MIXING AND ENERGY RELATIONSHIPS IN A FLOCCULATION UNIT

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IN A FLOCCULATION UNIT

by

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SCOPE AND CONTENTS:

The effects of energy input variation on the mixing obtained in a laboratory flocculation unit were studied. The main parameter was the size and shape of paddles used to induce the mixing. A variation in paddle geometry produced a variation in energy input. The effect of baffles was studied and the effect of flocculator geometry was considered by comparing two different length to width ratios for a single tank. Theoretical models were used in an attempt to describe a flocculation unit by comparing the experimental tracer output with the predicted theoretical output.

Some generally accepted design criteria for full-scale flocculation units were also compared with experimental results.

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INTRODUCTION

Coagulation - Flocculation

A principal process in the treatment of water and waste-water is that of coagulation - flocculation followed by sedimentation and/or filtration. With the continual introduction of new organic coagulants and coagulant aids, this clarification process will conceivably increase in importance, and so a better understanding of the chemistry, mechanics and hydraulics involved would be beneficial to improve design prodedures.

Coagulation is the process of reducing the coulombic barrier between particles so that they may join together as aggregates. This is achieved by the addition of electrolytes, coagulants or coagulant aids that either change the surface properties of particles so that chemical attachment becomes possible, or introduce a bridging agent that may attach to two or more particles. The actual aggregation of the destabilized colloidal particles into settleable flocs is termed flocculation. Perikinetic flocculation is the aggregation of particles of submicron size and is promoted primarily by Brownian movement. Orthokinetic flocculation is the aggregation of larger particles and is governed almost entirely by velocity gradients within the flocculator.

Physical Factors Affecting Flocculation

Much has been introduced into the engineering and chemical literatures during the past 50 years describing the chemical reaction and effects of salt concentrations, pH optimums, etc. involved with coagulation. Publications describing the physical factors affecting flocculation have not been so frequent, however, and there is still much to be learned regarding these.

The main physical parameters directly affecting the performance of a flocculation reactor are the residence time (T), its distribution, and the mixing energy input. The currently accepted design criteria for determining the reactor volume and the mixing arrangement are based on the concept that performance is a linear function of T and G, where G is the root mean square (r.m.s.) velocity gradient (sec^{-1}), a function of the energy input. Because G and T are generally accepted as independent, their dimensionless product (GT) is regarded as an adequate design parameter. Argaman and Kaufman (1969), however, state that this parameter is not valid for high values of G. They showed that for a given T, the performance of a flocculator increases almost linearly with G until a maximum value is reached beyond which any further increase results in a decrease in performance.

The power characteristics of mixing impellers have been studied extensively in the literature, particularly by Rushton et al (1950). Unfortunately, the type of paddle wheels that are commonly used in flocculation have not been studied. It is impossible to apply any of these power relationships to flocculator paddles because the rate at which the energy is supplied through the impeller is not only dependent upon the type of impeller used and how rapidly it is rotated, but also on the physical characteristics of the fluid, the shape of the reactor, and the relative location of all component parts of the system such as baffles and supports. Hence, to characterize the behaviour of any impeller, it is necessary to take into consideration the complete environment in which it operates.

LITERATURE REVIEW

Kinetics of Flocculation

The first person to study the theories of flocculation in detail was Smoluchowski (1918). He published a mathematical theory for the coagulation of colloidal suspensions describing the rate of change in particle concentration after the repulsive forces between particles were reduced sufficiently so that interparticle collisions were not hindered. He developed two kinetic equations. The first described the kinetic process when interparticle collisions occurred through Brownian diffusion. The second described the process when a laminar shear gradient causes particle transport at a point in the fluid. The usual form of the second equation is

$$H_{ij} = \frac{4}{3} (r_i + r_j)^3 \quad n_i n_j \frac{du}{dz} \tag{1}$$

where H_{ij} is the frequency of collision between i-fold and j-fold particles,

n_i, n_j are the numbers of i-fold and j-fold particles per unit volume of fluid,

r_i, r_j are the radii of the i-fold and j-fold particles, and

du/dz is the laminar velocity gradient.

Subsequently, Langelier (1921) demonstrated that agitation was necessary to obtain adequate flocculation in water purification, and so

Smoluchowski's equation is not truly valid for turbulent flocculation.

Camp and Stein (1943) were the first to introduce the theory of orthokinetic flocculation into the sanitary engineering literature. They realized that for practical reasons turbulent flocculation was of most importance in water treatment. By using the basic ideas of mechanics and introducing their own version of the value of Stoke's "dissipation function", they defined a root-mean-square (r.m.s.) velocity gradient for turbulent flow:

$$G = \sqrt{\frac{W}{\mu}} \tag{2}$$

where G is the r.m.s. velocity gradient in the flocculation chamber,

μ is the absolute viscosity of the fluid, and

W is the mean value of the dissipation function and is equal to the total power dissipation into the chamber divided by the volume of the chamber.

Camp and Stein also demonstrated theoretically that the speed of flocculation is directly proportional to the velocity gradient at a point, and so G was substituted in Smoluchowski's equation for the velocity gradient existing in laminar flow. Equation (1) now becomes

$$H_{ij} = \frac{4}{3} (r_i + r_j)^3 n_i n_j G \tag{3}$$

or, as is more often seen in the literature,

$$H = \frac{1}{6} (d_1 + d_2)^3 n_1 n_2 G \qquad (4)$$

Equation (4) states the relation between energy applied to the flocculation chamber as described by the velocity gradient, G, and the number of contacts, H, between n_1 particles of diameter d_1 and n_2 particles of diameter d_2 in unit time.

In general, equation (4) can be used to study the rate at which small floc particles join to form larger floc particles or it can be used to study the rate at which suspended particles in the raw water are entrapped by floc particles.

Evidence for experimental and/or theoretical verification of equation (4) has been presented by several workers. For example, Saffman and Turner (1956) used the concept of collision frequency between particles in a turbulent flow field for the special case of isotropic turbulence. Their final equation is in the form

$$H_{ij} = 1.30 (r_i + r_j)^3 n_i n_j G \qquad (5)$$

where $1 \leq r_i/r_j \leq 2$.

Frisch (1956), Levich (1962), Swift and Friedlander (1964), and Higuchi (1963, 1965) are others who have verified Smoluchowski's work. All these authors have arrived at similar kinetic equations.

Equation (4) has found wide application among sanitary engineers for design and operation purposes despite its several theoretical deficiencies. Some of these have recently been examined by several investigators.

Hudson (1965) stated that one of the most significant limitations of equation (4) is that it describes the reaction of only two sizes of particles. He attempted to avoid the restriction by assuming a monodispersed floc developing in a field of primary particles. He used equation (4) to analyze the rate at which particles in the raw water are entrapped by floc particles and defined the terms in the equation as follows:

n_1 is the number of suspended particles naturally present in a unit volume of water,

d_1 is the average diameter of these particles,

n_2 is the number of floc particles in the water, and

d_2 is the average diameter of these floc particles.

Based on studies by Robeck (1963) and Riddick (unpublished), Hudson postulated that the effect of d_1 on the equation was small since n_2 is only a small fraction of n_1 in flocculating water. Consequently, d_1 could be omitted with little error. He also introduced a "sticking ratio" to account for the phenomena that not all particles adhere to a floc upon collision. The "sticking ratio" is defined as the proportion of primary particles which stick to the floc particles upon collision. His final mathematical relation describing the kinetics of flocculation is in the form

$$\frac{n_i^1}{n_i^o} = e^{-(\frac{\phi}{6} n_2 d_2^3 GT)} \qquad (6)$$

where n_i^o, n_i^1 are the primary particles in the flocculation influent and effluent, respectively,

ϕ is the sticking ratio, and

T is the residence time.

This relation indicates that the entrapment of suspended matter by floc is influenced by the volume of floc produced rather than by the size or appearance of the floc particles since the volume of floc per unit volume of water V_2, is:

$$V_2 = n_2 \frac{\pi d_2^3}{6} \qquad (7)$$

Neither the Smoluchowski model nor the models used to verify the Smoluchowski equation consider the possibility of simultaneous floc aggregation and breakup due to the shear and turbulent forces which are generated in the reactor. Little consideration had been given to the effect of an upper limit of floc size on the rate process until Fair and Gemmell (1964) developed a mathematical model for floc growth by imposing various upper particle size limits and floc breakup modes on Smoluchowski's equation for flocculation in a laminar shear field.

Computer solutions indicated a particle size growth pattern that approached a steady-state mean size regardless of the velocity gradient-particle concentration product, provided the maximum particle volume and breakup mode were fixed. The particle growth pattern became oscillatory, however, as the velocity gradient-particle number product was increased, presumably because of the increased rate of particle aggregate breakup and reformation when the gross particle contact frequency was increased.

In recent work by Harris, Kaufman and Krone (1966), attention was directed to equations expressing the rate of change of primary particles when the maximum size of the floc was limited. They also continued the work of Hudson (1965) by deriving a size distribution function describing the mutual flocculation of all particles comprising the size distribution. Their final rate equation for a series of continuous stirred-tank reactors is

$$\frac{N_1^m}{N_1^0} = \left(1 + \frac{KDGN_1^0T}{m}\right)^{-m} \qquad (8)$$

and for a batch reaction is

$$\frac{N_1^1}{N_1^0} = e^{-KDGN_1^0T} \qquad (9)$$

where N_1^0 is the mass concentration of primary flocs in the influent,

N_1^1, N_1^m are the mass concentration of primary flocs in the effluent of one reactor and m reactors in series, respectively,

D is the experimentally determined rate variable which is analogous to their size distribution function, and

K is their experimentally determined reaction constant.

Harris et al (1966) used the mass concentration N_1^0 and N_1^1 instead of n_1^0 and n_1^1, the number concentration, as the measurement of the degree of flocculation because they could not devise a satisfactory means of experimentally determining the number concentration of primary particles in a partially flocculated system. The mass of floc suspension after a period of settling could be measured easily. They demonstrated experimentally the first order dependence of N_1^0/N_1^1 upon G for two different values of T, upon T for two different values of G, and upon N_1^0 for two different values of GT.

All of the work done thus far in the field of orthokinetic flocculation indicated that the performance of a flocculator is dependent upon G, T, and the distribution of T, but no author has indicated that there is an interdependence among these three factors.

Design Factors for Flocculation Units

Factors and standards used in the design of flocculating units are not so well established as in the case of other main treatment processes, and adequate performance data, in general, are lacking. The investigators of performance of flocculators differ somewhat in their criteria for design. It is worthwhile to mention some of these criteria and the basis for these values.

(a) Tank Shape and Mixing Devices:

A variety of combinations of flocculating devices and sedimentation tanks have been used in the past, either as separate units or as combined in rectangular or circular tanks. In most present pretreatment plants, the raw water is thoroughly mixed with the coagulant, after which it passes to rectangular basins where it is slowly mixed to build up a satisfactory floc. This slow mixing is accomplished by diffused air, by baffled chambers using around-the-end or over-and-under baffles, or by mechanically operated paddle wheel mixers of the horizontal- or vertical-shaft type.

Diffused-air flocculators are costly to operate for the following reasons as outlined in the WPCF Manual of Practice No. 8 (1967). For good flocculation, the air bubbles should be distributed uniformly throughout the volume of the unit and should be small enough so that the velocity gradients close to the bubbles are not great enough to disrupt the floc. If these precautions are not taken, the results are not of equal quality to those obtained with mechanical devices. Experience indicates that much of the floc is carried to the surface as a scum much like an air flotation unit. Therefore, addition and maintenance of skimming equipment adds to the cost of diffused-air flocculators.

Baffled mixing chambers attain good results at a fixed feed rate, but if the flow varies they are not satisfactory since G is proportional to u in the following manner for a mixing chamber of the channel type or a rectangular settling tank (Camp and Stein (1943)):

$$G = \sqrt{\frac{f}{\nu}\,\frac{u^3}{8R}} \tag{10}$$

where f is the friction factor,

u is the average channel velocity over the cross section area,

ν is the kinematic viscosity, and

R is the hydraulic radius.

At low flow flocculation is incomplete and deposits occur in the chamber, while at high flow the loss of head is excessive and velocities may be so rapid as to break up the floc. Baffled flocculators were very common at one time but their use is decreasing because of their inflexibility, high head loss, and construction cost.

Mechanical mixers are used most frequently today. They give constant rotational velocities with a minimum of head loss regardless of the rate of flow through the basin, and, therefore, within the usual variations of flow encountered at most plants, the water leaving these mechanical mixers is always uniformly flocculated. The vertical-shaft mixers consist of a vertical shaft with radial arms attached, to which paddle blades are fastened. The horizontal-shaft type is of similar construction, but uses a horizontal shaft placed either parallel to the direction of flow through the basin or at right angles to it.

