing diameter, d, was also considered. The data from the 6-in. reactor at L/D = 1.7 are shown in Figure 12. In this figure, k' from the fluidized bed alone are plotted at D/d = 1. It is clear that the apparent rate constants for the fluidized bed with packing increased approximately inversely with d, when the bed diameter, the bed height, the gas velocity and the bed temperature were held constant. It may also be expected that at larger values of D/d the value of k' for fluidized bed with packing would be close to the fixed bed value, especially at a high gas velocit

The Effect of Temperature on Apparent Rate Constant

The relationship between log k' and 1/T at a constant gas velocity (u = 0.10 ft./sec.) for both the fluidized bed with and withou

packing was examined. The temperature coefficients were estimated using Arrhenius' equation on the data obtained from two different reactors (1.8-in. and 6.0-in.), and these are plotted for two temperature ranges and as a function of the bed height in Figure 13. It is clear from these figures that the temperature coefficient of the apparent rate constant is strongly affected by the temperature region and by the bed height. At the higher temperature range 225 to 250°C, the temperature coefficient for both fluidized bed runs (without and with packing) decrease sharply with increasing bed height to about 5 Kcal. which may suggest that a diffusion-like controlling step has become dominant in the over-all reaction rate.

DISCUSSION OF RESULTS

Product Distribution

In Figure 14 the mole fraction of propylene produced, y, is plotted as a function of the mole fraction of cyclopropane converted, x. These data were obtained from runs with the reactor C, but approximately similar plots were also obtained for reactors A and B. The selectivity, y/x, is less than unity. In this paper, in general, the fraction of cyclopropane converted has been used as the basis for calculating catalyst activity. It has been assumed that the reaction involved the same mechanism over the whole experimental range.

Cyclopropane isomerization is suggested by Roberts (9) to occur by way of a carbonium ion intermediate,

$$\underset{H_2C-CH_2}{\overset{H_2}{C}} + H^+A^- \xrightarrow{k_1} \left[\underset{H_2C}{\overset{H_2}{C}} \diagdown CH_3 \right] A^- \xrightarrow{k_2} H^+A^- + CH_2{=}CH{-}CH_3$$

where H^+A^- represents the solid acid catalyst. If such a consecutive

reaction really occurs, the data in Figure 14 may be correlated by an equation, in which the time variable has been eliminated

$$\frac{dC_R}{dC_A} = -1 + \frac{k_2 C_R}{k_1 C_A} \qquad (5)$$

The solid and dashed line in Figure 14 were calculated for the mixed and piston flow systems at $k_2/k_1 = 8.0$, respectively. At high values of x, the concentration of propylene produced during reaction is slightly smaller than for the lines calculated by assuming the simple consecutive reaction. It may be concluded that the real reaction mechanism is more complicated than the simple consecutive reaction assumed, and some production of polymer, decomposition of propylene and so on occurs.

Fluidized Bed Without Packing

All the conversion data for the fluidized bed alone are plotted as a function of dimensionless group, kW/F, gas velocity and bed height in Figure 15. The dashed line represents the conversion curve if the reaction took place in a completely mixed bed without by passing. Except for the results at L/D = 0.3 in 6.0-in. reactor, which lie above the dashed curve, all the data fall below or on the completely mixed curve. The data in Figure 15-1 represent the conversions at four different velocities for a constant bed height, L/D = 1.9, in the 0.89-in. reactor. The data do not fall on the single curve and are a function of gas velocity. The data in Figure 15-2 represent the conversions for five gas velocities and two L/D ratios in the 1.8-in. reactor. All the data appear to be independent of the gas velocity and the bed height, and fall on a single curve which

approaches closely to the completely mixed system. Actually, k' does depend upon gas velocity and bed height as shown in Figure 9 but the insensitiveness of the curves in Figure 15 to velocity and bed height is surprising as has been noted by Orcutt et al. (12). The data in Figure 15-3 represent the conversions for five gas velocities at L/D = 0.3 and for eight gas velocities at L/D = 1.7. All the data obtained at L/D = 0.3 which was the same bed height as shown in Figure 15-2 fall on a single curve which is slightly higher than the line calculated for the completely mixed system. When such a shallow bed, i.e. L = 1.75-in., is used, the effects of the reactor diameter on conversion would be expected to be small and gas-catalyst contact would be expected to follow approximately the completely mixed system. When a bed of 10.2-in. height was used the conversions as a function of kW/F were clearly dependent upon the gas velocity. The data for the velocity range 0.03 to 1.0 ft./sec. at L/D = 1.7 show clearly that the higher the gas velocity the lower the conversion.

Fluidized Bed With Packing.

All the data as a function of dimensionless group, kW/F, gas velocity and bed height for fluidized bed with three different size screen packing, 0.2-, 0.4- and 1.0-in., and with 1.0-in. Pall-ring in the two different reactors, 1.8- and 6.0-in. diameter, are given in Figure 16. The two dashed lines were calculated with the assumption of piston flow or of a completely mixed system. All the data in those figures lie between the piston flow and the completely mixed lines, whereas almost all the data for fluidized beds alone were lower than the completely mixed line. This illustrates a major effect of the presence

of packing on the fluidized bed conversions.

In Figure 16-1 the results at 1.9-in. bed height for a fluidized bed with 0.2-in. screen packing in the 1.8-in. reactor are shown. The data represent five temperatures (150 to 200°C) and five gas velocities (0.03 to 0.16 ft./sec). All the data fall on the single curve which is close to the piston flow line. Thus it may be concluded that reaction temperature and gas velocity contribute no secondary effects. In Figure 16-2 the results for the fluidized bed with 0.4-in. screen packing in 1.8-in. reactor are shown. The data represent two bed heights (1.85- and 3.66-in.), five gas velocities (0.03 to 0.16 ft./sec. and 0.03 to 0.26 ft./sec., respectively) and five temperatures (150 to 250°C). These data also fall on a single curve which lies between the piston flow and the completely mixed system within a deviation of about ± 2.0%. The curve through the data is lower than that for the 0.2-in. screen packing used in Figure 16-1, and the effects of packing size are clearly indicated. However, again no significant difference due to the reaction temperature, the gas velocity and the bed height were observed. In Figure 16-3 the results obtained with 0.4-in. screen packing in 6.0-in. reactor are given. The data represent two bed heights (1.85-in. and 11.0-in.), five gas velocities (0.03 to 0.20 ft./sec. and 0.10 to 0.53 ft./sec., respectively) and five temperatures (150 to 250°C). Most of the data fall on a single curve within a deviation of about ± 2.0%. Although the same 0.4-in. screen packing as in Figure 16-2 was used in this case, the curve is very similar to the curve shown in Figure 16-1 for which a 0.2-in. packing was used. In Figure 16-4

the results for a 1.0-in. packing in 6.0-in. reactor are shown. The data represent two different packings (1.0-in. screen packing and 1.0-in. Pall ring), five gas velocities (0.1 to 0.52 ft./sec.) and five temperatures. All the data fall on a single curve for each packing with a deviation of about ± 2.0%. The curve for a 1.0-in. Pall ring is higher than for a 1.0-in. screen packing, and so it should be noted that packing characteristics other than size should be taken into account.

In general, for fluidized beds with packing the most important factors affecting the conversion as a function of kW/F seems to be the D/d ratio and packing characteristics such as size, geometric form etc. rather than the effects of bed height, reactor diameter, linear velocity, etc.

Application of Various Reactor Models

It is a well-known fact that the piston-flow model and the completely mixed model do not correlate well with fluidized bed data. In order to account for the deviations from these idealized models, other models have been proposed. These are summarized by Levenspiel (13).

Dispersion Model

The dispersion model assumes that the flow of gas through a bed involves some longitudinal mixing which is characterized by an effective diffusion or dispersion coefficient. The steady state differential equation in dimensionless form is

$$\frac{1}{Pe}\frac{d^2f(z)}{dz^2} - \frac{df(z)}{dz} - Rf(z) = 0 \qquad (6)$$

where the dimensionless Peclet group contains the dispersion coefficient. Assuming that there is no volume change during the reaction, equation (6)

has been solved analytically by Wehner and Wilhelm (14) for a first-order reaction. Their solution modified in this paper by a different ratio of the characteristic length, $z = w/W$, reduces to equation (7) for a closed vessel,

$$1-x_A = \frac{4a\exp\left(\frac{1}{2}\frac{FW}{D_e'}\right)}{(1+a)^2\exp\left(\frac{a}{2}\frac{FW}{D_e'}\right) - (1-a)^2\exp\left(-\frac{a}{2}\frac{FW}{D_e'}\right)} \qquad (7)$$

where

$$a = \sqrt{1 + 4k\frac{W}{F}\left(\frac{D_e'}{FW}\right)}$$

D' is a modified axial dispersion coefficient, [g. g-mole/hr.] It is assumed in this solution that no stagnant gas pockets or gross by-passing occurs. This model usually represents quite satisfactorily flow which deviates only moderately from piston flow. In the limit of an infinitely large diffusivity the solution is identical with that of the completely mixed model. With zero diffusivity the solution approaches that for a piston flow.

Since the conversion data reported in this study for non-packed fluidized beds are generally below the conversion line based on a completely mixed model (see Figure 15), it is concluded that the dispersion model can not be applied to this case. However, when packing is used the final conversions obtained fall in the region between the piston-flow and the completely mixed models, and so a dispersion model may be applicable. In Figure 17 the dispersion model has been used to correlate the data obtained from 6-in. reactor. The data represent two different bed heights (1.85- and 10.0- to 11.0-in.), the gas velocity ranges (0.03- to 0.2- and 0.1- to 0.52-ft./sec.) and

two packing sizes (0.4- and 1.0-in.). The two solid lines were calculated from equation (7) using assumed values for the dimensionless group D_e'/FW. The calculated curve is higher at small kW/F and lower at large kW/F, but the deviations were within ± 2% over the range of kW/F from 1.0 to 7.0. In this case, i.e. for a fluidized bed containing 0.4-in. packing, the curve represents a reasonably good fit.

Tanks-in-series model

For this model it is assumed that the real reactor can be represented by a series of n equally sized fully mixed compartments. The early studies were done by McMullin and Weber (15). The out-put conversions for this model can be shown to be

$$x = 1 - \frac{1}{(1 + kW/F)^n} \qquad (8)$$

where n is the number of compartments. Various researchers have attempted to increase the reactor efficiency of fluidized beds by using horizontal baffles to form compartments or by using fluidized beds in series. Overcashier et al. (5) reported that the spread in residence times may be effectively narrowed by the use of baffles, although not to the extent corresponding to an equivalent number of completely mixed stages.

Although gas-solid contacting behaviour between the fluidized beds with packing and beds divided by horizontal baffles may be essentially different, a comparison between the results on a fluidized bed with packing and with the tanks-in-series model is made in Figure 18 in order to estimate the equivalent number of complete mixing stages.

In Figure 16-2 and 16-3 the data for the fluidized bed with 0.4-in. screen packing in the 1.8-in. and in the 6.0-in. reactor are shown, respectively. These data also represent two different bed heights (1.85- and 3.66-in.) and two gas velocity ranges (0.03- to 0.16- and 0.03- to 0.26-ft./sec.) for the 1.8-in. reactor; and two different bed heights (1.85- and 11.0-in.) and two gas velocity ranges (0.03- to 0.2- and 0.1- to 0.52 ft./sec.) for the 6.0-in. reactor. If the tanks-in-series model is to be used for a fluidized bed with packing, the reactor with a high bed can be represented by a series of n equally sized completely mixed compartment. Assuming that the non-packed fluidized bed at 1.85-in. height, which was found to be a nearly completely mixed system, is a unit compartment, two solid lines were calculated from equation (8) for two compartments (n = 2 for L = 3.7-in. and six compartments (n = 6 for L = 11.1-in.), respectively. The data for 6.0-in. reactor fit sufficiently well for the tanks-in-series model but the data for 1.8-in. reactor are slightly higher than the calculated line in spite of the small D/d ratio. These results suggest that the fluidized reactor efficiency may be increased by the use of packing to or above the extent corresponding to an equivalent number of completely mixed unit stages.

Two phase model

Neither a piston flow nor a completely mixed model corresponds to the real behavior existing in most fluidized beds. Even the dispersion model or the tank-in-series model which allow for substantial deviation from the simpler extremes of piston flow and a fully mixed system are not satisfactory either. Various two phase models, therefore have been proposed to fit the conditions prevailing in a fluidized

bed (13). These models are based on the observation that in a fluidized bed, bubbles form at the bottom and grow or coalesce as they rise through the bed. The dynamics and size of these bubbles have a large effect on conversion in a fluidized bed.

Bubble-flow models have resulted from studies of bubble flow (12, 16, 17, 18). Orcutt et al. (12) developed a bubble-flow model which assumed that all but the gas required for minimum fluidization passes through as bubbles. For the completely mixed case, it is also assumed that within the dense phase there is complete mixing and a first-order reaction, and that within each rising bubble there is complete mixing and no reaction, and that there is an exchange of gas between the two phases. The fraction of the reactant which is unconverted as a function of kW/F is given by

$$\Phi(k) = 1 - x = \beta e^{-n_K} + \frac{1 - \beta e^{-n_K}}{\frac{K}{n_K} + 1} \qquad (9)$$

where

$$\beta = 1 - (u_f/u)$$

$$n_K = qh/u_b V$$

$$K = kW/F$$

For the piston-flow case, it is assumed that there is complete mixing within each rising bubble, but piston flow in the dense phase. The expression giving the fraction of unconverted reactant is,

$$\Phi(K) = 1 - x = \frac{1}{m_2 - m_1} m_2 e^{-m_1\left(1-\frac{m_1 u_f}{n_K u}\right)} - m_1 e^{-m_2\left(1-\frac{m_2 u_f}{n_K u}\right)} \qquad (10)$$

where m_1 and m_2 are the roots of

$$m^2 - m(n_K + K)/(1 - \beta) + Kn_K/(1 - \beta) = 0$$

the limiting cases of Equations (9) and (10) are,

a) $\lim_{K = 0} \Phi(K) = 1$

b) $\lim_{K = \infty} \Phi(K) = \beta e^{-n_K} = f_b$

c) $\lim \frac{d}{dK} \Phi(K) = -1$

As n_K approaches infinity, the system tends to complete mixing with no by-passing for the complete mixing case, and to piston flow with no by-passing for the piston flow case, respectively.

In Figure 19 correlations for all the data from the fluidized bed with packing runs and part of the data for fluidized bed without packing based on the bubble-flow model by Orcutt et al. are given. As shown previously, all of the data with packing are in a region between the idealized piston-flow line and the complete mixing line, and nearly all of the data without packing fall on the complete mixing line or below the line. The solid lines through the data for the packing runs were calculated using Equation (10) (piston-flow case), whereas Equation (9) (complete mixing case) was used for the non-packed fluidized bed runs. These solid lines are calculated at u_f/u equal to 0.1 for various values of f_b. Satisfactory agreement between the experimental data and the calculated lines was obtained. For the fluidized beds with packing the experimental data are more dependent on the packing than on reactor diameter, bed height and gas velocity, but for the fluidized bed alone the data are very

sensitive to these latter parameters. Thus, the gas velocity range was limited to the lower values from 0.1- to 0.13 ft./sec. for the fluidized bed alone. In general, for fluidized beds alone, the fractional degree of by-passing seems to depend on reactor size, bed height and flow conditions. Bubbles appear not to be uniformly distributed over the height of the bed. A good correlation over a wide experimental range is therefore unlikely. When packing is used in the fluidized bed, such scale effects as reactor size and bed height etc. are much smaller or perhaps negligible. Consequently, the comparatively simple models described earlier should be useful in designing fluidized beds containing packing.