Camp (1955) calculated W, G, and GT for twenty of the flocculation basins existing in the United States. As a result of these calculations and some theoretical developments, he stated that, since higher velocity gradients may be used for small floc particles rather than for large particles, the best economy should result where flocculation is carried out in several stages, in a series of tanks, with the velocity gradients progressively decreasing as the floc particles grow in size. This procedure was first developed by Langelier (1921) and is known as the "Langelier process". The sum of the GT values for the series of basins should be the same as the sum for a single basin to obtain satisfactory floc, but the detention period, and hence the cost for the series, should be less.

Because the mixing process normally used in a flocculator short-circuits some of the water quickly from the entrance to the outlet port of each basin, several tanks should be placed in series so that the liquid which passes through the upstream tank will have a chance to stay in the downstream tanks for a longer period.

The most common type of flocculation unit is rectangular in shape, consisting of several compartments separated by baffles, and mixed by the use of vertical- or horizontal-shaft mechanical mixers consisting of paddle blades fastened to radial arms extending from the shaft.

(b) Paddle Design:

There are several important considerations in paddle wheel design and application. One of these considerations is the clearance of the periphery of the wheel between walls, bottom, and liquid surface. With a great mass of liquid at the periphery of the wheel, a certain mixing speed is essential to cause circulation of the liquid from the center to the outside of the wheel where it displaces the liquid at that periphery. Also, the greater the distance of the blade from bottom and walls, the greater is the speed required to obtain the sweeping action required to prevent deposition. According to Nichols (1940), increased speed causes more rapid dilution (dispersion) of incoming liquid and consequently short-circuiting through the compartment. Therefore, the less the clearance, the less is the necessity for paddle speed. After a series of tests, Tolman (1942) suggested that the areas swept by the paddle wheels should be not less than 65 per cent of the cross-sectional area of the basin. Nichols (1940) claimed that although the clearance of walls and bottom, and depth under the liquid level varies with diameter and speed, for good results and minimum speed requirements this clearance should not be greater than five per cent of the wheel diameter.

The distribution of paddle blades throughout the mass of liquid being treated is important. The intensity of mixing is inversely related to detention time, and, therefore, increasing the intensity of mixing will reduce the detention time necessary. Paddle widths are definitely limited, however, if rolling of the water with the paddle wheels is to be avoided. On the other hand, paddles may not be too narrow, otherwise, the eddies produced along the edges will be weak or ineffective. This results from the fact that with narrower paddles greater proportions of the flow past paddle edges become laminar in character, and only turbulent flow is effective in mixing (Bean (1953)). There has been no definite effective length-to-width ratio mentioned in the literature.

Bean (1953) stated that paddle spacing should be sufficiently wide that water in front of the paddles may readily pass around to the rear of the moving paddles. If the spacing is too small for the paddle widths, then roll must result. Similarily, high velocities between the paddles may break some of the floc already formed. Bean suggested that the area of the paddles may generally be in the vicinity of 12 to 20 per cent of the cross-sectional area of the tank. 25 per cent will produce major rolling or rotation. In order to facilitate the design of stirring mechanisms, most tanks with paddles mounted on vertical shafts are made circular or approximately square, according to Camp (1955).

The speed of the paddle wheels depends upon at least two other elements: clearance between paddle wheel and walls and bottom, the depth under the liquid level and the distance between blades measured circumferentially. The larger the wheel, the greater is the distance between blades in their path of travel. Higher peripheral speeds are required with larger wheels to compensate for the greater distance between blades. Further, the larger the wheel diameter the greater is the floor area outside of the space where blades are in close contact. Therefore, higher peripheral speeds are necessary to produce a sweeping action that will prevent deposition when using horizontal-shaft paddles.

Despite these factors, most authors can give limits to the peripheral speed of paddles based on experimental work with actual flocculators. Wilcomb (1932) stated that the peripheral velocity should be in the range of 0.3 to 1.0 ft/sec. As a rule of thumb, Nichols (1941) suggested that a 5-foot diameter wheel having a sufficient number of paddle blades in each section and a proper clearance of walls and bottom, and depth under the liquid level, should be operated at approximately 1 ft/sec peripheral speed in the first section and approximately 0.5 ft/sec in the last section. An 18-foot diameter wheel should be operated at higher speeds, approximately 2 ft/sec in the first section and 1 ft/sec in the last section. Peripheral speeds vary from 0.75 to 1.50 ft/sec according to Tolman (1949). Some of the standard text books in the sanitary engineering field also give general ranges of peripheral speeds which should produce best results. Rich (1961) suggested 0.5 and 3 ft/sec, Clark and Weissmann (1965) 0.3 to 3 ft/sec, and the Water Pollution Control Federation Manual of Practice No. 8 (1967) suggested 0.9 to 1.2 ft/sec.

(c) Detention Time:

The length of the flocculation period varies according to the quality of the raw water. Some investigators and some standard text books, however, have given values of detention times that have worked satisfactorily in the past for most types of raw water. Tolman (1949), Bean (1953), and Rohlich and Murphy (1961) suggested detention times of 40 to 60 minutes for good flocculation. Rich (1961) suggested 10 to 30 minutes, the WPCF Manual of Practice No. 8 (1967) 15 to 30 minutes, and Linsley and Franzini (1964) suggested 20 to 30 minutes.

(d) Inlet Speed and Port Velocity:

Once formed, flocculated particles are quite fragile and must be treated gently until settled. Inlet and outlet disturbances must be held to a minimum, and so velocities in and after the flocculating unit generally are not allowed to exceed the peripheral paddle speed, (WPCF (1967)).

Nichols (1940) felt that port velocities through baffles should not exceed 1.5 ft/sec, otherwise floc would be broken or shredded. He found that a port velocity of 1 ft/sec for maximum flow is sufficient for creation of good hydraulic conditions, even though the flow may vary considerably.

Use of Tracers

The type of tracers used in different types of waters to investigate vessel flow-through characteristics has changed over the past 20 years. Prior to 1950, salts or dyes were used extensively. Archibald (1950) introduced the idea of using radio-active tracers, or "radiotracers", in an attempt to avoid errors due to chemical action of the salt with the water's contents and errors due to density changes. His paper listed the main advantages of using radiotracers:

1) They give a more accurate picture of the actual flow-through curves of basins and conduits by bringing out the longer tails of the curves which sometimes go undetected by using salts or dyes.
2) The power of these tracers are unaffected by changing chemical and physical conditions and can be detected in very low concentrations - much lower than is possible with dye or salt.
3) Density effects are improbable as only very small amounts of the material are required.
4) In many cases, especially in model studies, the radiotracer may be followed through the system without disturbing the flow pattern by taking readings through the pipe or tank walls.

These advantages over salts were verified experimentally by Thomas and Archibald (1952).

The 1960's introduced the widespread use of fluorescent tracers. Carpenter (1960), and Pritchard and Carpenter (1960) used these tracers and their results were of the same accuracy as radioactive tracers.

Thus, the main objection to radiotracers, the hazard involved to the personnel handling them, was removed.

Feuerstein and Selleck (1963) carried out a laboratory investigation to determine the behaviour of the fluorescent tracers Rhodamine B, Pontacyl Brilliant Pink B, and fluorescein in waters of various quality. Effects of temperature, salinity, pH, background level, and turbidity on the analytical determinations of the tracers were ascertained. Pontacyl Brilliant Pink B demonstrated no absorption on suspended sediments, whereas Rhodamine B exhibited significant absorption. Fluorescein has exceedingly high photochemical decay rate and natural background levels. Therefore, it should be used in waters of only the highest quality. Although the cost of Pontacyl Brilliant Pink B is higher than Rhodamine B its use in most cases for water quality conditions generally encountered is believed to be justified for attainment of meaningful results from tracer studies. For laboratory purposes, however, when using tap water for short-term studies, the lower cost of Rhodamine B justifies its use over Pontacyl Brilliant Pink B.

Another new fluorescent tracer is Rhodamine WT. It was compared with the three previous fluorescent tracers by Wilson (1968). It is far less susceptible to sorption than Rhodamine B, although its fluorescent properties are similar. It is more susceptible to sorption than Pontacyl Billiant Pink B, however. Because it is not as readily available as Rhodamine B, this latter tracer is still recommended for short-term laboratory work using tap water.

Mixing Models for Reactors

There has been a great deal of information in the literature concerning the flow characteristics of chemical reactors. Although the ideal mixing extremes are plug flow and complete back-mix flow, real reactors never fully satisfy the conditions implied in these extremes. In many cases the deviation from ideality can be considerable. This deviation can be caused by the channeling of fluid through the vessel, by the recycling of fluid within the vessel, by the existance of stagnant or "dead" regions, or by a combination of any of these conditions.

One of the many types of models that can be used to characterize non-ideal flow patterns within reactor vessels is the dispersion model (dispersed plug flow). The foundation for this approach is the assumption that the mixing process involves a redistribution of material either by slippage or eddies, and this occurrs enough times to be statistical in nature. Therefore, an equation analogous to Fick's law of diffusion may by used:

$$\frac{\partial C}{\partial t} = \frac{D \partial^2 C}{\partial x^2} \qquad (11)$$

where D is the longitudinal or axial dispersion coefficient characterizing the degree of back-mixing during flow through the vessel.

The terms "longitudinal" and "axial" are used to distinguish mixing in the direction of flow from mixing in the lateral or radial direction, which is not considered. Many authors, including Danckwerts (1953), Wilhelm (1953), Taylor (1953, 1954), and Wehner and Wilhelm (1956), have pointed out that longitudinal mixing can be treated like diffusion, and much has been done to develop the results of such a treatment.

Levenspiel and Smith (1957) have expanded this concept to show that the actual parameter which correctly characterizes the role played by dispersion is the dimensionless group D/uL, the reciprocal of the Peclet number. This group is called the vessel or reactor dispersion number where L is the length of the vessel and u is the average flow velocity through the vessel. Levenspiel and Smith have also shown that the variance of the experimental response curve can be used to determine the value of the dispersion number. The relationship between the variance and the dispersion number depends on the end conditions of the vessel. For a closed vessel, Van der Laan (1958) has shown this relationship to be

$$\sigma^2 = 2\left(\frac{D}{uL}\right) - 2\left(\frac{D}{uL}\right)^2 \left(1 - e^{-\frac{uL}{D}}\right) \qquad (12)$$

The variance, or second moment about the mean, can be determined from the experimental curves by using the following formula:

$$\sigma^2 = \frac{\sum x_i^2 \ f(x_i)}{\sum f(x_i)} - \mu^2 \tag{13}$$

where μ is the mean or first moment about the origin,

and is determined by using the following formula:

$$\mu = \frac{\sum x_i \ f(x_i)}{\sum f(x_i)} \tag{14}$$

This technique of equating variance with the dispersion number is inadequate, however, when using ordinary laboratory equipment to measure the response curve for a flocculation unit. Timpany (1967) has indicated that even for a relatively low degree of dispersion (D/uL = 0.85), accurate readings to 4.9 detention times are required to yield 80 per cent of the actual D/uL, and using common laboratory equipment, accurate readings at this level are not usually possible. Because a flocculation unit has a high degree of dispersion, it is not possible to use this standard variance technique effectively.

The tanks-in-series model is an alternate approach to the dispersion model. Stein (1940) developed the equation for the instantaneous dispersion curve for n tanks of equal size in series:

$$\frac{C}{C_o} = \frac{n^n}{(n-1)!}\left(\frac{t}{T}\right)^{n-1} e^{-nt/T} \tag{15}$$

where C_o is the initial concentration at the inlet.

For a single tank (ie. n=1), equation (15) reduces to the basic equation for a first-order reaction:

$$\frac{C}{C_o} = e^{-t/T} \tag{16}$$

The variance of the tanks-in-series model is given by the equation:

$$\sigma^2 = \frac{1}{n} \tag{17}$$

The value of n can be found from experimental variance measurements. Combining equations (12) and (17), a relationship between the tanks-in-series model and the dispersion model can be obtained:

$$\frac{1}{n} = 2\left(\frac{D}{uL}\right) + 2\left(\frac{D}{uL}\right)^2 (1-e^{\frac{-uL}{D}}) \tag{18}$$

Levenspiel (1962) concluded that the agreement between these two models is not clear-cut. For small values of D/uL, and, hence, large values of n, the models are similar, but for larger deviations from plug flow, the response curves for the two models differ more and more in shape. Because a flocculation unit has a very large deviation from plug flow, the tanks-in-series model cannot be used to determine the degree of dispersion or mixing within the unit. It may be possible, however, to use equation (15) alone for predicting the performance of a flocculation unit. This method would eliminate the discussed problems associated with variance and dispersion index.

LABORATORY APPARATUS AND TECHNIQUES

The laboratory model flocculation unit and associated equipment are shown schematically in Figure 1. The flocculation unit, including baffles, was constructed of three-eighths inch acrylic plastic with dimensions as shown in Figure 2. When used, the baffles were fitted in slots giving three equal compartments, each 5 inches by 5 inches by 7 inches high. The baffle part-holes were 3/8 inches in diameter. These baffles were sealed with stop-cock grease to prevent hydraulic short-circuiting around the baffles. When operating, the unit had a volume of 6 liters, giving a water height of approximately 5 inches.

The paddles were constructed from one-eighth inch acrylic plastic. Figure 3(a) is a drawing of one of the two arms of a paddle with the attached blades. The paddles were designed in accordance with the criteria summarized in the previous section. The paddles were driven by small permanent magnet synchronous motors (115V, 60c, 27W), products of Cramer Electromechanical Products. They had a constant speed of 60 revolutions per minute giving a paddle peripheral velocity of 1.177 feet per second.

The power input to the water was determined by using spring balances (0-100 grams), products of Chatillon Company. The torque applied to the water by the paddles was the difference between the spring balance readings in water and air times the perpendicular

Figure 1: SCHEMATIC OF APPARATUS

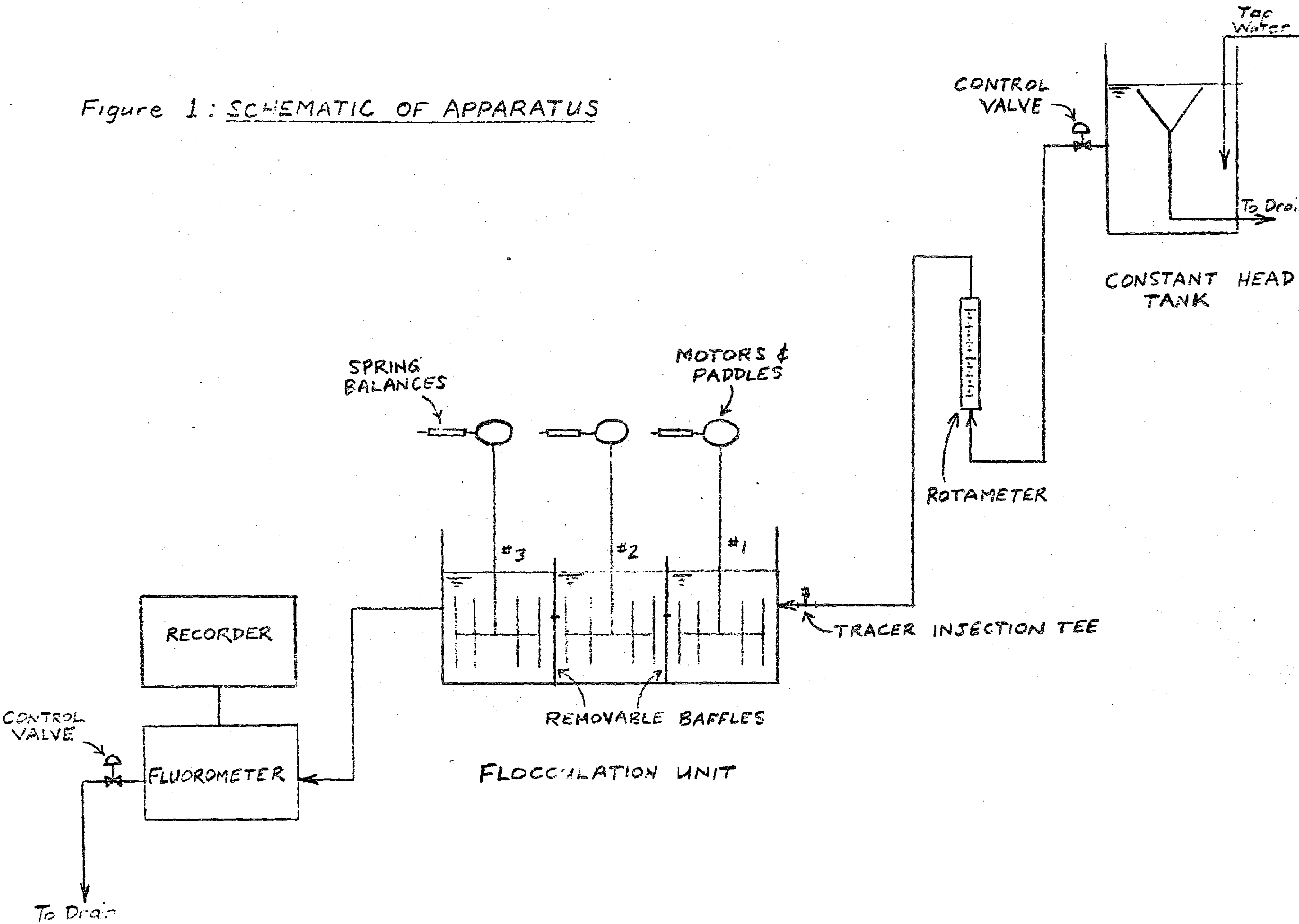
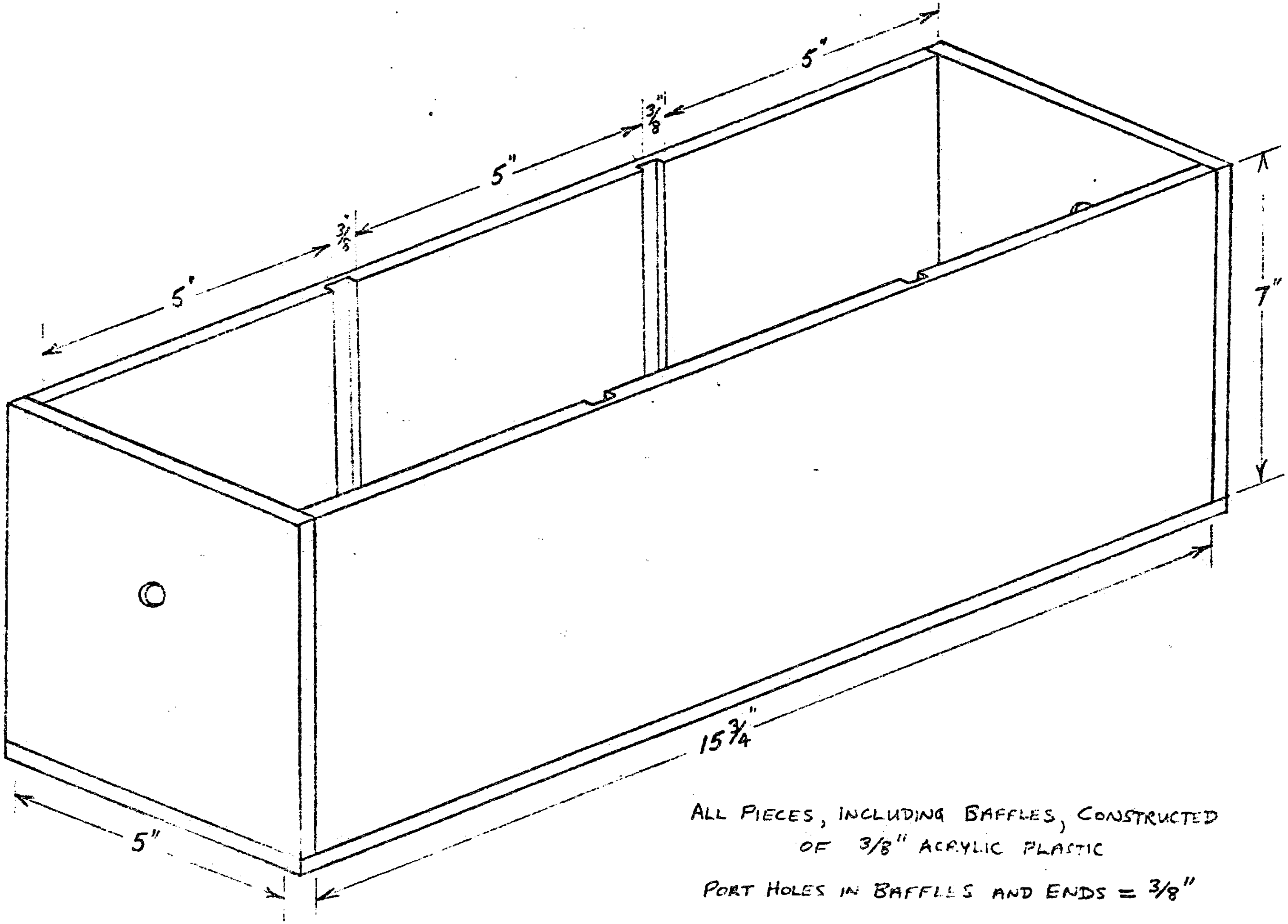


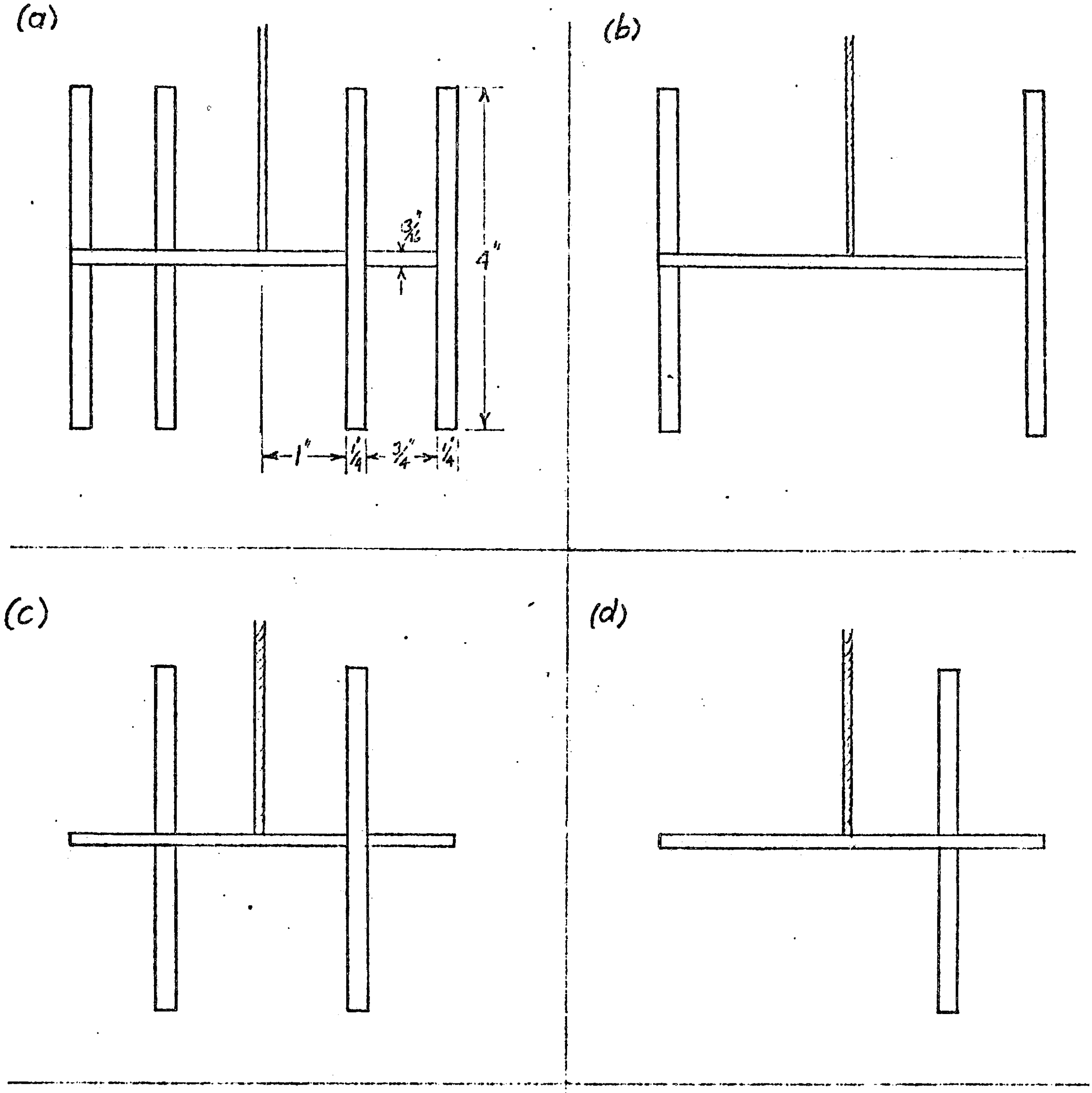
Figure 2: MODEL FLOCCULATION UNIT

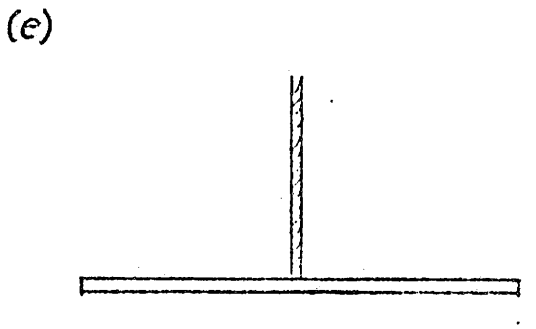


ALL PIECES, INCLUDING BAFFLES, CONSTRUCTED OF 3/8" ACRYLIC PLASTIC

PORT HOLES IN BAFFLES AND ENDS = 3/8"

Figure 3: PADDLE DESIGN & ARRANGEMENTS





PADDLE ARRANGEMENTS:

#1. TWO OF (a)

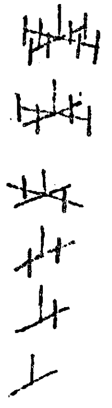
#2. ONE OF (b) & ONE OF (c)

#3 ONE OF (c) & ONE OF (e)

#4 ONE OF (c)

#5 ONE OF (d)

#6 ONE OF (e)



PADDLES IN EACH OF THE 3 COMPARTMENTS HAD THE SAME ARRANGEMENT FOR EACH RUN distance between the center of the motor shafts and the respective projection lines of the spring balance arms.