Assuming that bubbles in the bed are uniformly distributed over the height of the bed, the number of mass transfer units, n_K, as a function of the equivalent bubble diameter, d_b, was estimated at various fluidized bed heights at minimum fluidization, h_f, using the method proposed by Davidson (19). The molecular diffusion coefficient of propylene, 1.04 ft.^{2}/hr., instead of cyclopropane through the air was used and was calculated by the method of Fair and Lerner (20). Then, n_K as a function of the fraction gas by-passing the bed, f_b, at $\beta = 0.9$ was calculated from the following equation,

$$f_b = \beta e^{-n_K} = (1 - \frac{u_f}{u}) e^{-n_K} \qquad (11)$$

Combining these two relationships, d_b as a function of f_b at a constant value of u_f/u may be estimated.

The mean equivalent bubble diameter, $\bar{d}_b$, as a function of the catalyst bed height at a linear velocity of 0.1 ft./sec. for both the fluidized bed, with and without packing, in the 1.8- and

6.0-in. reactor are shown in Figure 20. The curve for the fluidized bed increases with increasing catalyst bed, and shows that the bubble sizes are increasing as they rise through the bed. Since the curve with packing is flat, it is concluded that the bubbles are approximately similar in size and are uniformly distributed over the height of the bed owing to the packing which impedes the growth of large bubbles.

$\bar{d}_b$ as a function of the linear velocity at a bed height of 9.9- to 11.0-in. in 6.0-in. reactor is also shown in Figure 21. $\bar{d}_b$ for the fluidized bed increased with increasing linear velocity, whereas with the packing present the effect of the linear velocity on $\bar{d}_b$ is small enough to ignore. $\bar{d}_b$ as a function of the packing size at L/D = 1.0~2.0 and u = 0.1 ft./sec. for a fluidized bed with packing in 1.8- and 6.0-in. reactor are shown in Figure 22. This curve shows that $\bar{d}_b$ increases proportionately with the packing diameter, which suggests that the packing diameter $\bar{d}_b$ when the same kind of packing is used is a very important factor.

GENERAL CONCLUSIONS

Direct experimental comparisons of the kinetic data between the fixed bed operating (with and without packing) and the fluidized bed operating (with and without packing) in reactors of various diameters have shown that the use of a packing in the fluidized bed greatly improved the final conversions obtained. When a well designed packing is used in the fluidized reactor, the final conversions should be close to the fixed bed values while still

retaining most of the attractive operating characteristics of the fluidized bed.

For the fluidized bed alone, difficulties were encountered in correlating the data over the whole range of operating conditions using simplified models such as the dispersion model, the tanks-in-series model or the two phase model, because of the complex effect of gas velocity and catalyst bed height on conversion.

For a fluidized bed with packing, simplified models such as a dispersion model, a tanks-in-series model or a two phase model can be applied to correlate the data. The assumption that the fluidized bed efficiency was independent of bed height and bed diameter, that the bubbles in the bed were uniformly distributed over the height of the bed, and that the effects of gas velocity on the conversion as a function of kW/F were negligible etc. appear to be approximately justified.

NOMENCLATURE

a = mole fraction of cyclopropane in feed

C_A = concentration of cyclopropane, mole/ft.^{3}

C_R = concentration of reaction intermediate, mole/ft.^{3}

D = diameter of reactor, in.

D_e' = modified axial dispersion coefficient, g. g-mole/hr.

d = diameter of packing, in.

$\bar{d}_b$ = mean equivalent bubble diameter, in.

F = feed rate, mole/hr.

f = fraction of reactant remaining

f_b = fraction gas by-passing the fluidized bed with packing

f_b' = fraction gas by-passing the fluidized bed without packing

h = fluidized bed height, in.

k = first order rate constant, g-mole/g-hr.

k' = apparent rate constant for fluidized bed, g-mole/g-hr.

k_m = mean value between k_o and k_p, g-mole/g-hr.

k_o = first order rate constant for no-packing present, g-mole/g-hr.

k_p = first order rate constant for with packing present, g-mole/g-hr.

L = catalyst bed height, in.

n = number of compartments

n_K = number of mass transfer units

Pe = Peclet group, Lu/D

Δp = pressure drop through the bed, in. H_2O

q = volumetric flow through a bubble, ft.^{3}/sec.

R = dimensionless rate group, kL/u

r = reaction rate, g-mole/g-hr.

T = reaction temperature, °C or °K

t = time

u = superficial linear velocity, ft./sec.

u_b = bubble rising velocity, ft./sec.

u_f = minimum fluidization velocity, ft./sec.

V = bubble volume, $ft.^3$

W = weight of catalyst bed, g

x = fraction of cyclopropane reacted

y = fraction of propylene produced

z = dimensionless distance in the axial direction w/W

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Table 1. Summary of runs.

RUN	REACTOR		PACKING	CATALYST BED		
	No.	INSIDE DIAMETER D (IN.)	DIAMETER d (IN.)	WEIGHT $\overline{W}$ (G.)	HEIGHT L (IN.)	L/D (-)
1-FL-0	P	0.87	NO	14.5	1.76	2.0
2-F1X-0	A	0.89	"	"	1.72	1.9
2-FL-0	"	"	"	"	"	"
3-F1X-0	B	1.81	"	62.8	1.72	1.0
3-FL-0	"	"	"	"	"	"
4-F1X-P	"	"	0.2	"	1.89	"
4-FL-P	"	"	"	"	"	"
5-F1X-P	"	"	0.4	"	1.85	"
5-FL-P	"	"	"	"	"	"
6-F1X-0	"	"	NO	125.5	3.42	1.9
6-FL-0	"	"	"	"	"	"
7-F1X-P	"	"	0.4	"	3.66	2.0
7-FL-P	"	"	"	"	"	"
8-F1X-0	C	6.0	NO	697.0	1.72	0.3
8-FL-0	"	"	"	"	"	"
9-F1X-P	"	"	0.4	"	1.85	"
9-FL-P	"	"	"	"	"	"
10-FL-0	"	"	NO	4185.0	10.3	1.7
11-FL-P	"	"	0.4	"	11.0	1.8
12-FL-P	"	"	1.0	4019.0	10.1	1.7
13-FL-P	"	"	1.0 Pall ring	3970.0	10.2	"

Table 2. The over-all first order rate constants for a fixed bed and a fixed bed with packing.

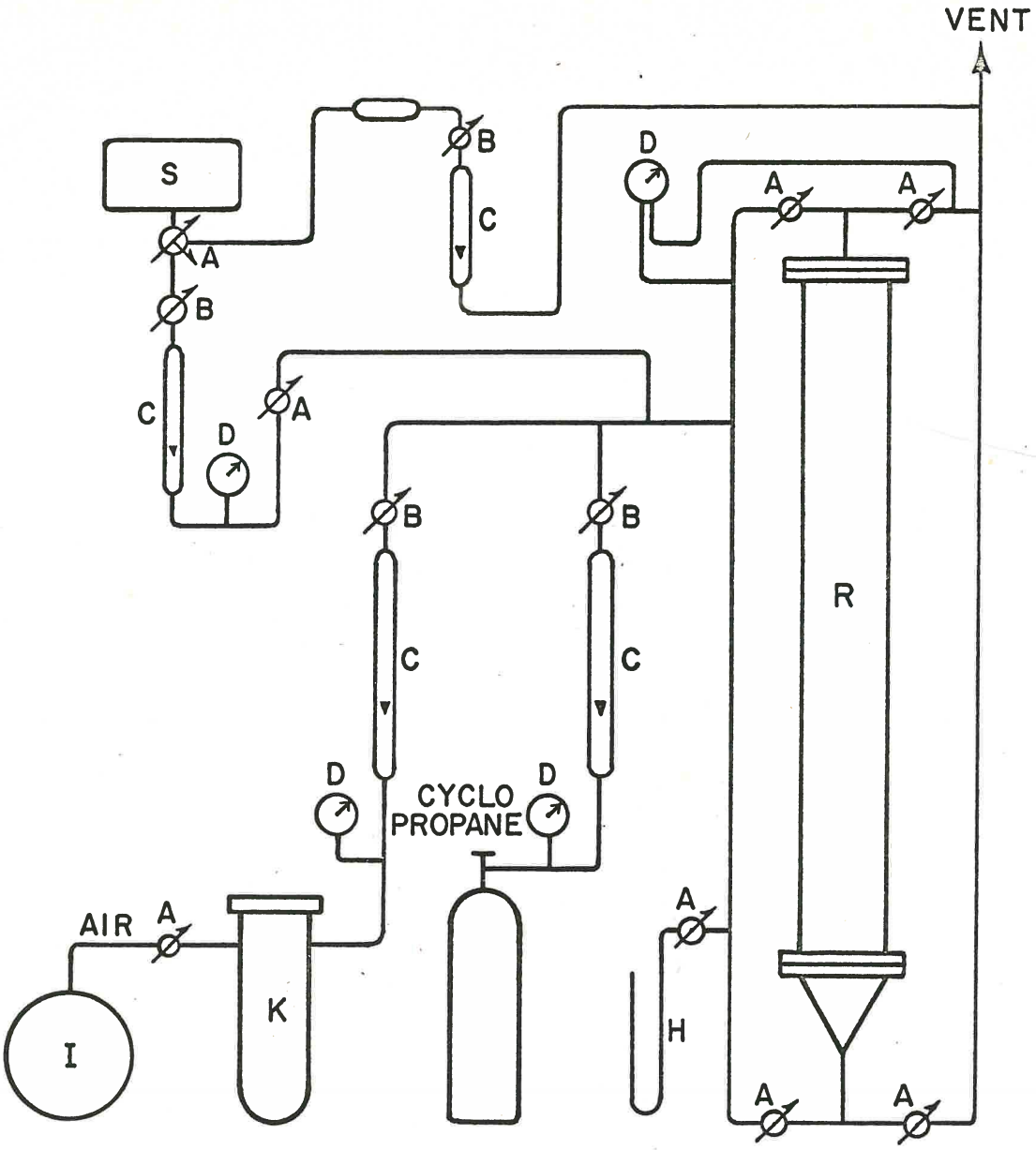
TEMPERATURE °C	RATE CONSTANTS		MEAN RATE CONSTANTS	$\frac{k_p}{k_o}$
	NO PACKING $k_o \times 10^2$	WITH PACKING $k_p \times 10^2$	$k_m \times 10^2$	
150	0.36	0.41	0.39	1.14
175	1.05	1.12	1.09	1.07
200	2.88	3.17	3.03	1.10
225	6.65	7.10	6.88	1.07
250	13.30	14.30	13.80	1.08

Table 3. Experimental results for a fluidized bed without packing.

Run No.	2-FL-0		3-FL-0		6-FL-0		8-FL-0		10-FL-0	
Reactor	A		B		B		C		C	
D in.	0.89		1.81		1.81		6.0		6.0	
L in.	1.72		1.72		3.42		1.72		10.3	
W g	14.5		62.5		125.5		697.0		4185.0	
Temp. °C	W/F	x	W/F	x	W/F	x	W/F	x	W/F	x
150	17.5	5.2	17.3	3.9	17.2	7.3	27.7	12.4	14.2	6.7
	28.8	9.5	28.8	8.1	34.5	9.5	54.5	18.4	27.5	8.5
	65.7	16.2	60.4	17.8	57.5	18.6	116.2	36.5	56.5	15.6
			135.0	37.4	142.2	32.8			113.0	28.8
									257.0	50.4
175	17.5	11.5	17.3	14.0	17.2	16.2	13.9	12.9	14.2	15.8
	28.8	21.6	28.7	19.4	34.5	27.4	27.7	22.1	27.5	23.5
	65.8	46.1	60.4	46.2	57.5	42.6	54.5	39.3	56.5	33.7
	117.0	60.1	135.0	77.1	142.2	68.2	116.2	68.8	113.0	50.4
									257.0	66.9
200	17.5	20.4	17.3	31.4	17.2	33.9	13.9	27.1	14.4	28.3
	28.8	41.3	28.8	39.8	34.5	49.4	27.6	44.1	27.5	40.4
	67.8	72.0	61.6	68.0	57.5	64.4	54.4	63.1	56.5	57.0
	117.0	86.2	135.0	95.9	142.2	91.4	115.8	94.4	113.2	72.8
									257.0	82.5
225	17.5	41.3	17.3	50.0	17.2	53.0	13.9	50.2	14.4	41.3
	28.5	55.8	28.7	62.6	34.5	70.4	27.7	66.7	27.5	53.9
	67.7	92.2	59.8	88.5	57.5	78.4	54.5	82.0	56.5	69.5
					142.2	93.6			113.0	82.8
									257.0	89.0
250	17.5	49.2	17.3	66.0	17.2	65.0	7.2	52.1	14.2	52.0
	28.5	67.0	28.7	76.7	34.5	81.2	13.9	68.7	27.5	63.4
			59.3	93.3	57.5	87.8	27.7	83.0	56.5	78.2
					142.2	96.6	54.5	93.5	113.0	88.6
									257.0	92.4

Table 4. Experimental results for a fluidized bed with packing.

Run No.	4-FL-P		5-FL-P		7-FL-P		9-FL-P		11-FL-P		12-FL-P		13-FL-P	
Reactor	B		B		B		C		C		C		C	
D	1.81		1.81		1.81		6.0		6.0		6.0		6.0	
L	1.72		1.85		3.66		1.85		11.0		10.1		10.2	
W	62.5		62.8		125.5		697.0		4185.0		4019.0		3970.0	
Packing,d	0.2-in.Screen		0.4-in.Screen		0.4-in.Screen		0.4-in.Screen		0.4-in.Screen		1.0-in.Screen		1.0-in. Pall ring	
Temp., °C	W/F	x	W/F	x	W/F	x	W/F	x	W/F	x	W/F	x	W/F	x
150	8.6	5.5	8.6	3.4	17.2	5.7	27.7	12.5	14.2	6.7	13.7	7.6	25.2	10.6
	17.3	8.6	28.8	9.9	34.5	10.6	54.4	23.2	27.5	12.1	26.4	12.7	53.6	20.0
	28.8	9.4	71.5	21.7	57.6	19.5	116.0	39.0	56.5	21.3	54.3	24.0	107.2	34.1
	70.9	19.4			143.0	38.6			113.0	34.3	109.0	42.0		
175	8.6	9.6	8.6	11.4	17.2	17.4	14.0	16.6	14.2	19.4	13.7	19.6	13.5	17.3
	17.3	17.1	17.3	15.9	34.5	32.5	27.7	30.9	27.5	28.7	26.4	28.0	25.2	28.0
	28.8	30.6	28.7	23.5	57.6	47.2	54.2	51.4	56.5	47.3	54.3	46.0	53.6	45.4
	71.2	53.8	71.5	50.9	143.0	70.2	115.8	76.8	113.0	70.5	109.0	62.4	107.2	63.8
200	8.6	26.7	8.6	23.3	17.2	39.3	13.9	33.7	14.2	38.4	13.7	34.3	13.5	36.9
	17.3	40.0	17.3	34.1	34.5	59.5	27.7	56.7	27.5	56.4	26.4	50.0	25.2	51.7
	28.7	57.8	28.8	49.1	57.6	79.1	54.3	80.3	56.5	79.4	54.3	73.1	53.6	74.2
	71.2	84.8	71.5	86.7	143.0	93.4	115.8	96.8	113.0	95.0	109.0	86.7	107.2	89.7
225	8.6	49.2	4.9	29.8	17.2	59.7	7.1	38.1	14.2	64.4	13.7	57.3	13.5	61.8
	17.3	68.6	8.6	44.6	34.5	82.6	13.9	57.0	27.5	80.9	26.4	70.3	25.2	75.1
	28.8	82.2	17.3	58.0	57.3	93.0	27.7	81.8	56.5	94.8	54.3	85.8	53.6	88.8
	71.4	98.6	28.7	78.6	142.5	97.8	54.3	96.2	113.0	98.7	109.0	95.7	107.2	95.3
			71.7	95.7										
250	8.6	68.7	4.9	41.6	17.2	76.4	7.1	58.7	14.2	84.7	13.7	73.8	13.5	77.9
	17.3	87.3	8.6	58.0	34.5	91.8	13.9	77.7	27.5	93.7	26.4	83.2	25.2	87.5
	28.8	94.4	17.3	78.0	57.6	97.9	27.7	93.2	56.5	98.7	54.3	92.2	53.6	95.3
	71.4	99.5	28.8	91.6	143.0	100.0			113.0	99.7	109.0	96.3	107.2	97.5
			71.5	99.6										



A - REGULATING VALVE
B - FLOW CONTROL VALVE
C - ROTAMETER
D - PRESSURE GAUGE
H - WATER MANOMETER

I - AIR COMPRESSOR
K - SEPARATOR
R - REACTOR UNIT
S - ANALYSIS UNIT

Fig.1 Arrangement of experimental equipment

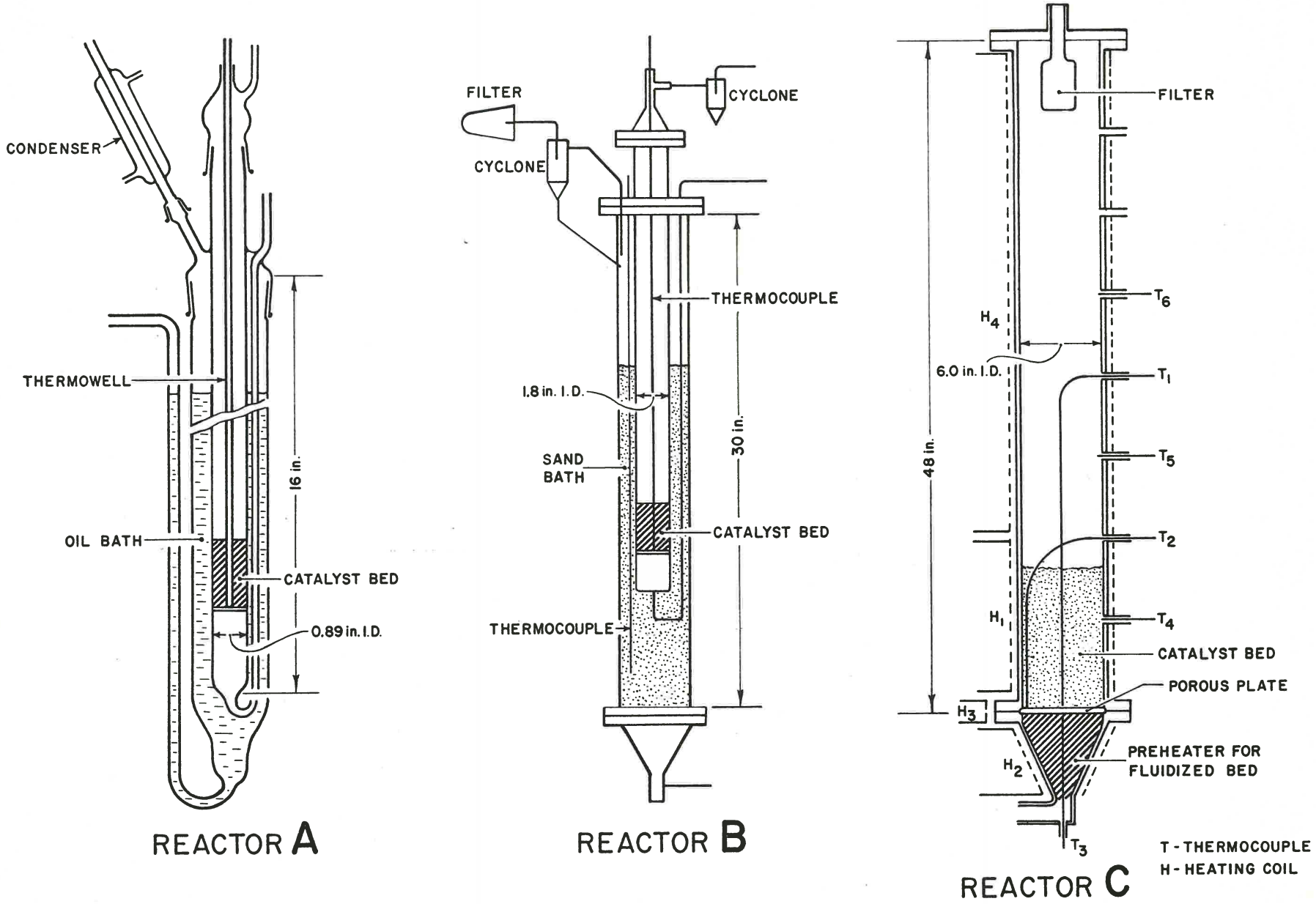


Fig. 2 DETAILS OF REACTORS

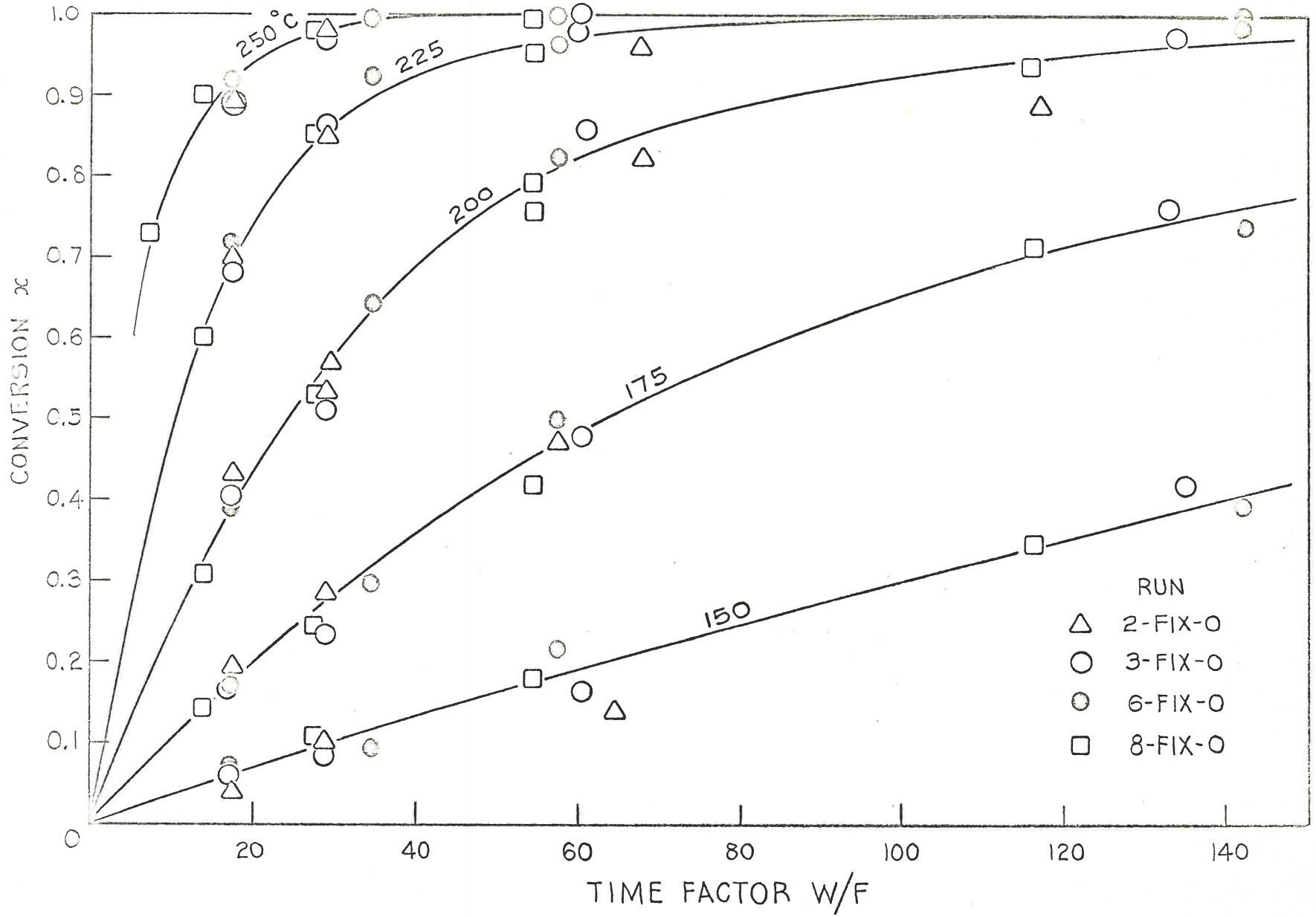


Fig. 3 CONVERSION AS A FUNCTION OF W/F FOR A FIXED BED

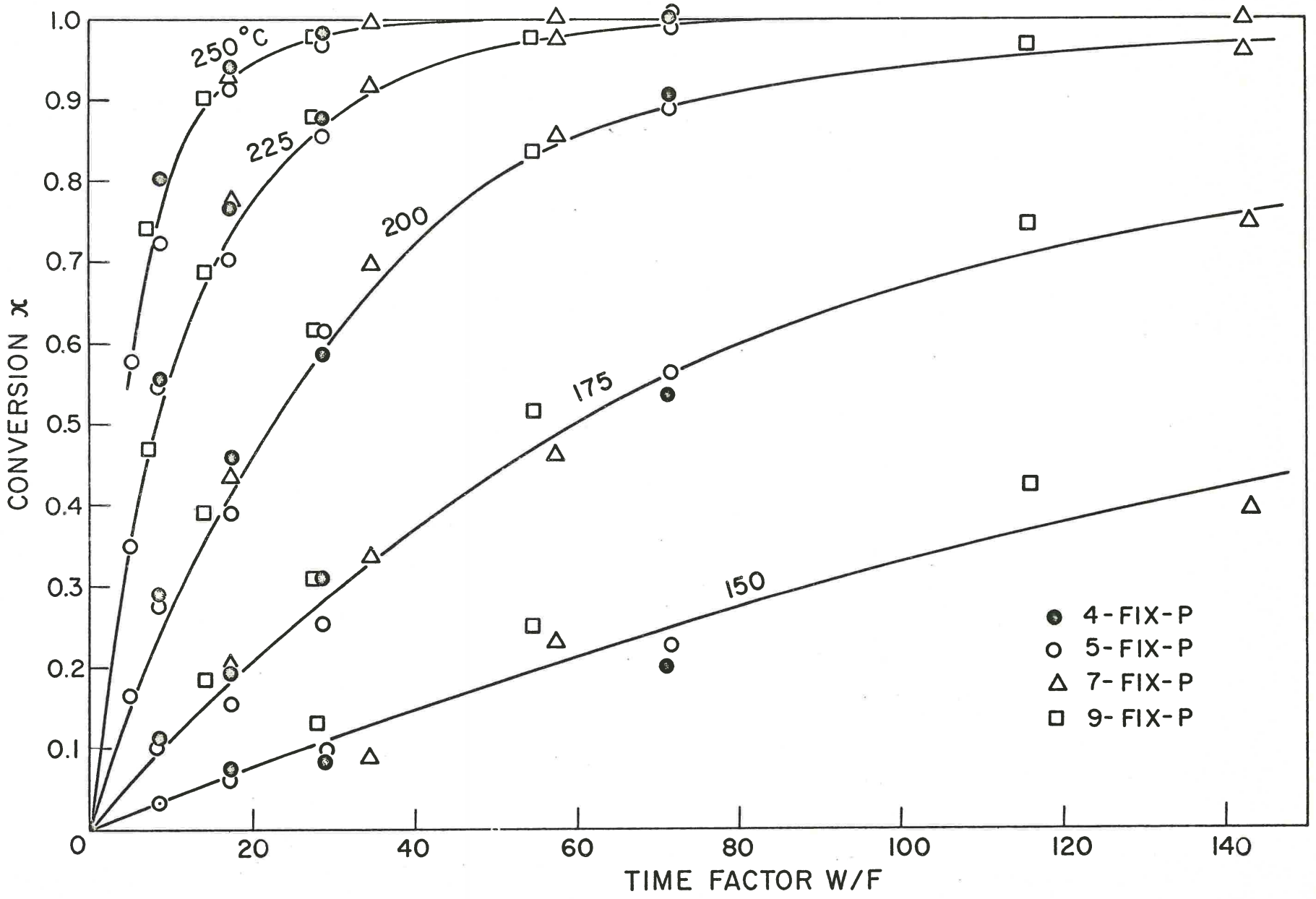


Fig. 4 Conversion as a function of W/F for a fixed bed with packing

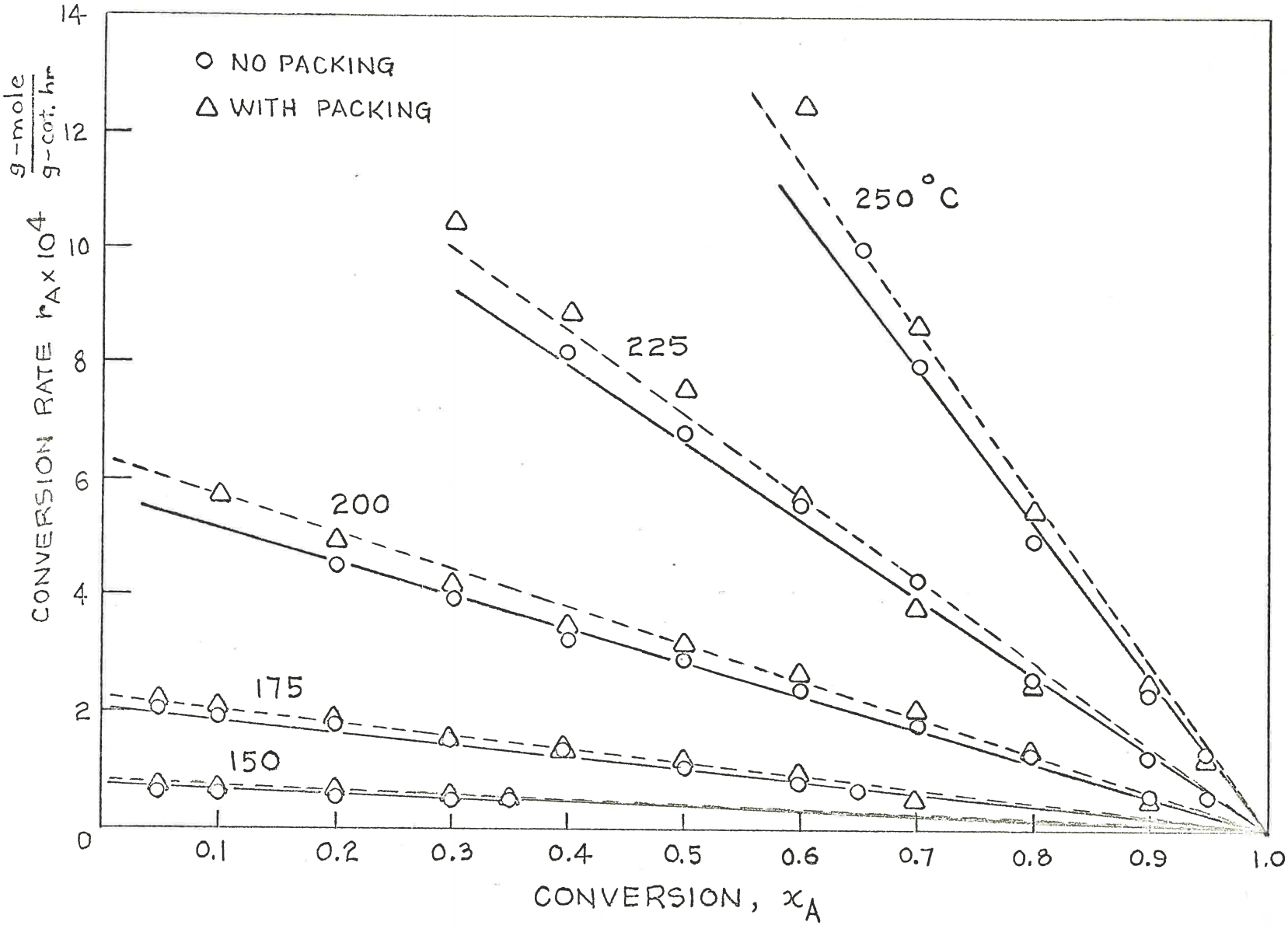


Fig. 5 CONVERSION RATE AS A FUNCTION OF CONVERSION FOR A FIXED BED WITH AND WITHOUT PACKING.