To begin an experimental run, the flocculation unit was filled with tap water to the desired height giving a volume of 6 liters for 3 compartments and 2 liters for 1 compartment. The two control valves were then adjusted so that the desired flow rate and the desired reactor volume were maintained. After this equilibrium had been reached, a small volume of dye was injected by a syringe into the injection point as illustrated in Figure 1. This injection point consisted of a glass tee covered by a rubber membrane. The fluorescent dye used was Acid Rhodamin B, a product of Allied Chemical. This method of dye injection was considered to represent an ideal pulse function because the duration of injection was negligible compared to the detention time of the unit. The time of dye injection was marked on the recorder. The effluent was passed through the fluorometer and the fluorescence readings were recorded. The holdup time in the fluorometer was considered during the data analysis. Thus, continuous sampling of the effluent was obtained. The fluorometer used was a G. K. Turner Association, Model 111. It was calibrated at 18°C using the 1-60 primary filter and the 23A secondary filter in the 1X range. These filters and this range were used throughout the experimental work. A temperature correction was necessary during the data analysis as the runs were not completed at 18°C. The recorder used was a Honeywell "Electronik 19".

In order to decrease the energy input for successive runs, some of the blades and arms of each paddle were removed. For each run, all three paddles had the same physical construction. In all cases the speed of the mixers was maintained at a constant 60 rpm. The paddle arrangements used are illustrated in Figure 3. Some runs used all three compartments, while others used one compartment only. For the latter cases, the inlet was connected to the second baffle so that compartment three was used as the representative single tank.

During each run, the following data were recorded: flow rate, reactor volume, dye volume and concentration, fluorometer effluent temperature, spring balance readings, and the physical arrangement of the flocculation unit as well as the experimental response curve. The spring balance readings were also recorded prior to and following each run when the unit was empty so that the readings in air were available.

DATA ANALYSIS

During the laboratory tests, a recorder strip chart continuously recorded the flocculator effluent concentration. Each resulting curve was transferred to I.B.M. data cards taking readings at equal intervals. A computer programme was written to calculate the dimensionless response curve, the area under the curve and the mean of the curve. A typical programme is reproduced in Appendix A. Each programme included several corrections. The length of tubing from the reactor exit to the fluorometer was 56 inches, and so a correction was made to reduce the chart time to the actual time by this delay. There was no attempt to control the temperature of the water, and so the temperature of the fluorometer effluent was recorded. Using the same type of fluorometer that was used in this study, Feuerstein and Selleck (1963) found through experimental procedures that the fluorescence-temperature relationship could be described mathematically as:

$$C_\theta = C_s e^{k(\theta_s - \theta)} \qquad (19)$$

where C_θ is the fluorescence at any temperature, θ,

C_s is the fluorescence at a standard temperature, θ_s, and

k is a rate constant in reciprocal temperature units.

The rate constant was found to be $0.027^{o}C^{-1}$ for Rhodamine B by these same investigators. Butts (1969) verified their work and found a value of $0.026^{o}C^{-1}$ for k. Therefore, a value of $0.0265^{o}C^{-1}$ was used in this study. Θ_s was $18^{o}C$. The values of C_s were determined from the fluorometer calibration curves shown in Appendix B.

The method of dye injection contributed to what was probably the greatest source of error, the value of C_o used for each run. The area under each dimensionless response curve should have been unity, but the areas were generally greater than unity. Therefore, the values of C/C_o were adjusted by the corresponding value of the experimental area to bring all the areas to unity. An exception was made for the runs involving a single tank only. Because the response curves for these runs were truncated at approximately 2.2 residence times, the adjustment described above could not be made. Therefore, for these runs the maximum value of C/C_o for each run was adjusted to a value of unity, and all the other points on the same curve were adjusted by the same ratio.

Table 1 lists the variables used for each run, the calculated values of G (the root-mean-square velocity gradient) and the mean under the curve. All of the experimental runs are not included in Table 1 because some were aborted due to experimental difficulty while others were replicates of some of those listed. Runs with residence times of 30 minutes only are included. There were not enough experimental runs conducted at the other residence times of 10 minutes to 90 minutes from which to draw any conclusions, and so they were omitted from this report. 30 minutes was chosen as the operating retention time since

Table 1: EXPERIMENTAL RESULTS

Run	Temp. (°C)	Number of Compartments	Baffles	Mixing	Paddle Arrangement	G ($sec.^{-1}$)	Mean (a)	Mean (b)
4	12.95	3	yes	no	-	—	—	—
9	14.35	1	-	yes	1	99.2	0.705	—
14	8.15	3	yes	yes	1	106.0	—	0.960
18	12.85	1	-	yes	2	88.8	0.750	—
19	7.95	3	yes	yes	2	96.3	—	0.994
20	8.3	3	no	yes	2	134.0	—	0.875
21	9.4	3	no	no	-	—	—	—
22	9.65	3	no	yes	3	73.9	—	0.910
24	9.6	3	yes	yes	5	43.3	—	0.995
25	9.8	3	yes	yes	6	36.3	—	0.995
26	10.8	1	-	yes	6	30.1	0.748	—
27	11.3	1	-	yes	4	42.4	0.774	—

Notes: 1. (a) = Mean for 2.2 residence times only.
(b) = mean for complete response curve.

2. Residence time for all runs = 30 minutes.

3. Paddle arrangement illustrated in Figure 3.

this value is within the various ranges described in the literature as the recommended retention time.

The values of G were calculated by using equation (2). W, the mean value of the dissipation function, was calculated for each run by using the following relationship:

$$W = \frac{S}{V} \cdot T_p \qquad (20)$$

where T_p is the torque input for each mixer,

S is the speed of each mixer,

V is the total volume of the vessel, and

W is the power input per unit volume.

T_p was determined from the spring balance readings and the perpendicular distance from the center of the mixer shaft to the line of force of the balances. The spring balances were calibrated using weights, but the readings were identical to the weights and so no calibration curve was necessary.

Figure 4 shows the effect of baffling during mixing. For comparison purposes, the response curves for one continuous stirred tank reactor (C.S.T.R.) and three C.S.T.R.'s are also shown. Equation (15) was utilized in drawing these latter two curves. With baffles, the experimental response curve is similar to the three C.S.T.R.'s curve, and without baffles, the experimental curve approximates the one C.S.T.R. curve. Baffling, therefore, has an important effect when mixing is being used in a flocculation unit.

The effect of baffling is not so pronounced, however, when there is no mixing induced by the stirrers. Figure 5 illustrates that

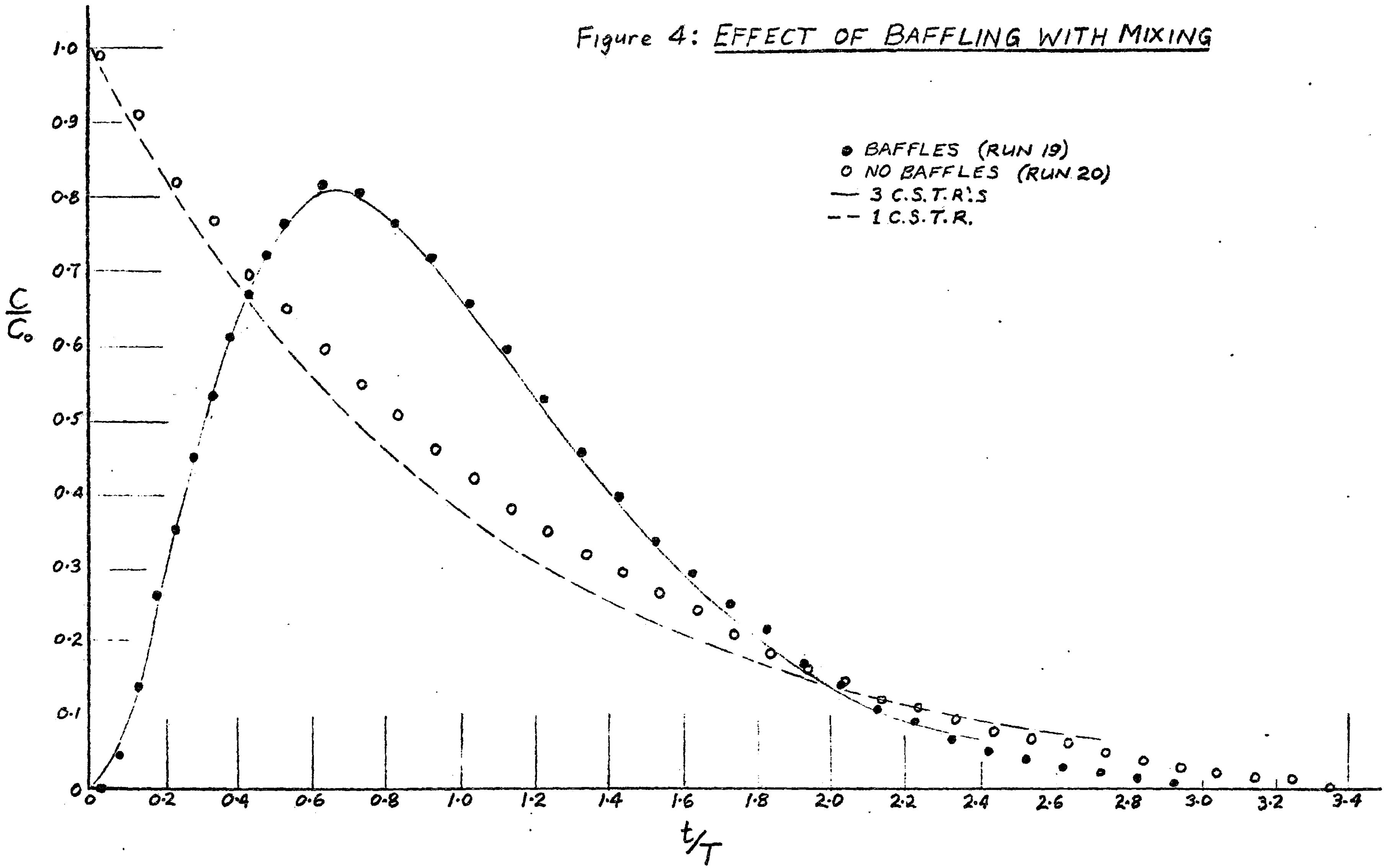


Figure 4: EFFECT OF BAFFLING WITH MIXING
BAFFLES (RUN 19)
NO BAFFLES (RUN 20)
3 C.S.T.R.'S
1 C.S.T.R.
C/C₀
t/T
0
0·1
0·2
0·3
0·4
0·5
0·6
0·7
0·8
0·9
1·0
0·2
0·4
0·6
0·8
1·0
1·2
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2·2
2·4
2·6
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3·0
3·2
3·4