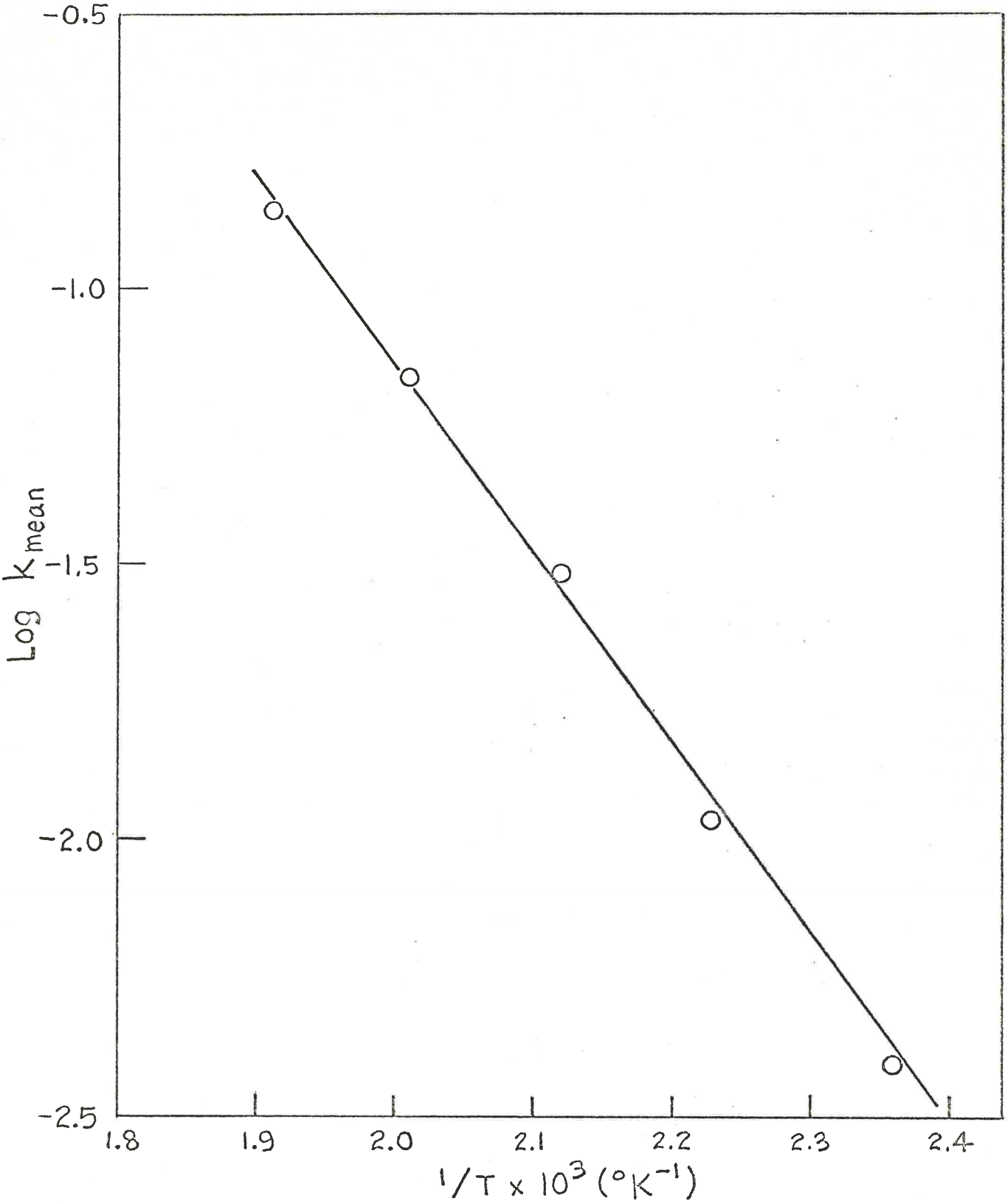


Fig. 6 ARRHENIUS PLOT FOR THE ISOMERIZATION OF CYCLOPROPANE.

THE OVER-ALL ACTIVATION ENERGY = 15.7 K cal.

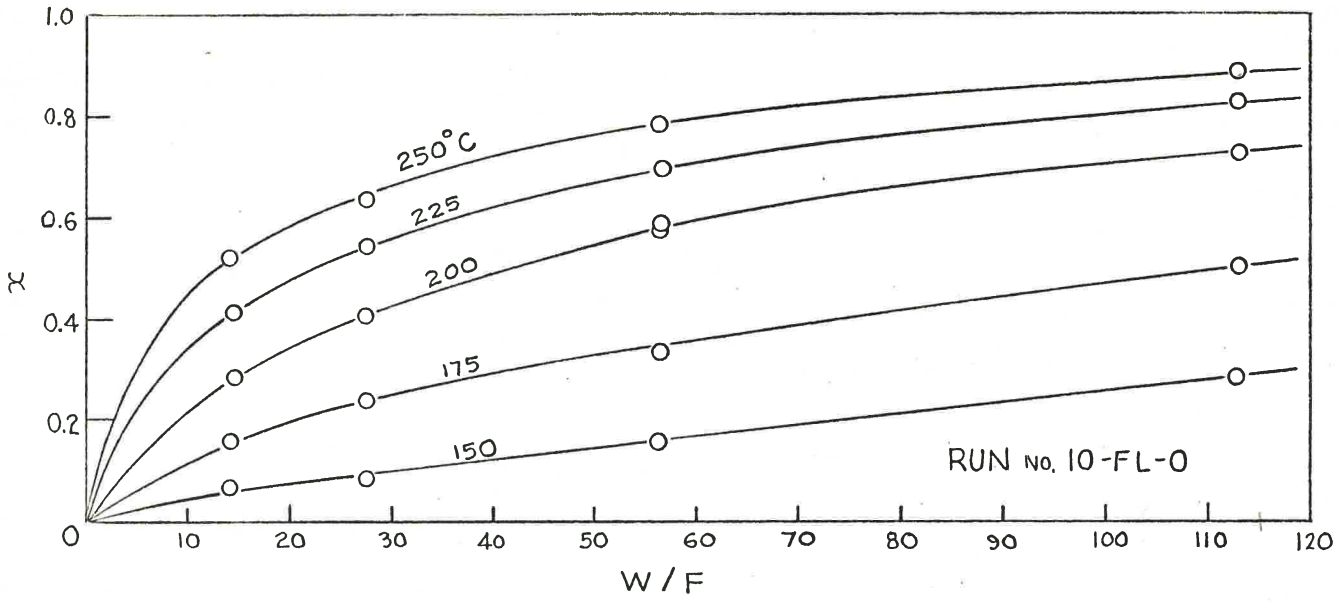


Fig. 7 CONVERSION AS A FUNCTION OF W/F AND TEMPERATURE FOR A FLUIDIZED BED WITHOUT PACKING.

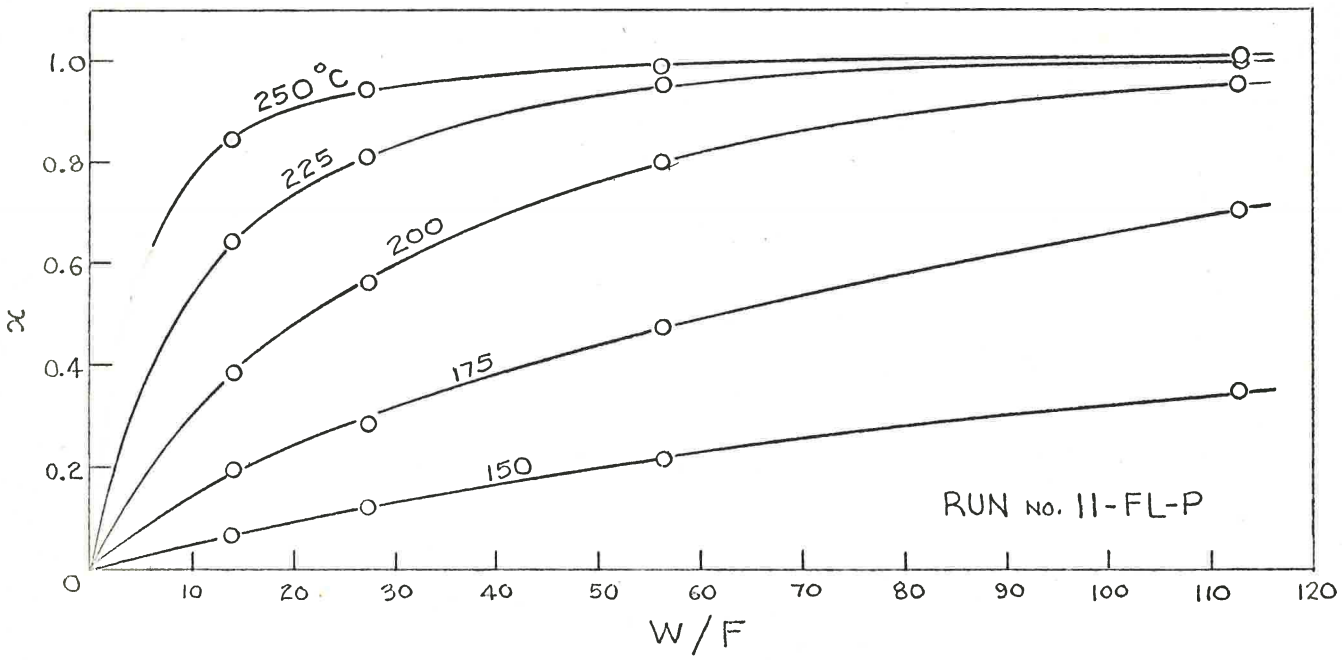


Fig. 8 CONVERSION AS A FUNCTION OF W/F AND TEMPERATURE FOR A FLUIDIZED BED WITH PACKING

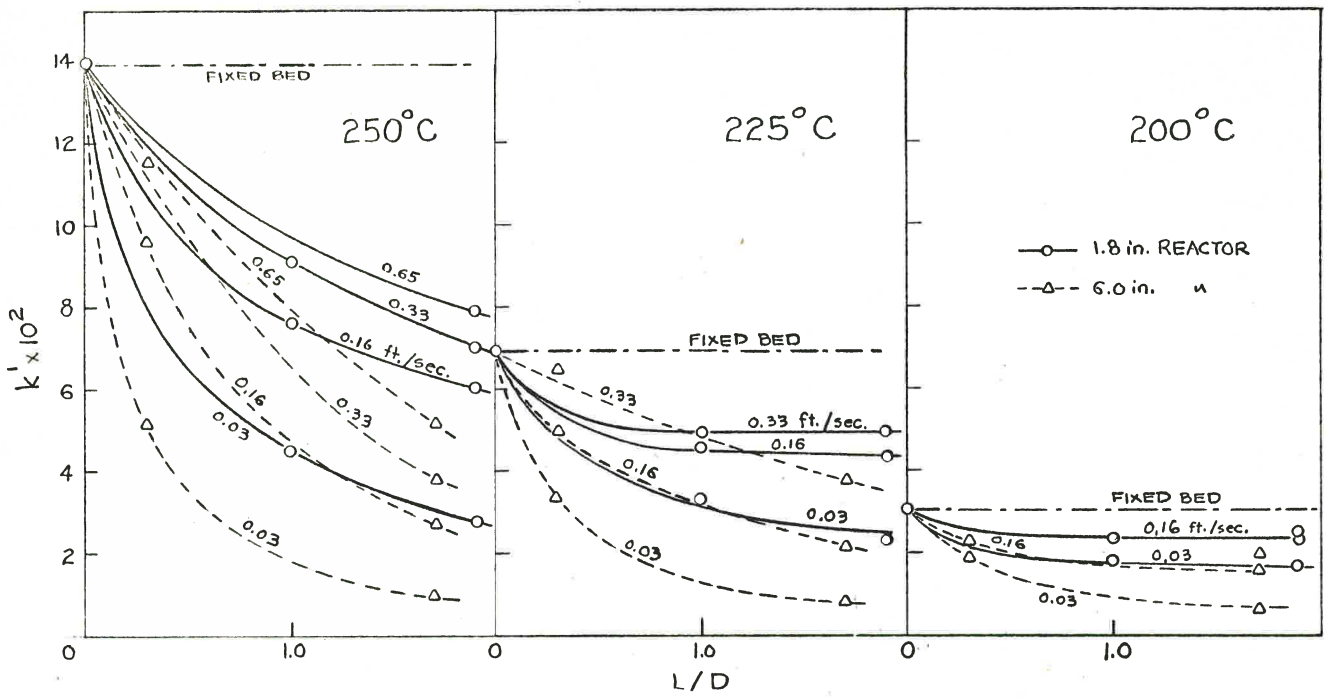


Fig. 9 APPARENT RATE CONSTANT AS A FUNCTION OF L/D RATIO, GAS VELOCITY AND TEMPERATURE FOR A FLUIDIZED BED WITHOUT PACKING IN 1.8 AND 6.0 in. REACTOR

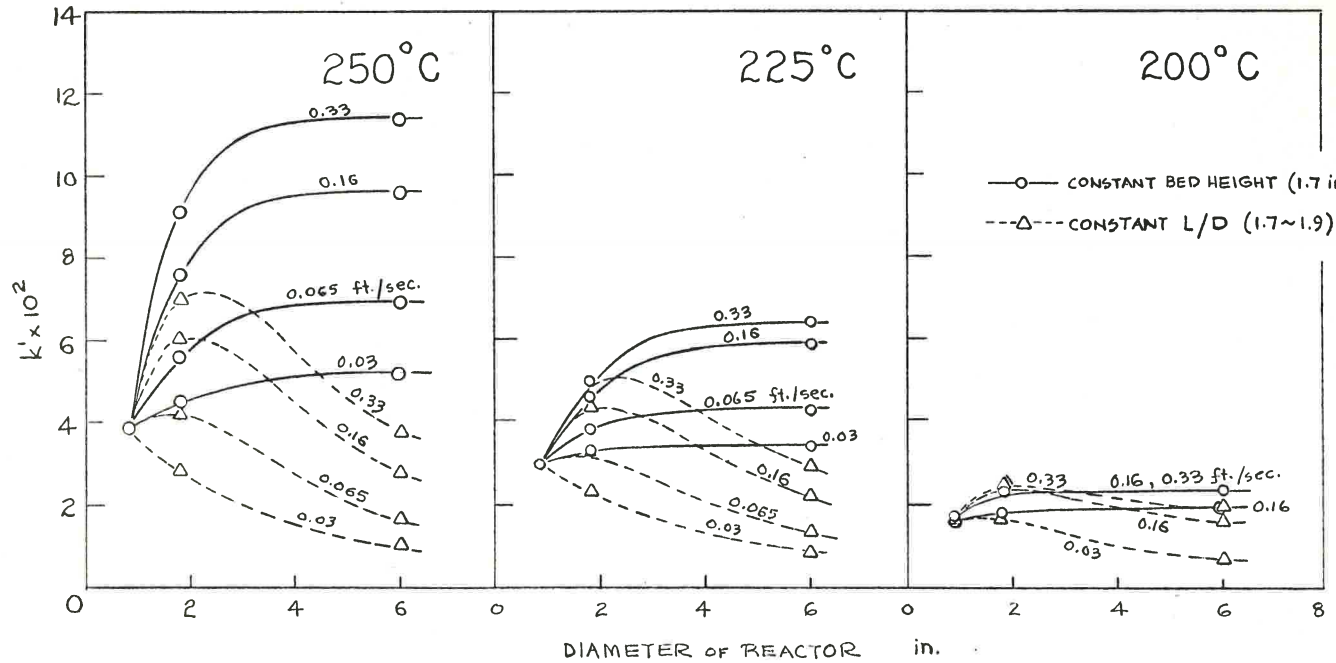


Fig. 10 APPARENT RATE CONSTANT AS A FUNCTION OF REACTOR DIAMETER, SUPERFICIAL GAS VELOCITY AND TEMPERATURE AT CONSTANT BED HEIGHT (1.7 in.) AND L/D (1.7~1.9) FOR A FLUIDIZED BED WITHOUT PACKING

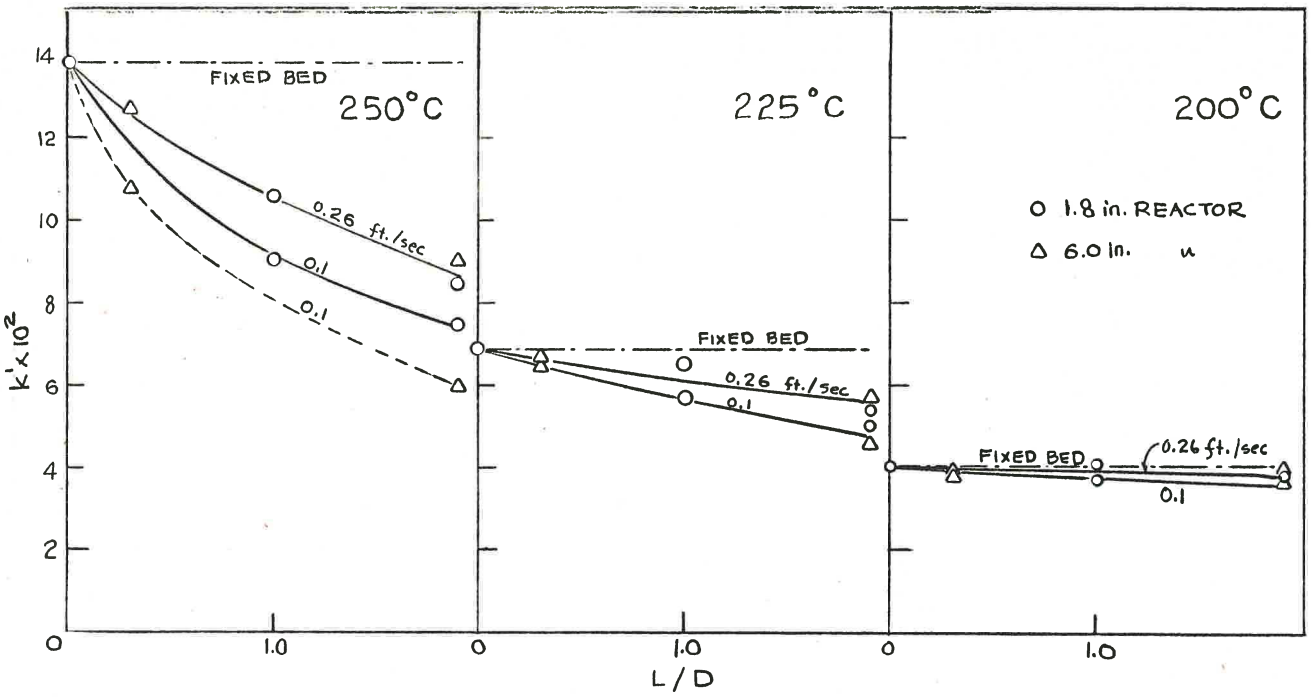


Fig. 11 APPARENT RATE CONSTANT AS A FUNCTION OF L/D RATIO, SUPERFICIAL GAS VELOCITY AND TEMPERATURE FOR A FLUIDIZED BED WITH 0.4 in. SCREEN PACKING IN 1.8 AND 6.0 in. REACTORS.

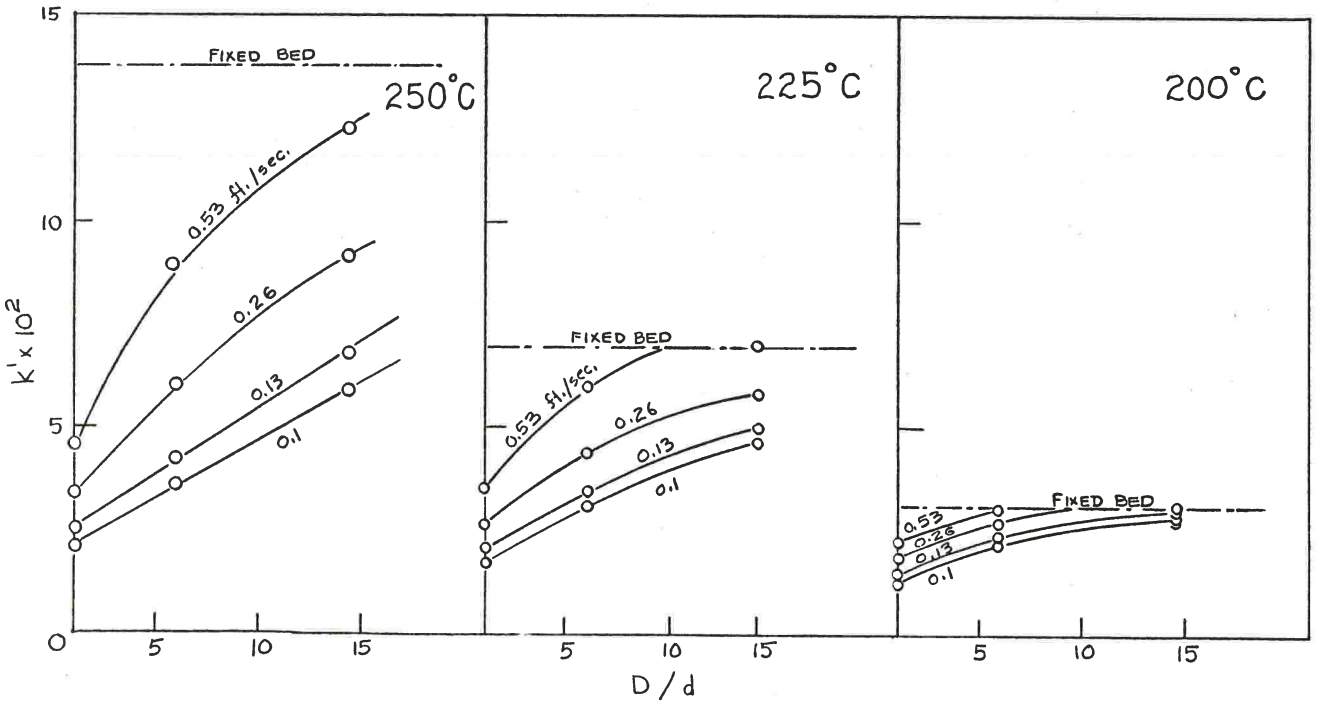


Fig. 12 THE EFFECT OF PACKING SIZE ON k' AS A FUNCTION OF D/d RATIO, GAS VELOCITY AND TEMPERATURE AT L/D = 1.7 FOR A FLUIDIZED BED WITH PACKING IN 6.0 in. REACTOR

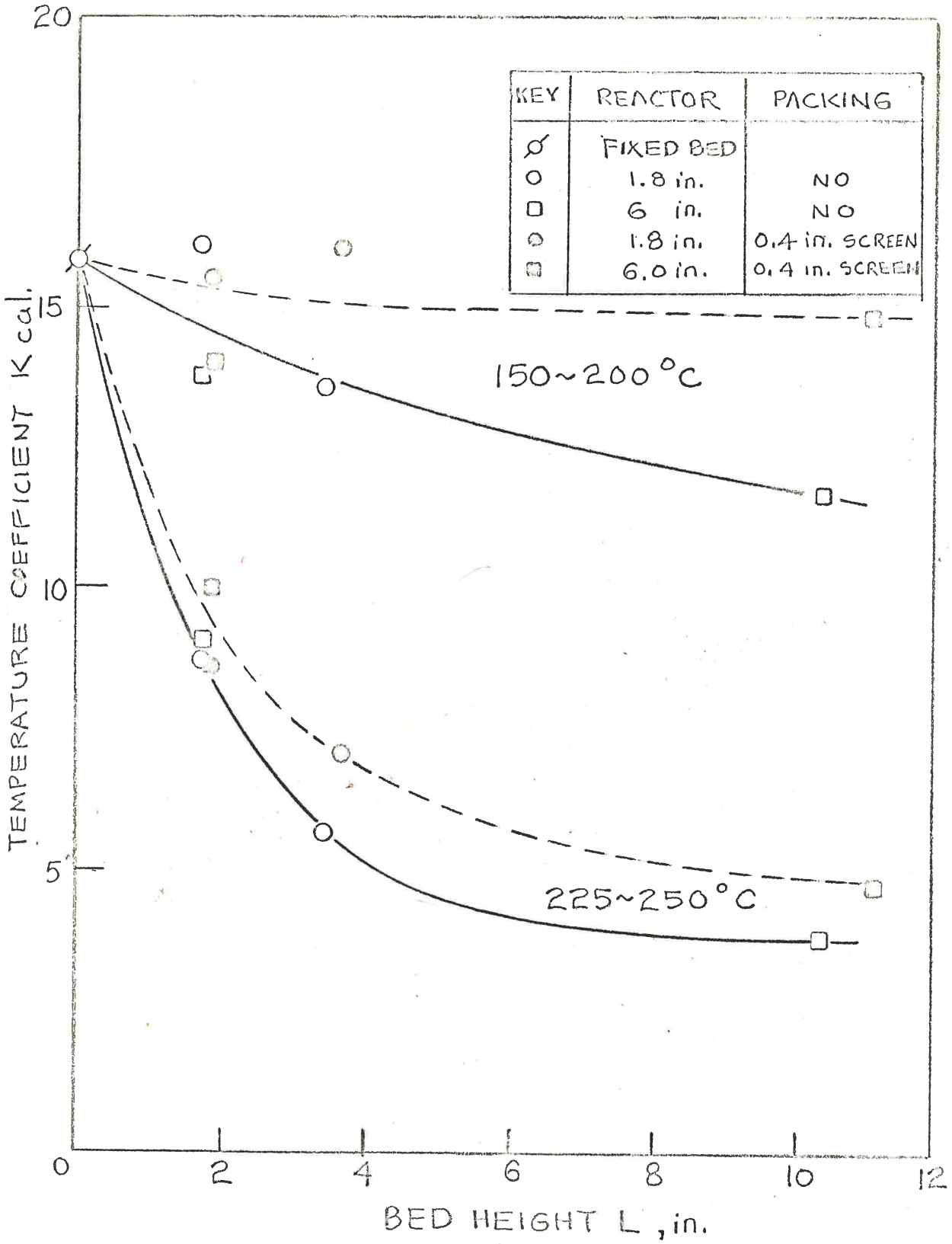


Fig. 13 TEMPERATURE COEFFICIENT OF APPARENT RATE CONSTANT AT CONSTANT GAS VELOCITY 0.1 ft./sec. FOR A FLUIDIZED BED WITH AND WITHOUT PACKING IN 1.8 in. AND 6.0 in. REACTOR