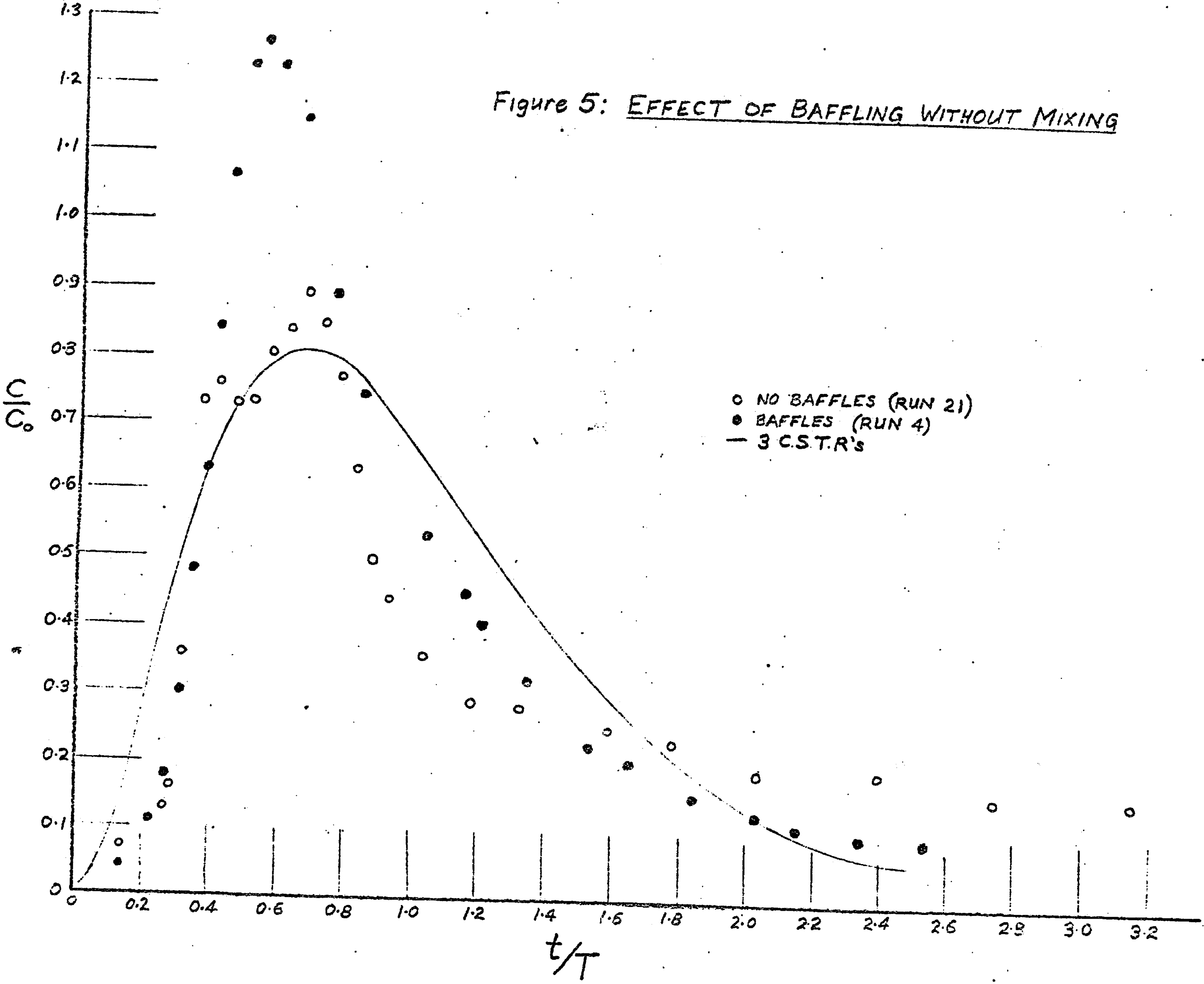
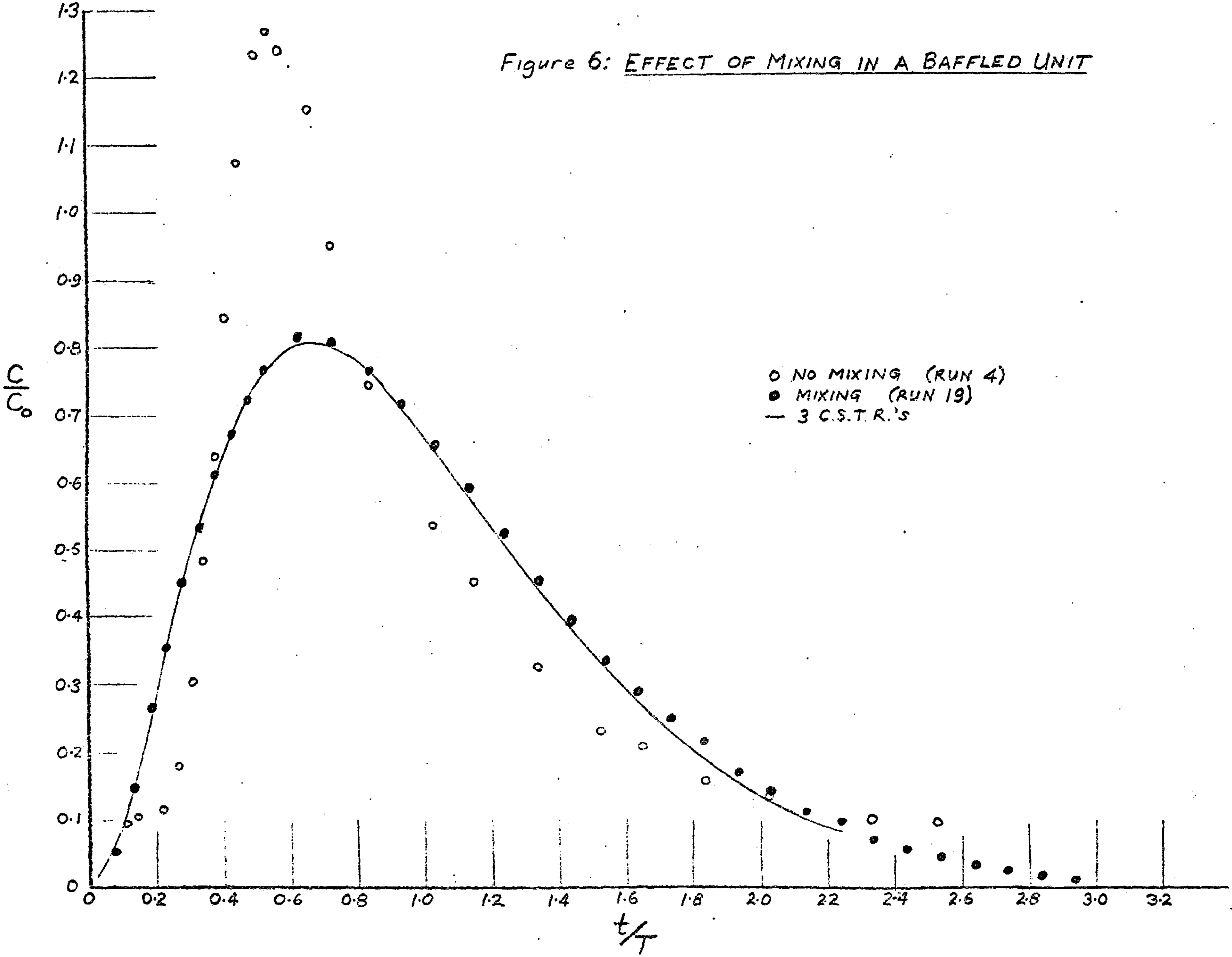


Figure 5: Effect of Baffling Without Mixing

Figure 6: EFFECT OF MIXING IN A BAFFLED UNIT



the peaks of both curves rise and fall sharply in both cases, indicating short-ciruiting, although there is much more short-circuiting in the case utilizing baffles. This phenomenon could be seen easily during the experimental runs. After the dye injection, the dye passed from one port almost directly through each compartment to the next port with very little dispersion. Small baffles were placed about one-third of an inch downstream from the ports to give a small amount of dispersion, although they did not prevent this severe short-circuiting to any great extent. There was a greater amount of dispersion of the dye without baffles than with baffles as indicated by the shorter peak and larger slope of the tail of the response curve.

Figure 6 combines Run 19 of Figure 4 and Run 4 of Figure 5 to illustrate the effect of mixing when baffles were used. There is a great deal of short-circuiting without baffles while baffling causes very little short-circuiting.

Because Figures 4, 5, and 6 show that the experimental curves using baffles with mixing approximate the curve for the three C.S.T.R.'s, an attempt was made to utilize the tanks-in-series model in order to predict a response curve for any energy input. Figure 7 shows the response curves for n equal tanks-in-series for n = 1, 2, 3, ...7. The locus of the peaks of these curves has been drawn in Figure 8, illustrating the relationship between the number of equal tanks-in-series and the dimensionless peak time, t_p/T. This curve was used as the basis for attempting to correlate the tanks-in-series model with the experimental curves.