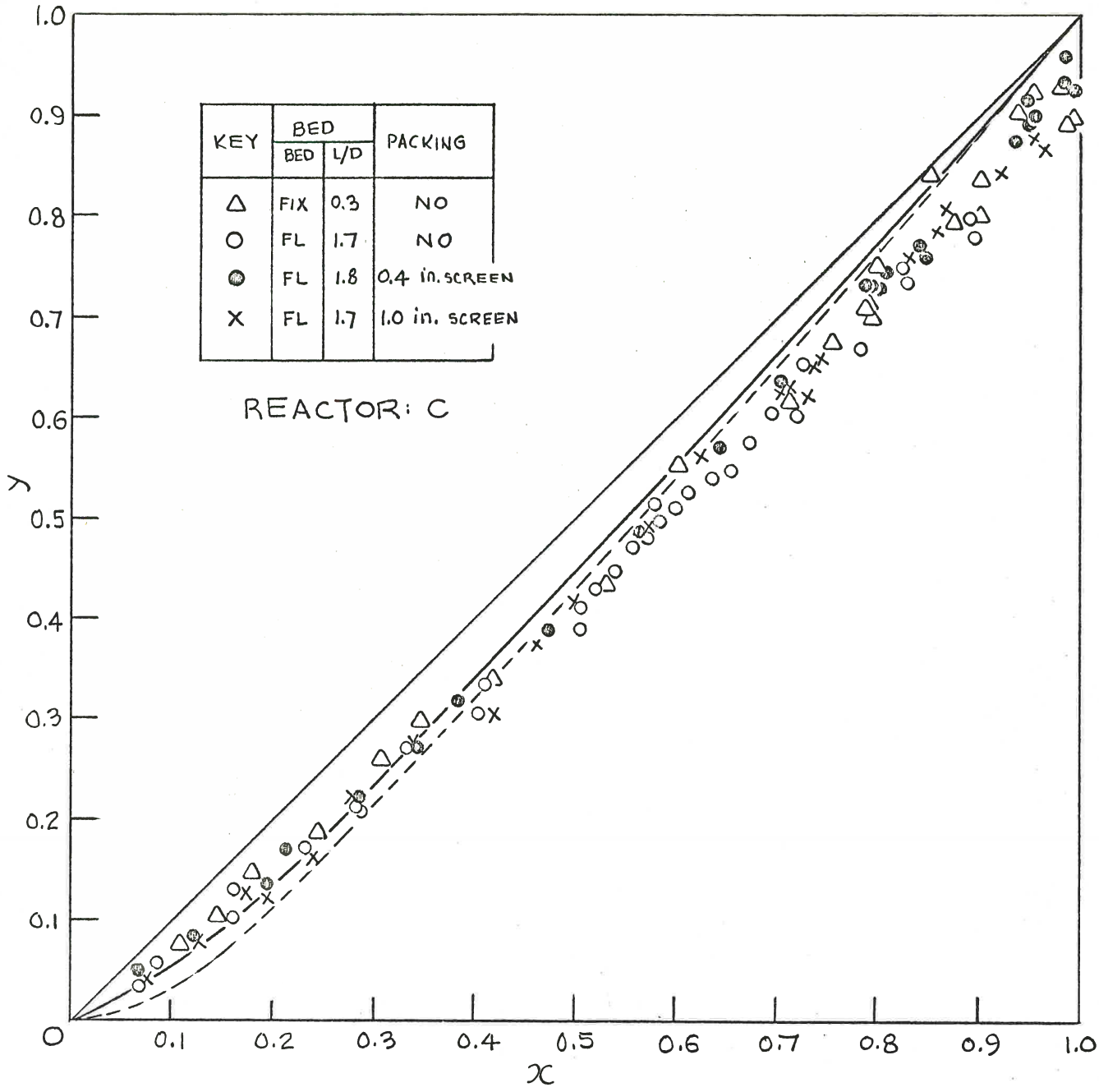


Fig. 14 PROPYLENE PRODUCED VS. CYCLOPROPANE CONSUMED FOR REACTOR C

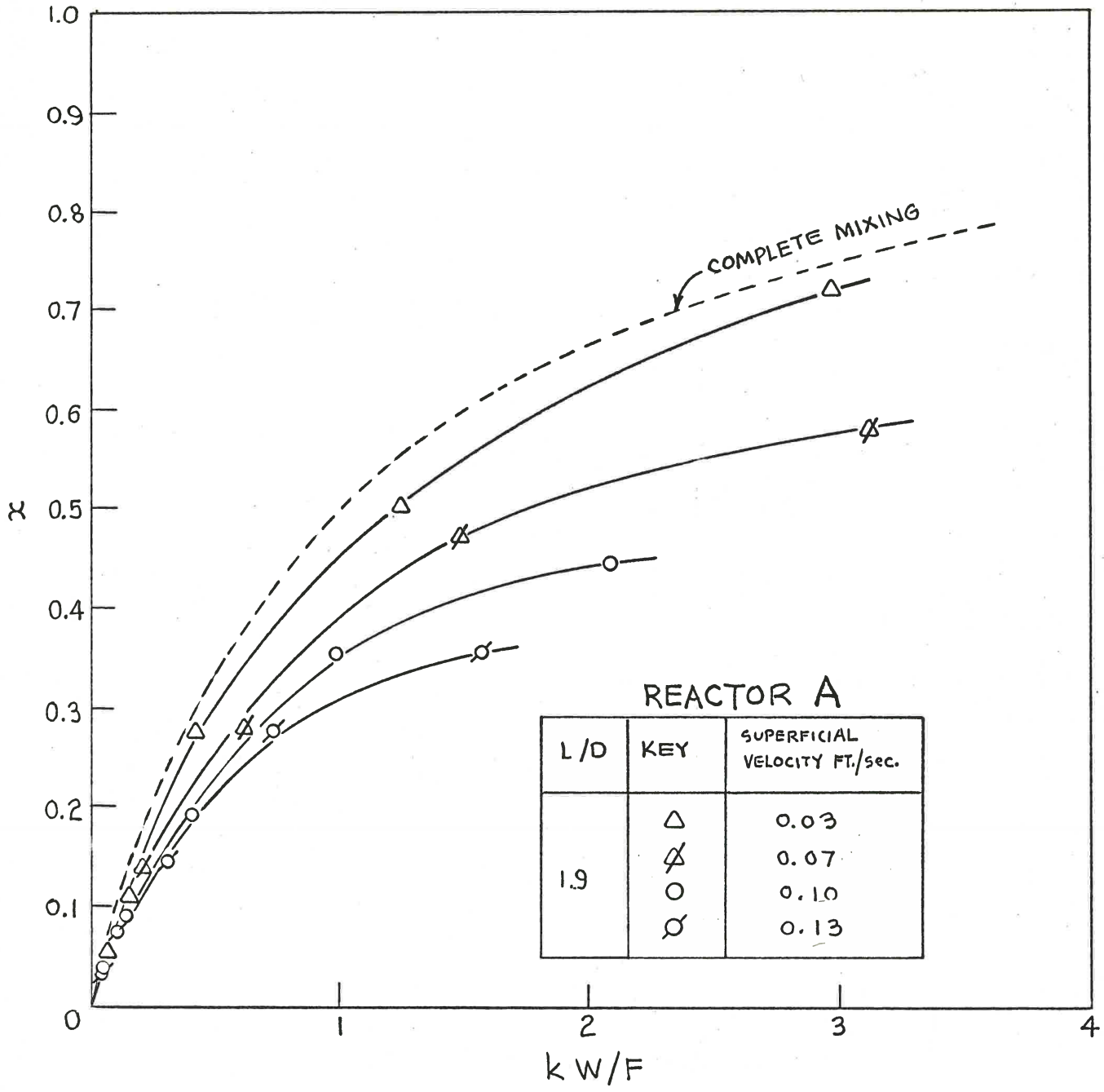


Fig. 15-1 CONVERSION AS A FUNCTION OF kW/F FOR A FLUIDIZED BED WITHOUT PACKING

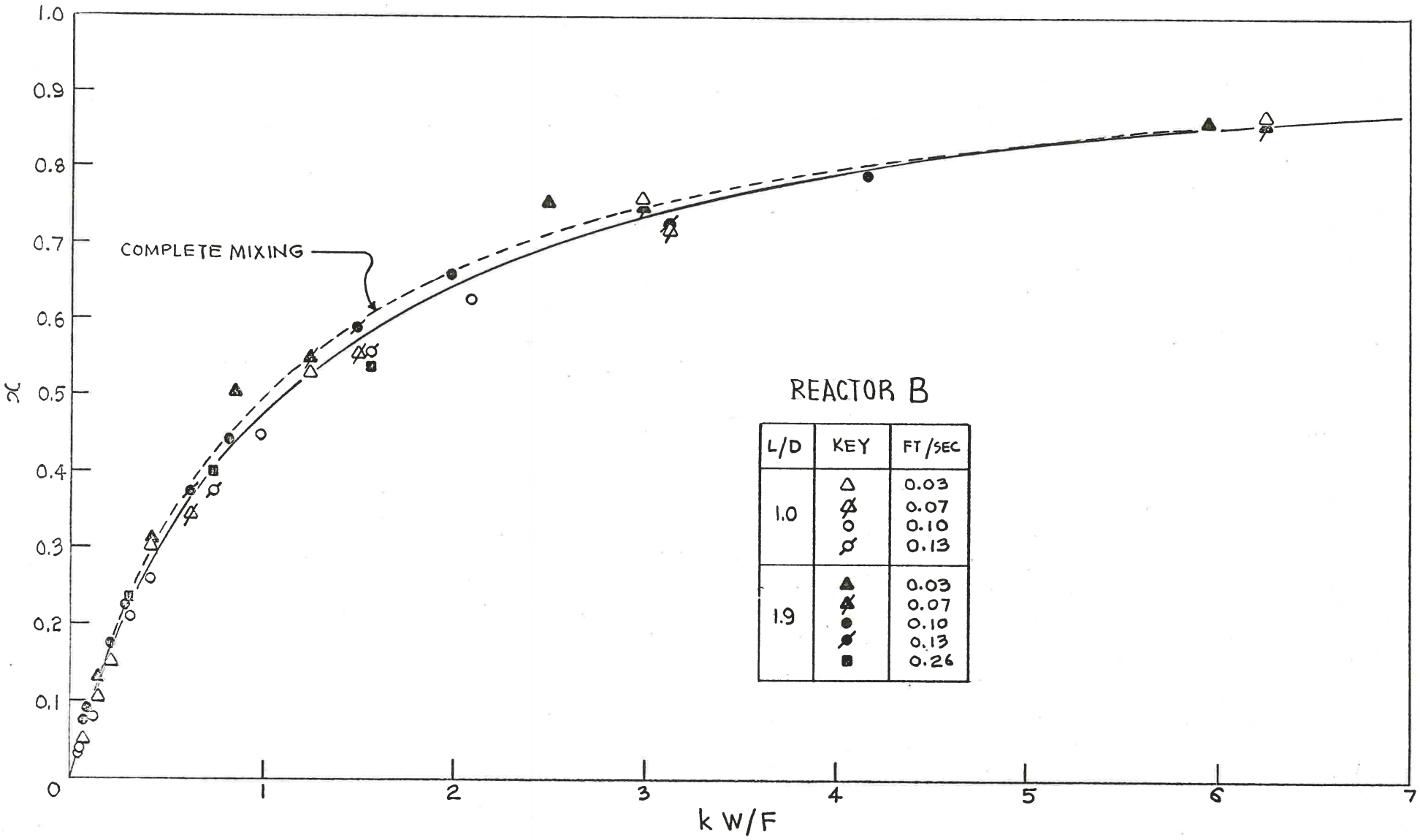


Fig. 15-2 CONVERSION AS A FUNCTION OF k W/F FOR A FLUIDIZED BED WITHOUT PACKING

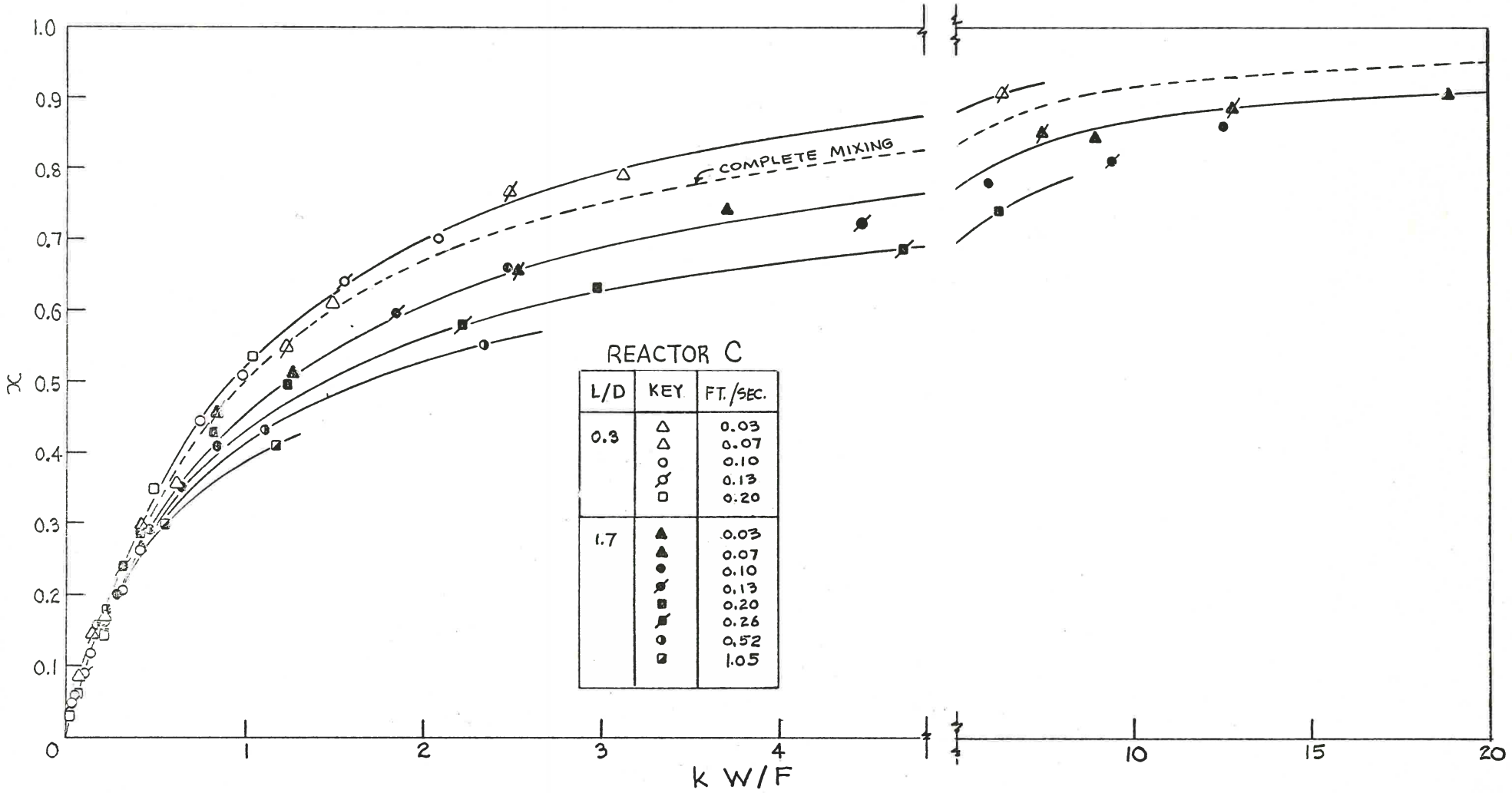


Fig. 15-3 CONVERSION AS A FUNCTION OF kW/F FOR A FLUIDIZED BED WITHOUT PACKING

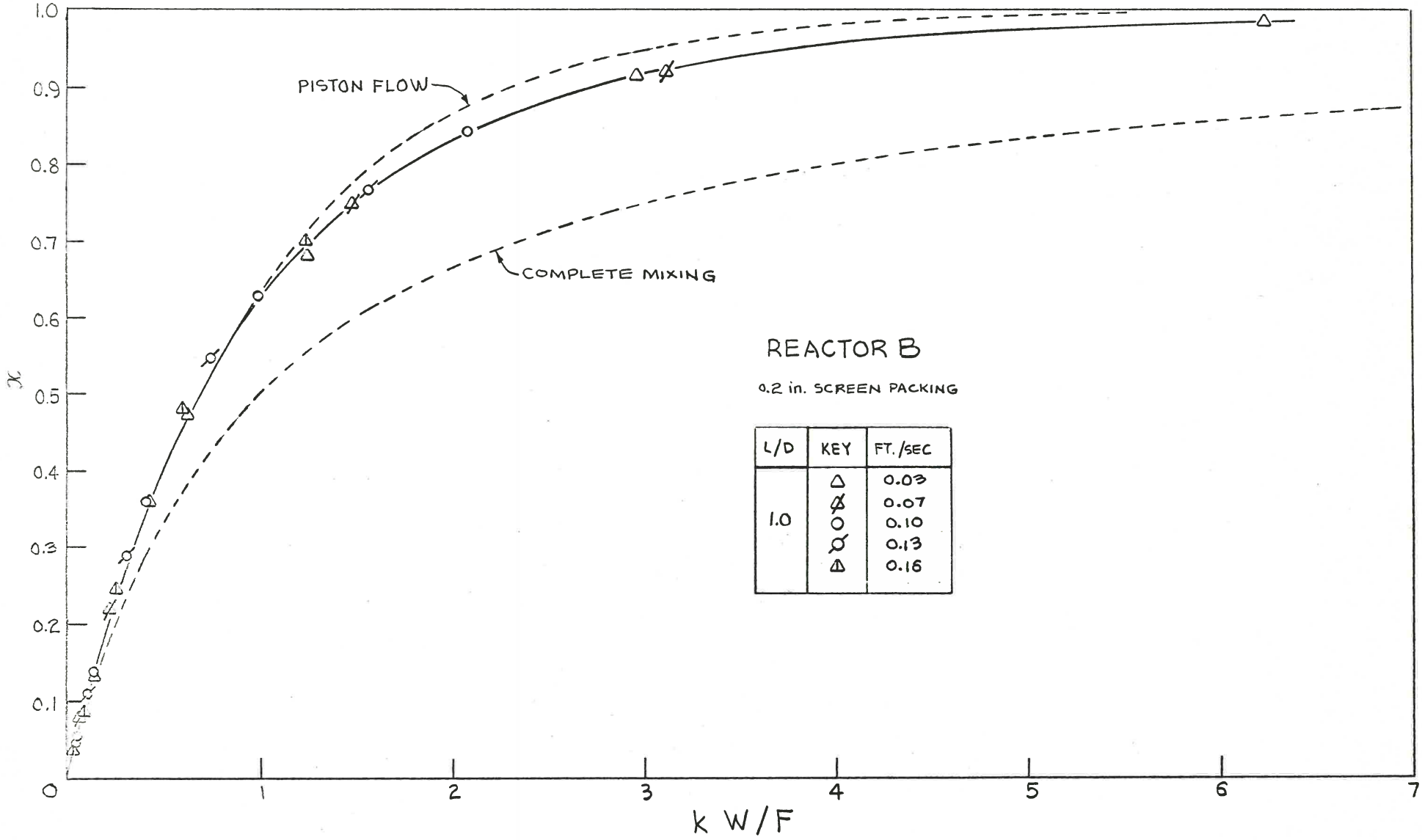


Fig. 16-1 CONVERSION AS A FUNCTION OF k W/F FOR A FLUIDIZED BED WITH SCREEN PACKING

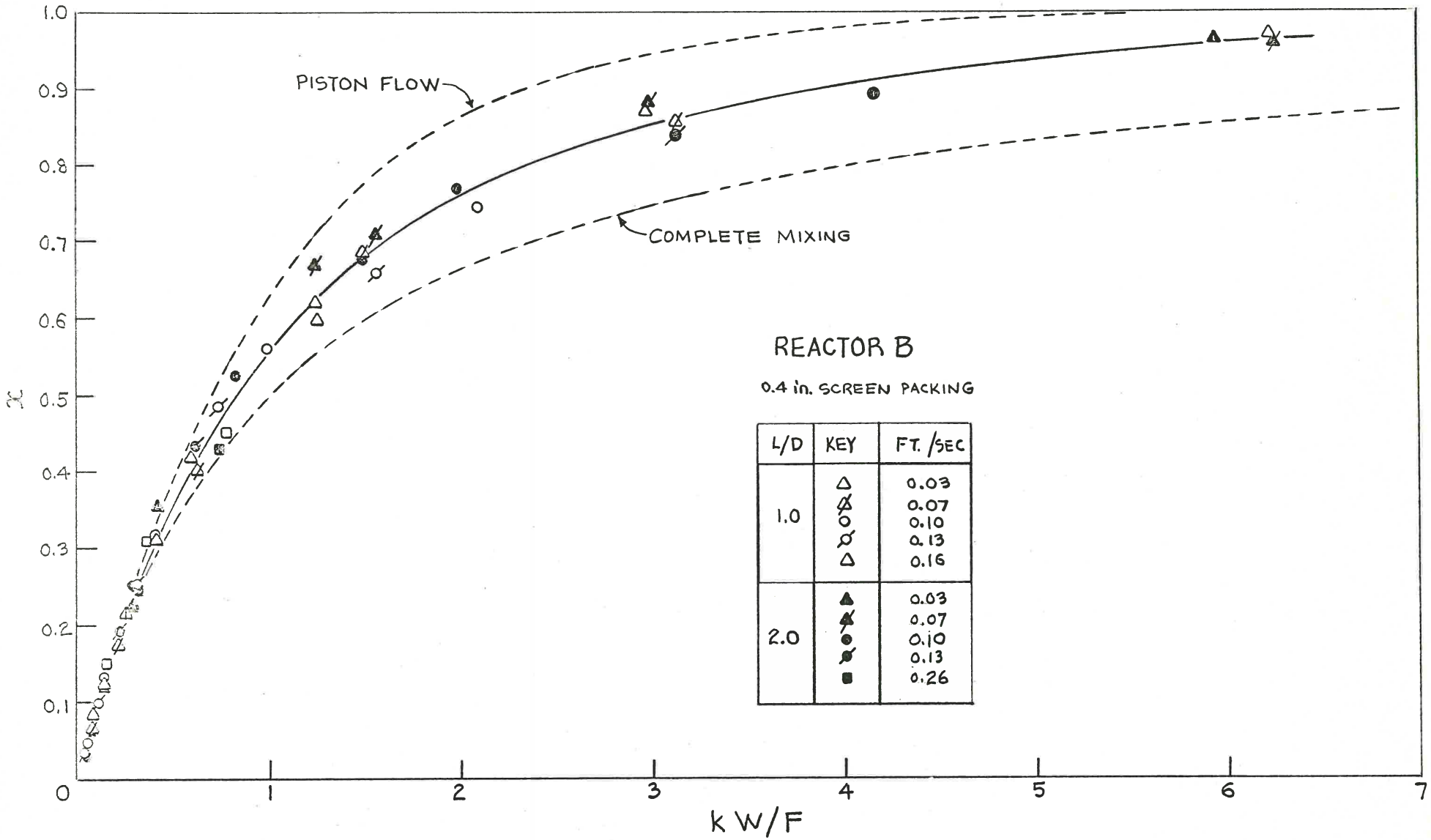


Fig. 16-2 CONVERSION AS A FUNCTION OF kW/F FOR A FLUIDIZED BED WITH SCREEN PACKING

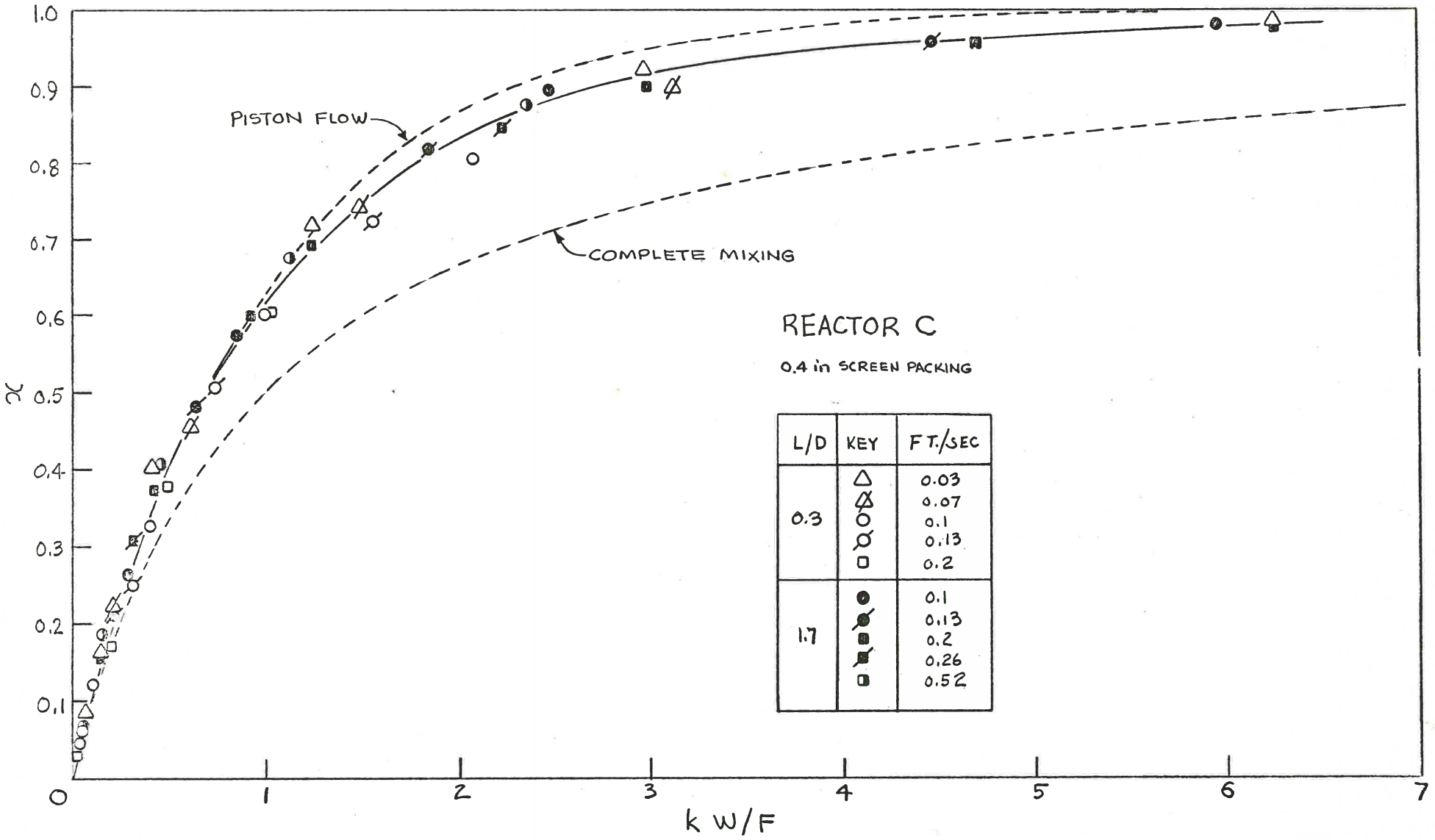


Fig. 16-3 CONVERSION AS A FUNCTION OF k W/F FOR A FLUIDIZED BED WITH SCREEN PACKING