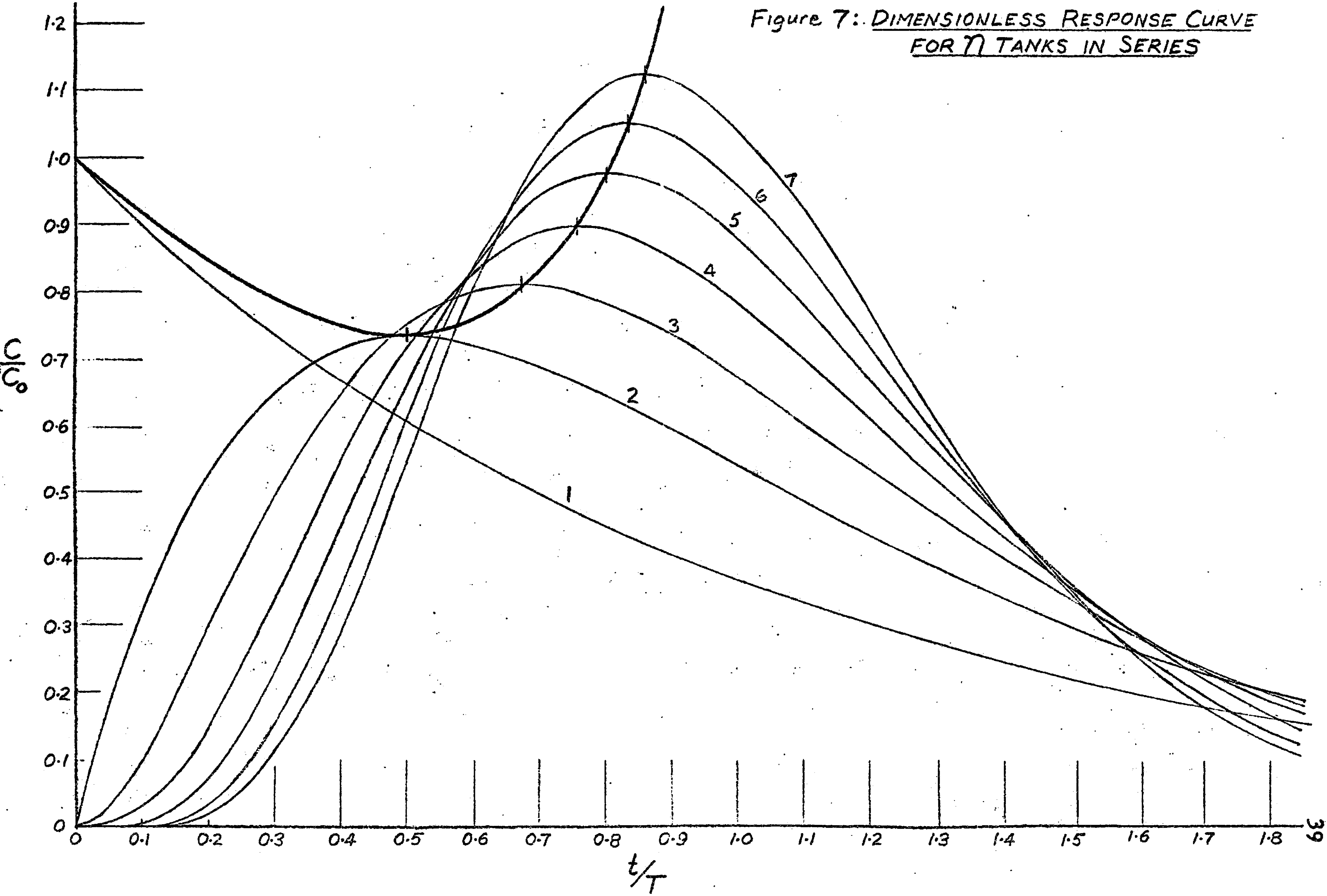


Figure 7: DIMENSIONLESS RESPONSE CURVE FOR n TANKS IN SERIES

Figure 8: Occurrence of Peak Time Associated With n

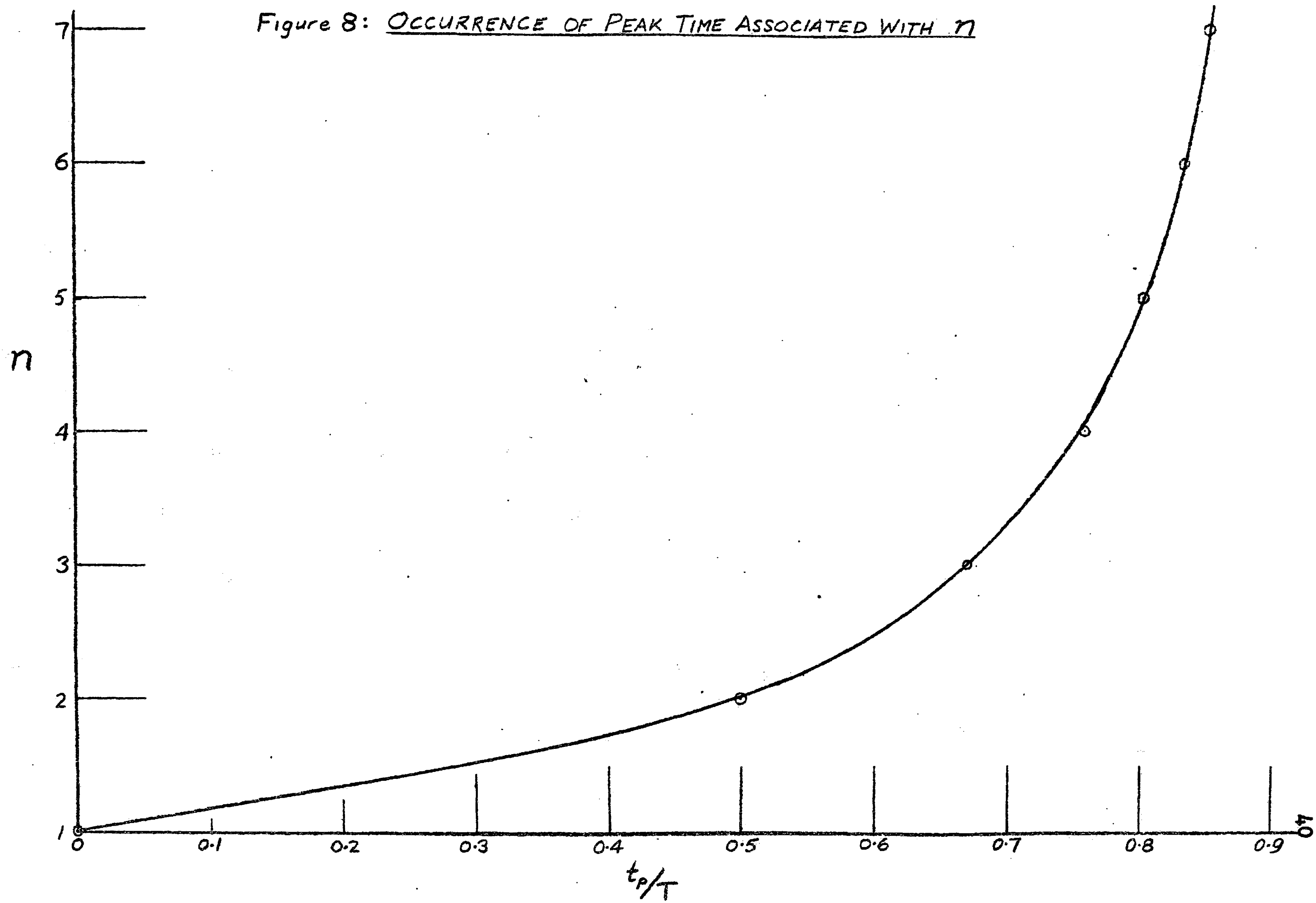
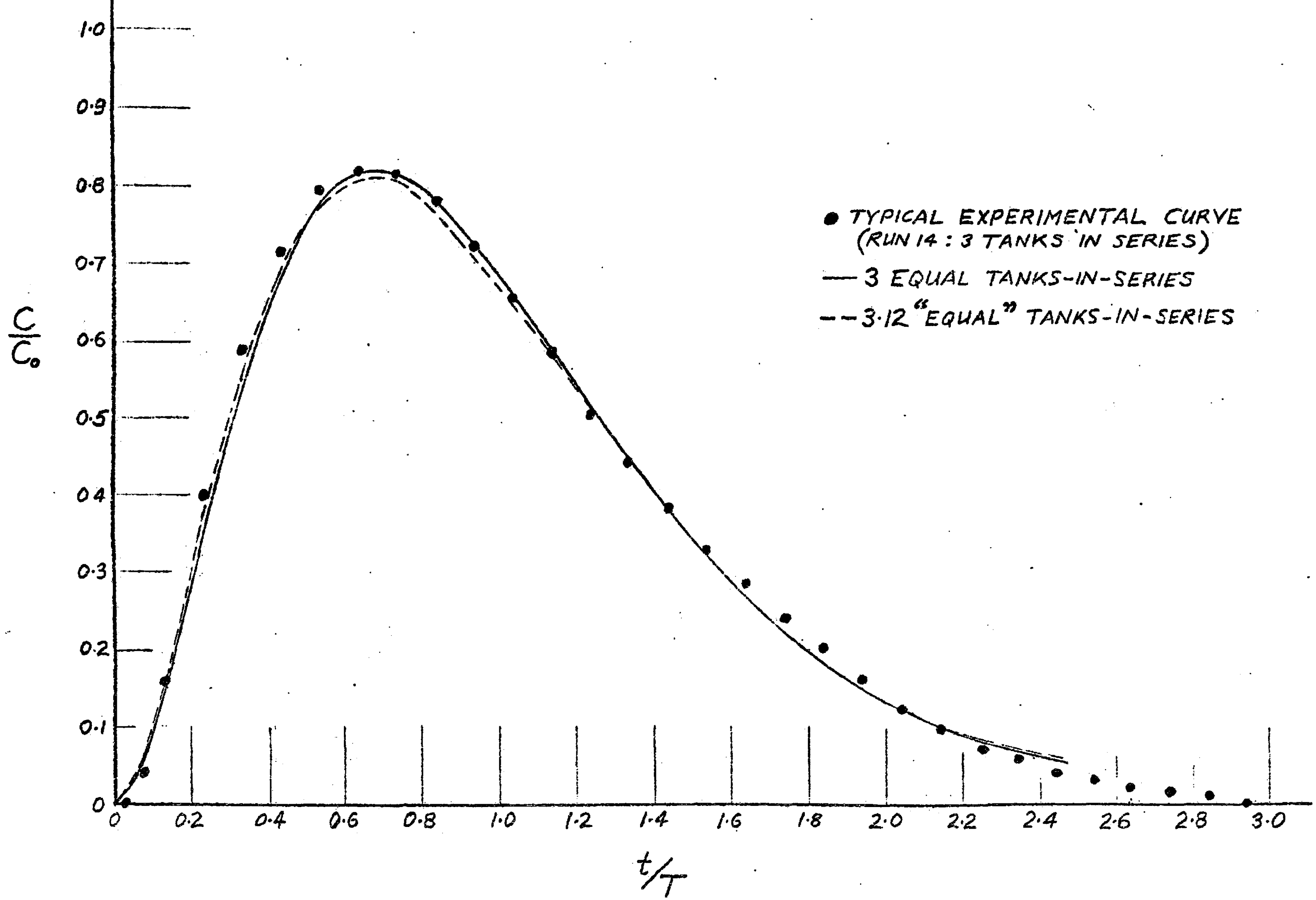


Figure 9 : COMPARISON OF TANKS-IN-SERIES MODEL WITH EXPERIMENTAL CURVE



Run 14 was reduced to dimensionless co-ordinates and plotted in Figure 9. The peak time occurred at a value of $t/T = 0.68$. Reffering to Figure 8, $n = 3.12$ for $t_p/T = 0.68$. Equation (14), the equal tanks-in-series model, was then used to estimate the response curve for 3.12 tanks-in-series. The gamma function was used to estimate the non-integer factorial by the following relationship:

$$\Gamma(n) = (n-1)! \qquad (21)$$

Although this relationship is true for integers only, it can be used as a very close approximation for non-integers, particularly for small values of n. The computer programme used for this curve is reproduced in Appendix A.

The response curves for three tanks-in-series and 3.12 tanks-in-series are plotted in Figure 9 with Run 14 for comparison purposes. Although the areas under the theoretical curves and the experimental curve are almost identical and the peaks appear at the same time, the curves are definitely not identical. The experimental curve shows a higher peak and a more rapid decline than the theoretical curves. This type of phenomenon is to be expected because the theoretical curves assume perfect mixing in all three compartments, and perfect mixing is not obtained in the experimental unit.

The criticisms of the tanks-in-series solution and the dispersion method could be overcome by using a "mixed model". This method consists of combining plug flow sections and tanks-in-series sections along with such things as by-passing, recycling and deadwater regions until the theoretical curves very closely approximate the experimental curves.

However, as Levenspiel (1962) pointed out, this generalized model has the disadvantage that as the number of parameters increase, the model may have very little correspondence with actual conditions, and

> "an unrealistic many-parameter model may closely fit all present data after the fact, but may be quite unrealistic for prediction in new untried situations."

Because perfect mixing is not obtained in the experimental unit, it may be possible to predict the performance of a flocculation unit by determining the amount of mixing that is to be expected as measured by the root-mean-square velocity gradient. This velocity gradient is a function of the energy input and, therefore, should be a measure of the degree of mixing since greater energy input should induce greater mixing.

EXPERIMENTAL RESULTS

In the following sections, the three physical arrangements of the experimental flocculation unit shall be defined as follows: a "single tank" shall be the case in which only compartment three of the unit was used; "three tanks in series" shall be the case in which the complete unit was used and separated into three sections by baffles; and a "large compartment" shall be the case in which the complete unit was used but the baffles removed.

The dimensionless response curves for Runs 14, 19, 24, and 25, representing three tanks in series are superimposed in Figure 10. Figure 11 is an expansion of the peak regions of this plot. Although differences among the peaks and ascending and descending portions of the curves exist, there is no general trend concerning variations of energy input and mixing levels. Experimental error could account for the variations that do exist. Similar results were obtained when the response curves for the runs for a single tank and runs for a large compartment were plotted in Figures 12 and 13, respectively. The latter plot has two curves only, and so generalizations as to variations cannot be made.

In the majority of the runs, however, the curves are higher than the corresponding theoretical curves, indicating that

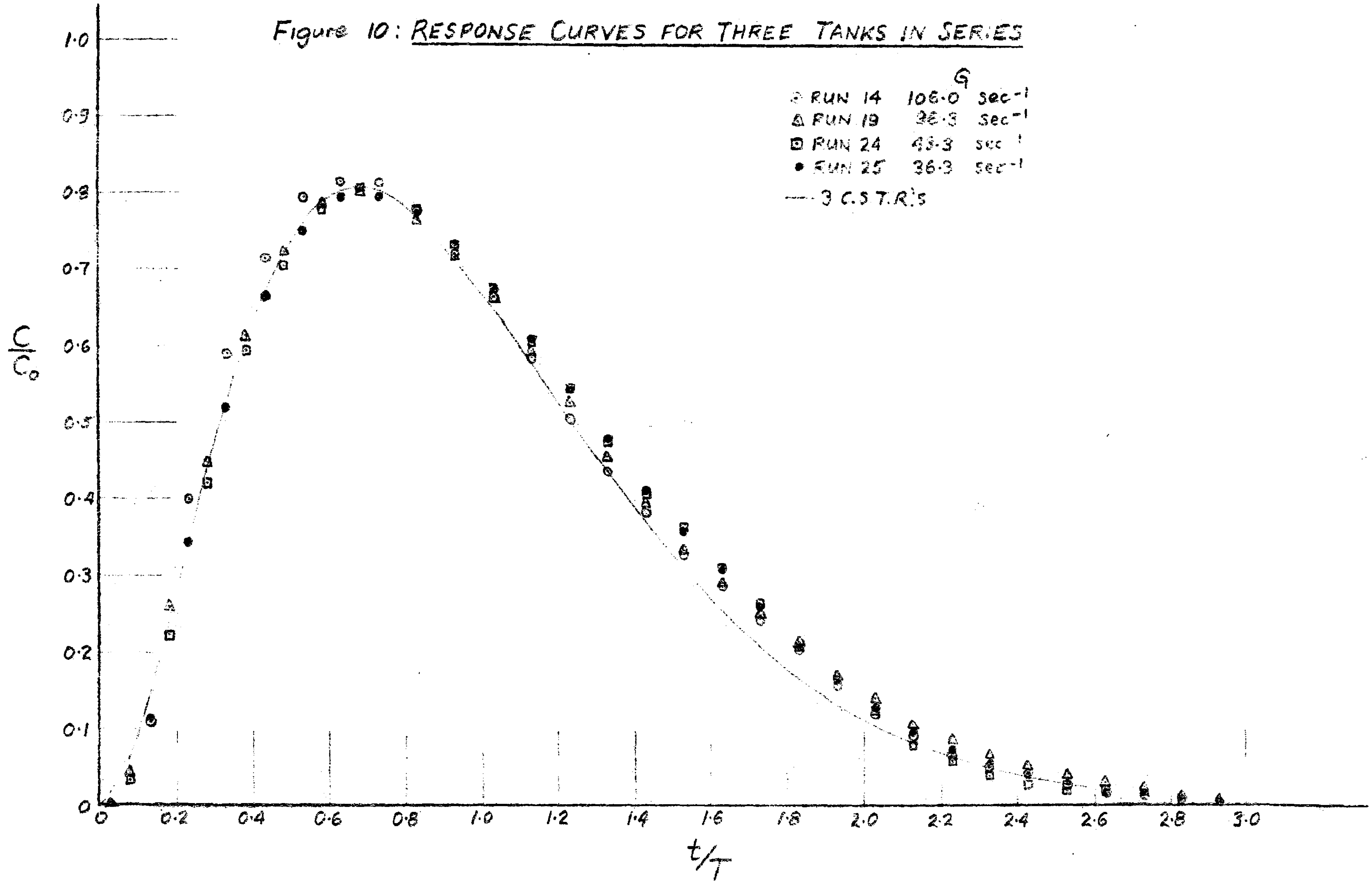


Figure 10: RESPONSE CURVES FOR THREE TANKS IN SERIES

Figure 11: PEAK OF RESPONSE CURVES (3 TANKS IN SERIES)

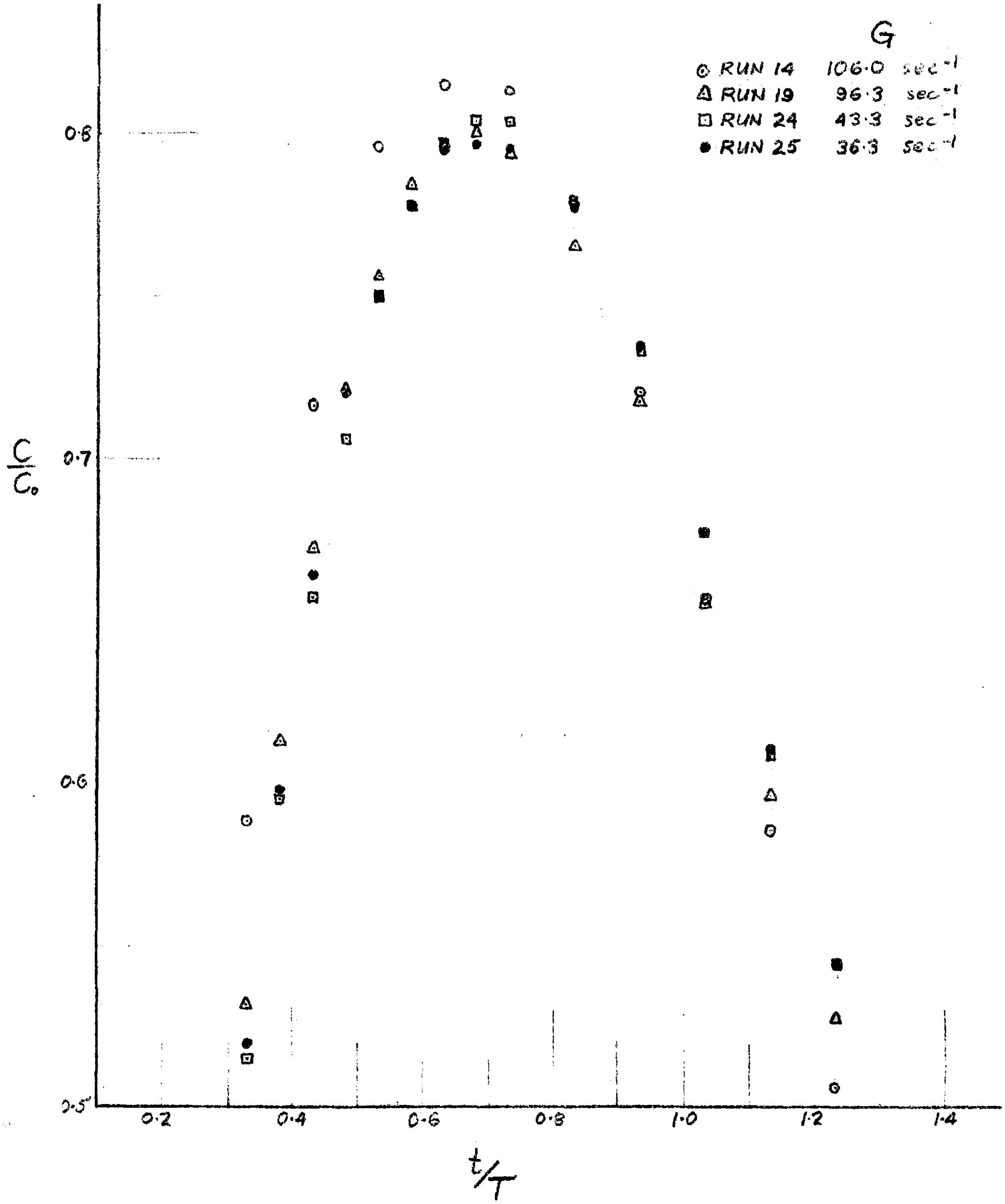


Figure 12: RESPONSE CURVES FOR ONE TANK

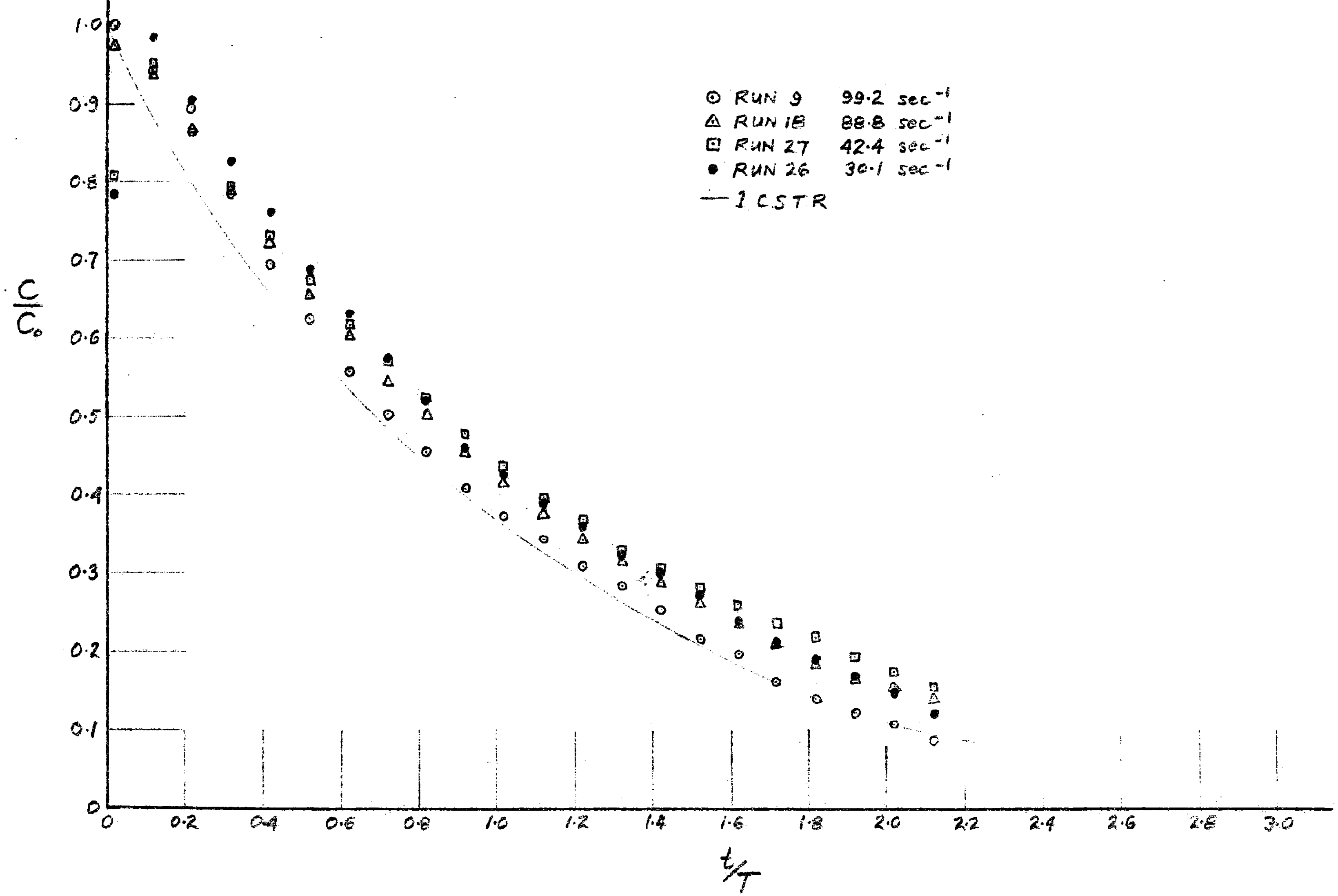
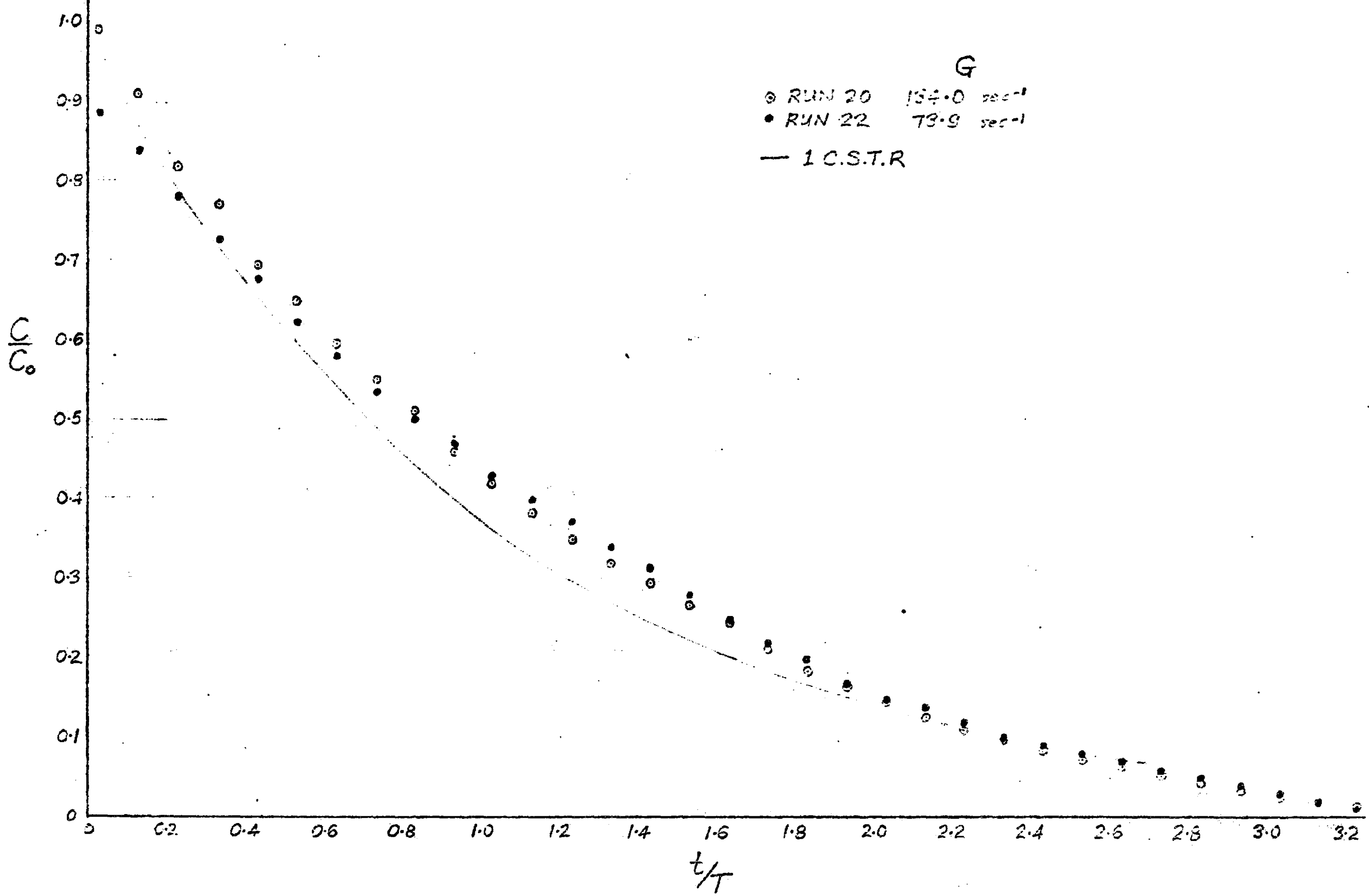


Figure 13: RESPONSE CURVES FOR ONE LARGE COMPARTMENT



complete mixing is not taking place in the experimental flocculation unit.