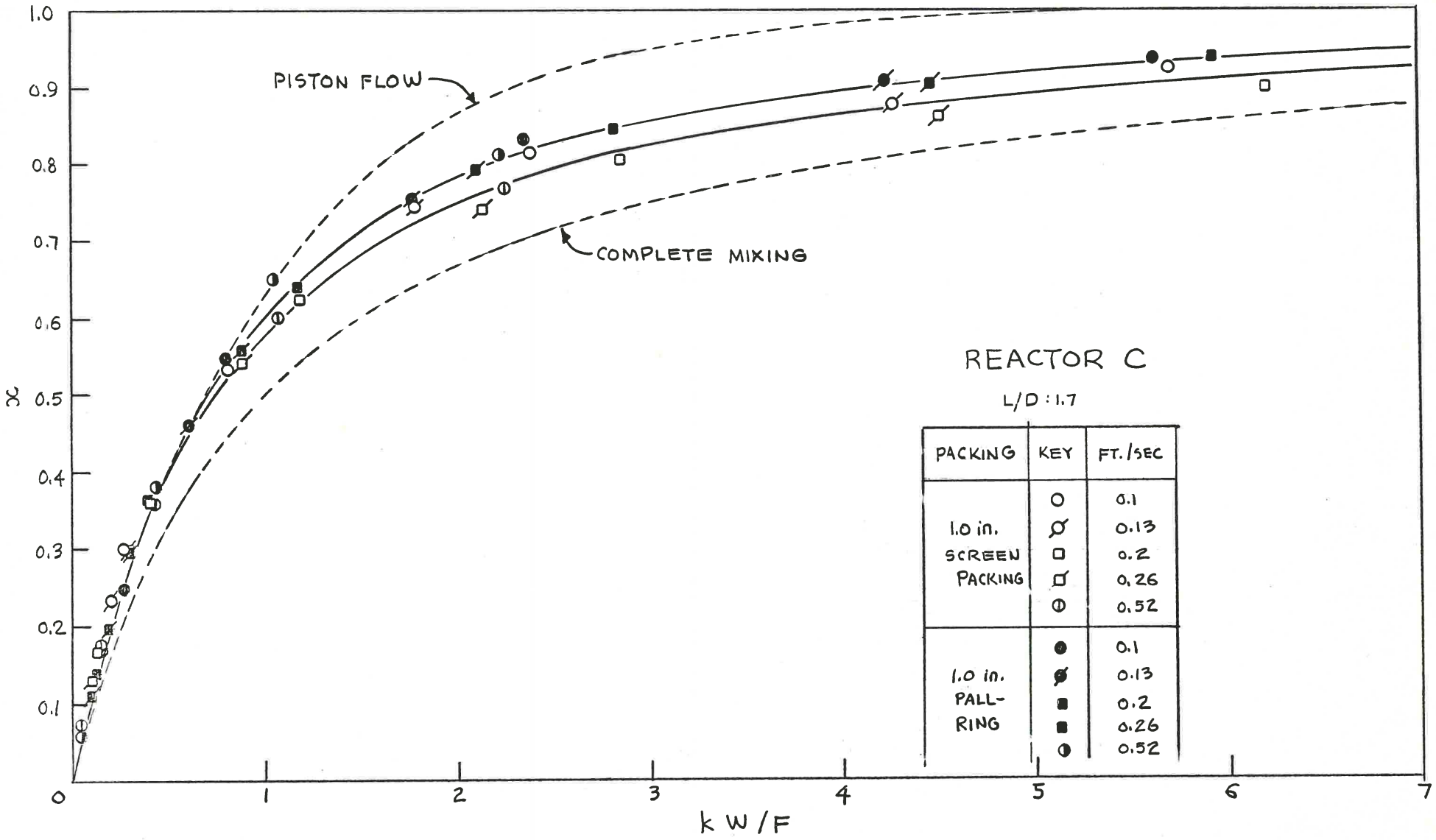


Fig. 16-4 CONVERSION AS A FUNCTION OF k W/F FOR A FLUIDIZED BED WITH SCREEN PACKING

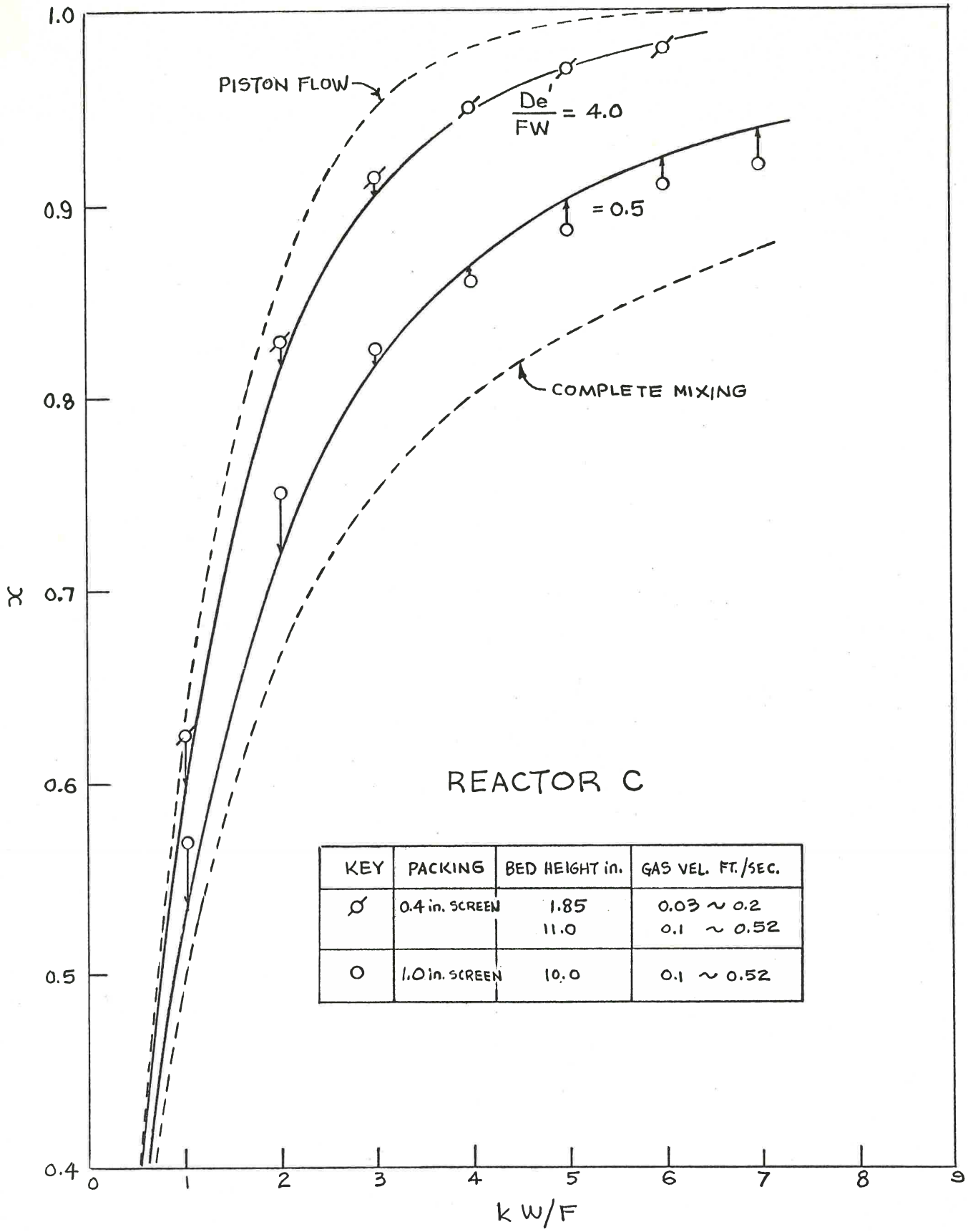
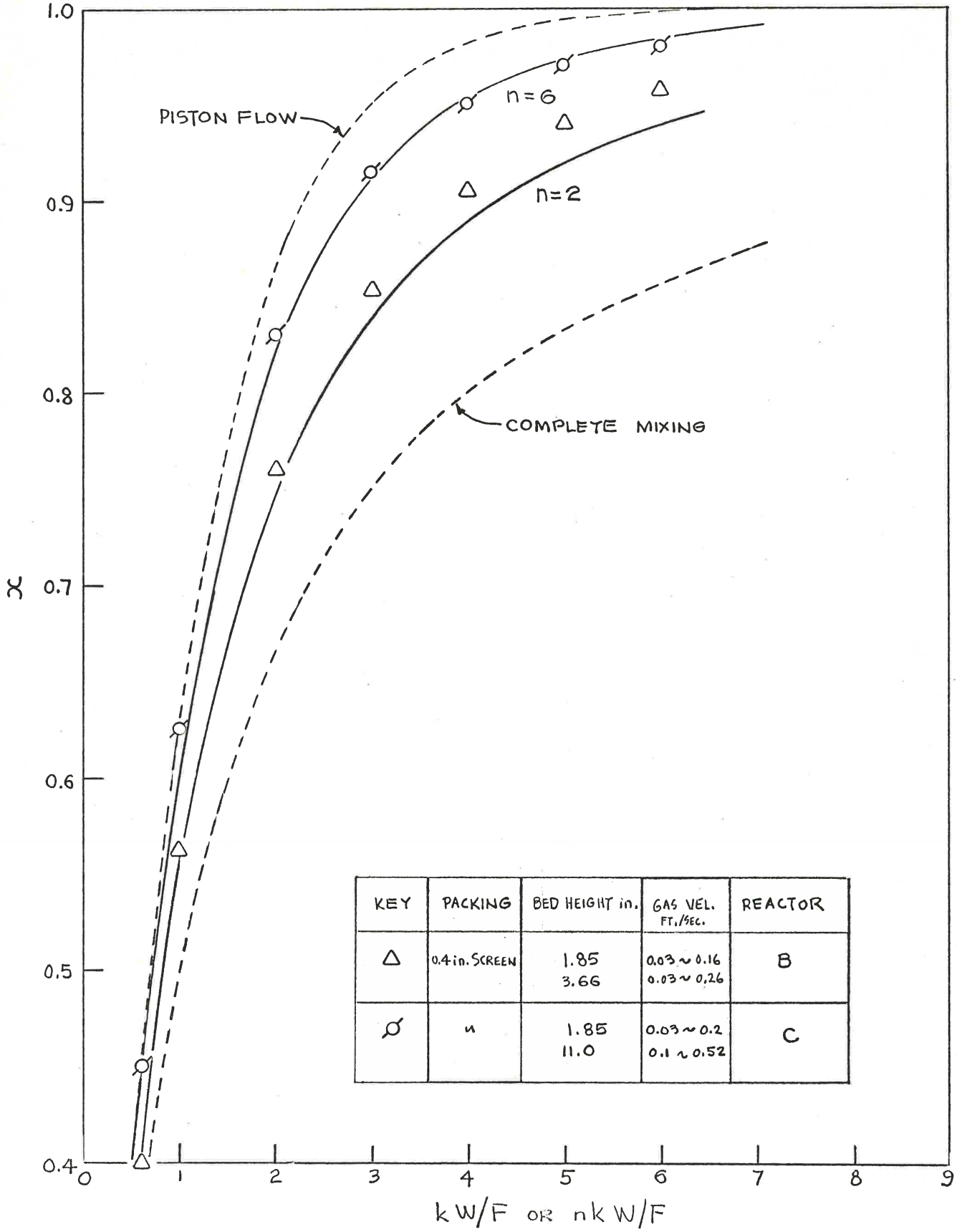
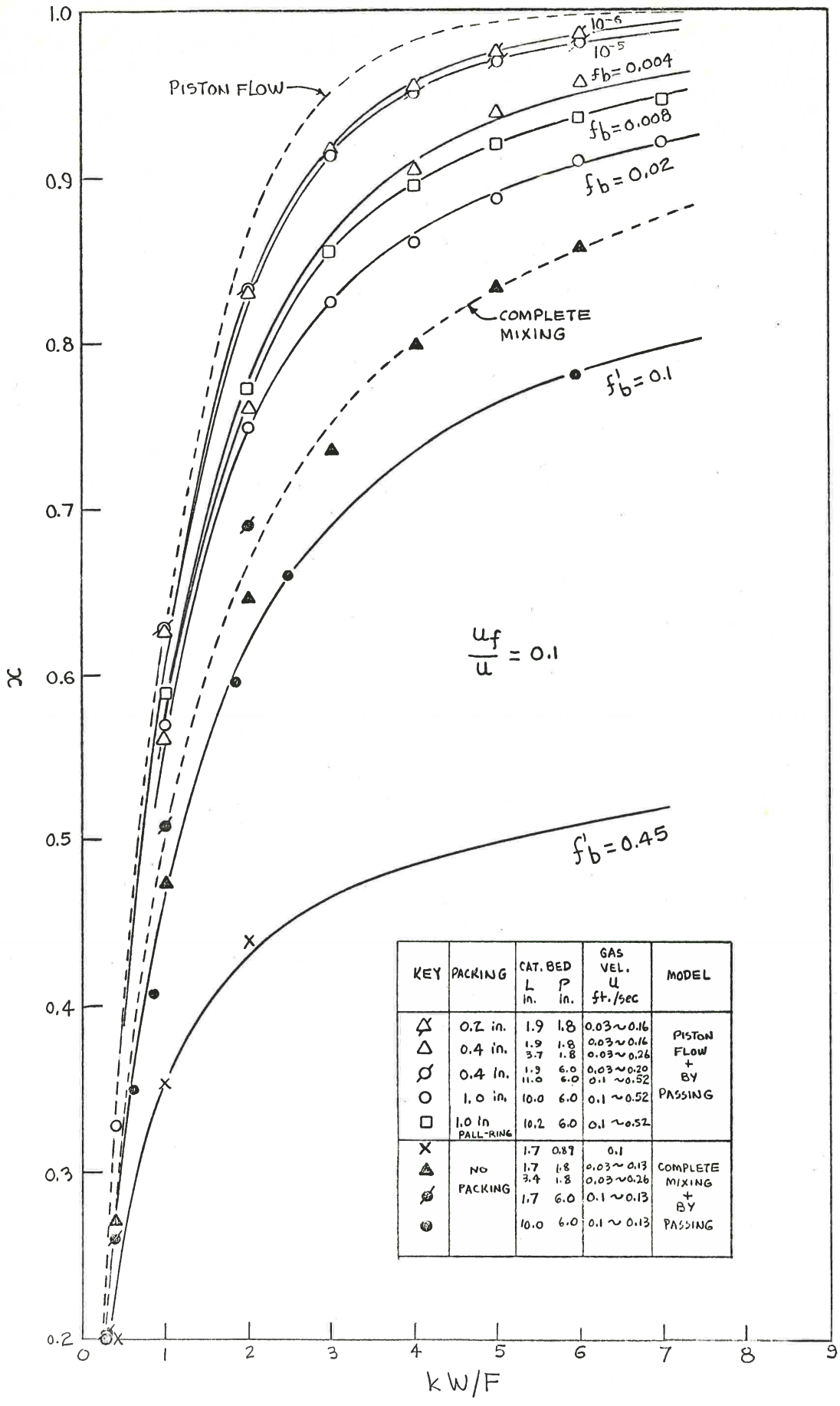


Fig. 17 THE CORRELATION OF DATA FOR FLUIDIZED BED WITH PACKING IN 6 in. REACTOR BY A DISPERSION MODEL.



KEY	PACKING	BED HEIGHT in.	GAS VEL. FT./SEC.	REACTOR
△	0.4 in. SCREEN	1.85 3.66	0.03 ~ 0.16 0.03 ~ 0.26	B
⌀	"	1.85 11.0	0.03 ~ 0.2 0.1 ~ 0.52	C

Fig. 18 CORRELATION OF DATA FOR FLUIDIZED BED WITH 0.4 in. PACKING BY A TANKS-IN-SERIES MODEL.



KEY	PACKING	CAT. BED L in.	CAT. BED P in.	GAS VEL. u ft./sec	MODEL
⧋	0.2 in.	1.9	1.8	0.03~0.16	PISTON FLOW + BY PASSING
△	0.4 in.	1.9	1.8	0.03~0.16	
		3.7	1.8	0.03~0.26	
⌀	0.4 in.	1.9	6.0	0.03~0.20	
		11.0	6.0	0.1 ~0.52	
○	1.0 in.	10.0	6.0	0.1 ~0.52	
□	1.0 in PALL-RING	10.2	6.0	0.1 ~0.52	
×	NO PACKING	1.7	0.89	0.1	COMPLETE MIXING + BY PASSING
▲		1.7	1.8	0.03~0.13	
		3.4	1.8	0.03~0.26	
⌀		1.7	6.0	0.1 ~0.13	
●		10.0	6.0	0.1 ~ 0.13	

Fig. 19 The correlation of data for a fluidized bed with and without packing using the bubble-flow model [12.]

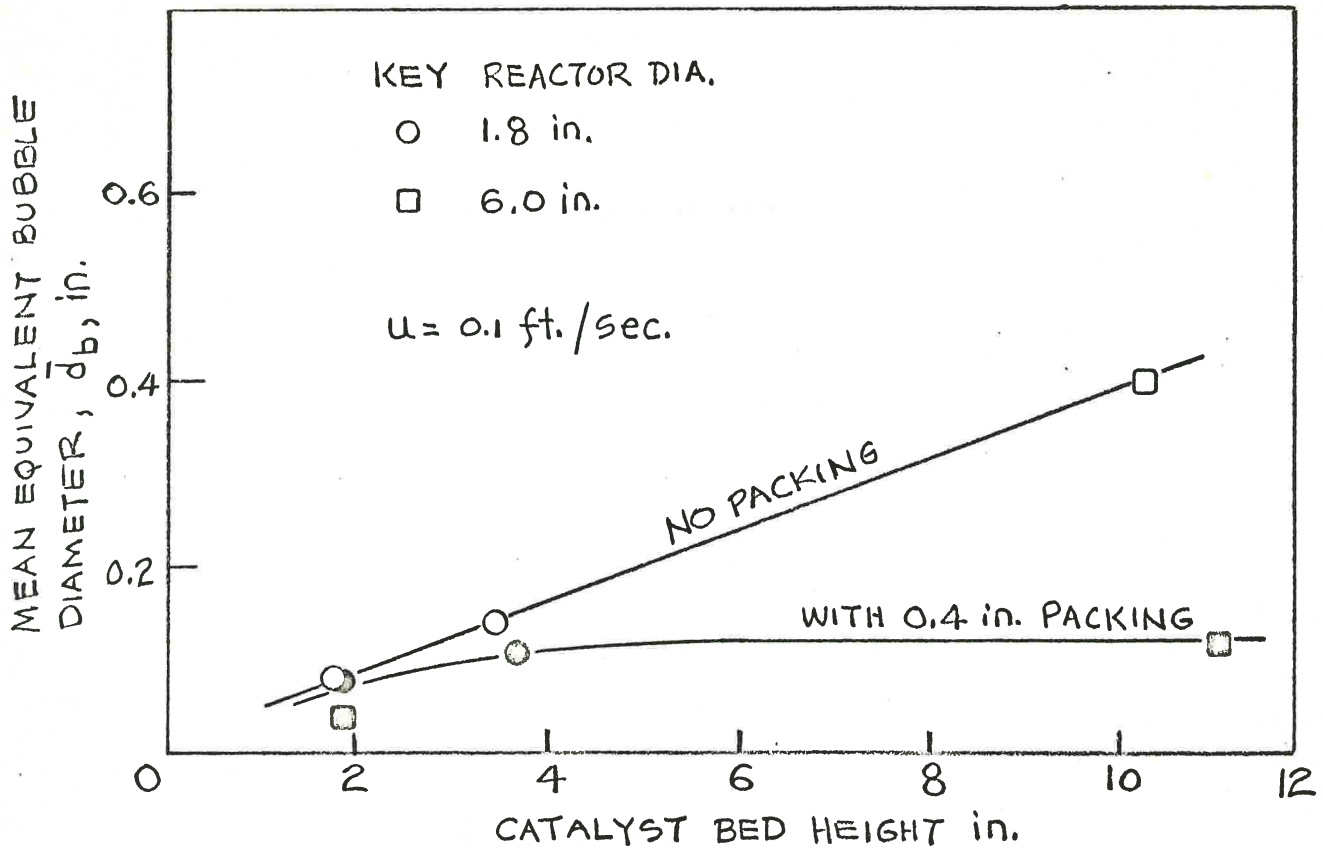


Fig. 20 THE MEAN EQUIVALENT BUBBLE DIAMETER AS A FUNCTION OF BED HEIGHT AT u = 0.1 ft./sec FOR FLUIDIZED BED WITH AND WITHOUT PACKING IN 1.8 in. AND 6.0 in. REACTOR

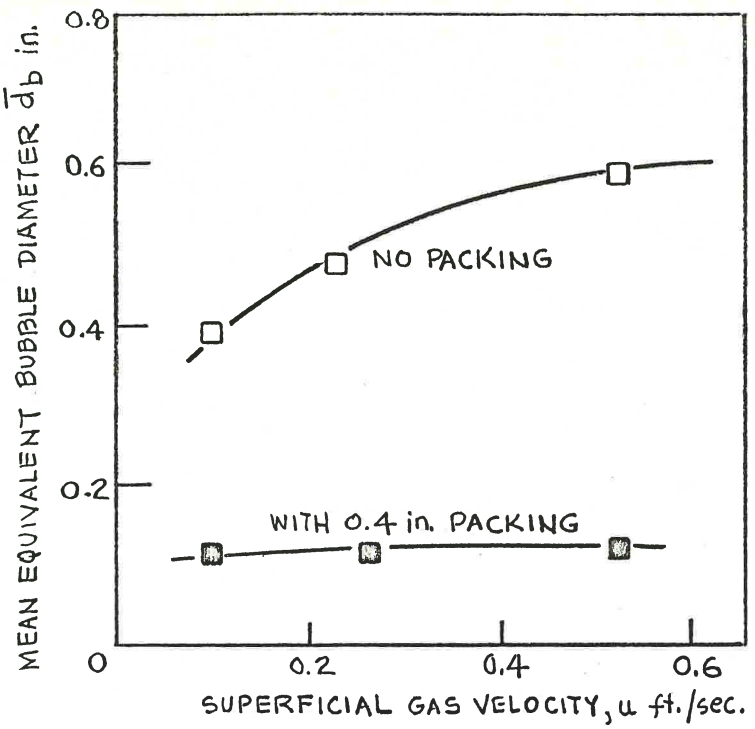


Fig. 21 THE MEAN EQUIVALENT BUBBLE DIAMETER AS A FUNCTION OF GAS VELOCITY AT L= 10.3 ~ 11.0 in. FOR FLUIDIZED BED WITH AND WITHOUT PACKING IN 6.0in. REACTOR.

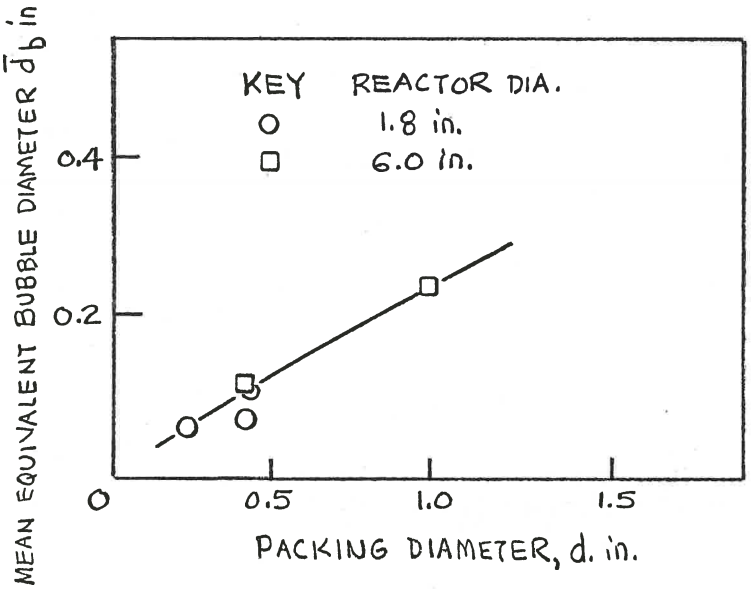


Fig. 22 THE MEAN EQUIVALENT BUBBLE DIAMETER AS A FUNCTION OF PACKING SIZE AT L/D = 1.0 ~ 2.0 AND u = 0.1 ft./sec. FOR FLUIDIZED BED WITH PACKING IN 1.8 AND 6.0 in. REACTOR