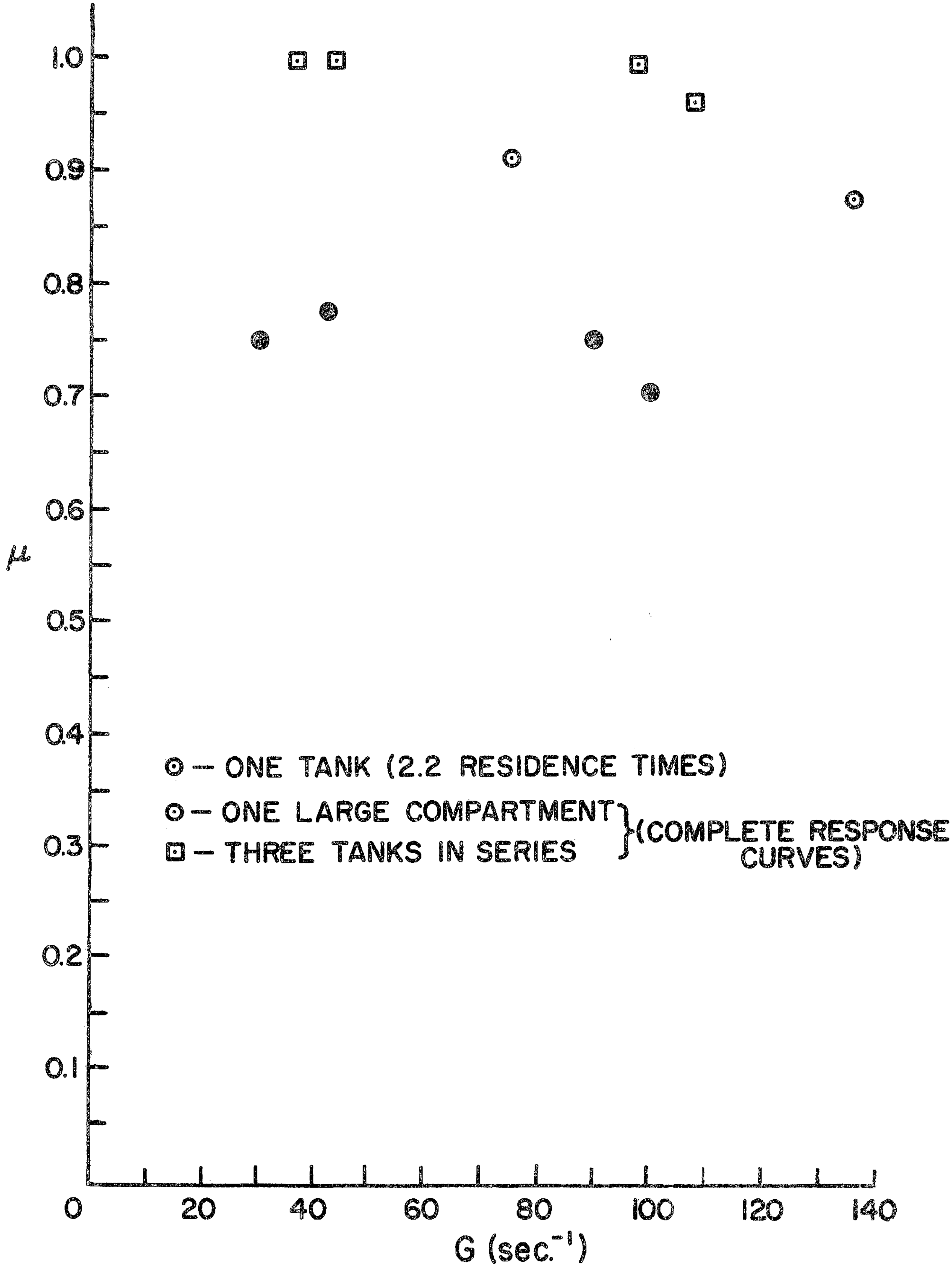
Because the lag time and peak time do not show any consistent changes with energy input at the range of energy input and residence time used in this study, the curves were analyzed for a variation in mean with energy input changes. Figure 14 shows the effect of energy input on the mean of the curves. G, the root-mean-square velocity gradient, is a function of the energy input and is used as the abscissa.

For each of the three physical configurations of the flocculation unit, there does not appear to be any major variation of mean over the range of energy input used in this study.

It should be pointed out here that the ranges of energy input used are consistent with the values used in actual practice. During his study of 20 flocculation basins in water-treatment plants in the United States, Camp (1955) showed that the maximum velocity gradients ranged from approximately 20 $sec.^{-1}$ to 74 $sec.^{-1}$ and the values of the product GT at plant capacity ranged from approximately 23,000 to 210,000. The values of G used in this study ranged from 30.1 $sec.^{-1}$ to 108.7 $sec.^{-1}$ and the values of GT ranged from 54,180 to 195,660.

Figure 14 does illustrate, however, that greater mixing takes place in a baffled unit than in a similar sized unit with the baffles removed. The means for the former case approach a value of unity, the theoretical value at which complete mixing occurs, while the means for the latter case have a value of approximately 0.9. Because

FIGURE 14: EFFECT OF ENERGY IMPUT ON THE MEAN.



the means are less than unity, deadwater regions exist, and the greater is the deviation from unity, the greater is the amount of deadwater. This fact is illustrated from the following equation taken from Levenspiel (1962a):

$$\frac{V_d}{V} = 1 - \frac{u_a}{u} \cdot \mu \qquad (22)$$

where V is the total volume of the vessel,

V_d is the "dead" volume of the vessel,

μ is the mean of the dimensionless response curve,

u is the average flow velocity through the vessel,

u_a is the flow velocity into and out of the "dead" portion of the vessel, and

$\frac{u_a}{u}$ is the area under the dimensionless response curve.

Deadwater accounts for that portion of the fluid within the vessel which is almost stagnant or completely stagnant. Only in the theoretical case of complete mixing are deadwater regions eliminated. Thus, baffling of a flocculation unit decreases the amount of deadwater and promotes better mixing.

Unfortunately, the points in Figure 14 representing the case of one tank cannot be compared with the other points on the same graph because the experimental curves for the one tank runs were truncated after 2.2 residence times. An attempt was made to extrapolate the points on these curves to obtain the complete curves so that the corresponding means could be calculated. This was impractical, however, because the complete descending portion of the response curves did not form a straight line when plotted on semi-logarithmic graph paper. The curves could have been approximated, but the error in the final results would have been too large to make a justifiable comparison. This non-linearity is illustrated in Figure 15.

The purpose of Figures 15, 16, and 17 is to illustrate graphically the existence of deadwater regions. If no such regions existed in a stirred vessel, the slope of the descending portion of the dimensionless response curves would equal minus unity, and equation (16), the basic equation for a first order reaction, would hold. This equation actually has the form:

$$\frac{C}{C_o} = e^{-mt/T} \qquad (23)$$

where m is the slope of the descending portion of the dimensionless response curve when plotted on semi-logarithmic graph paper. For the theoretical case of a C.S.T.R., m equals unity, but when deadwater regions exist, m should have a value of less than unity.

For a single tank, the slope of the experimental curves is approximately -0.93, while for a large compartment, the slope is approximately -0.81. Each of these values is greater than minus unity, or, expressed in a different way, the value of m for each case is smaller than unity. These values indicate that while both physical

Figure 15: DESCENDING PORTION OF RESPONSE CURVES

I. SINGLE TANKS

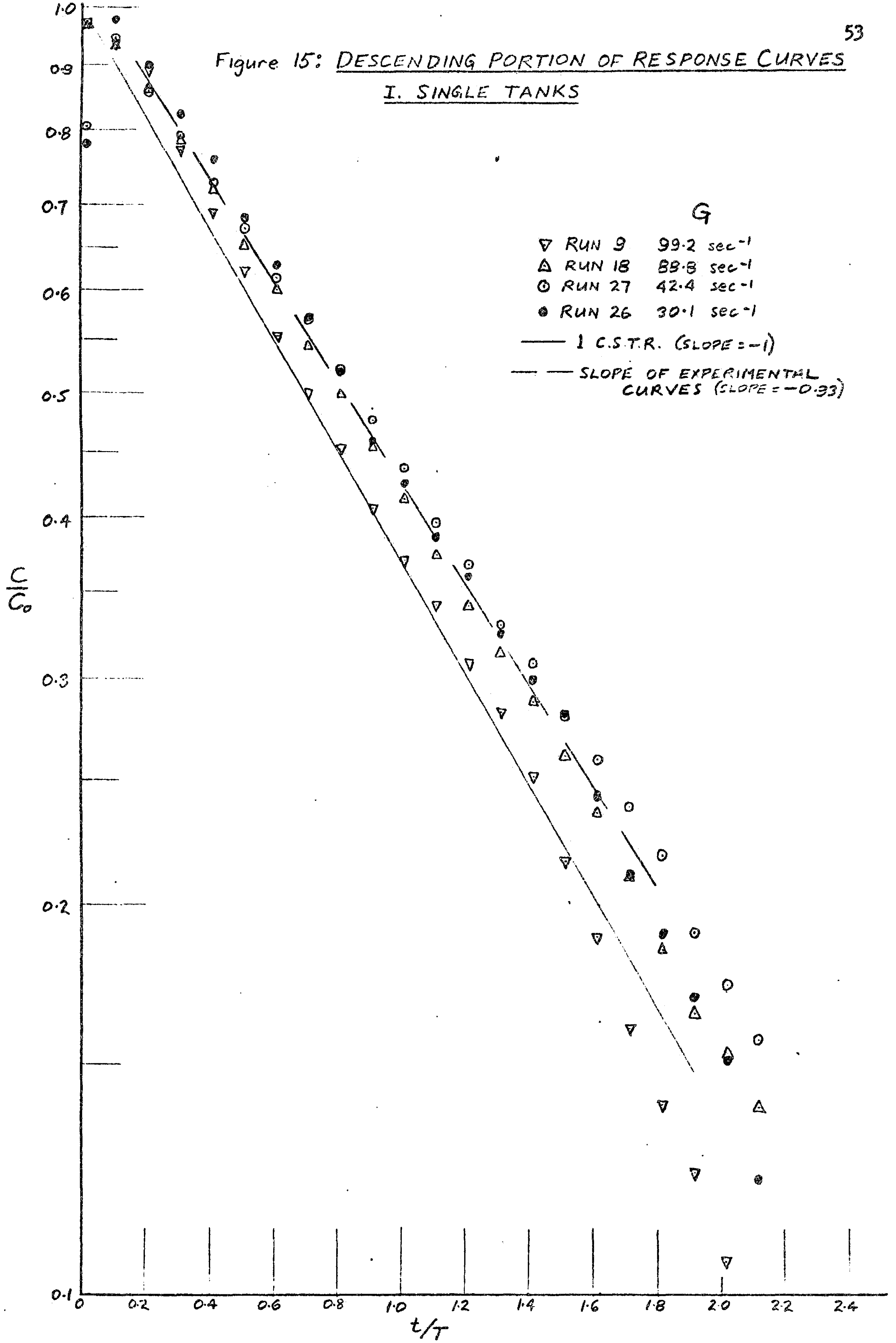


Figure 16: DESCENDING PORTION OF RESPONSE CURVES

II. LARGE COMPARTMENTS

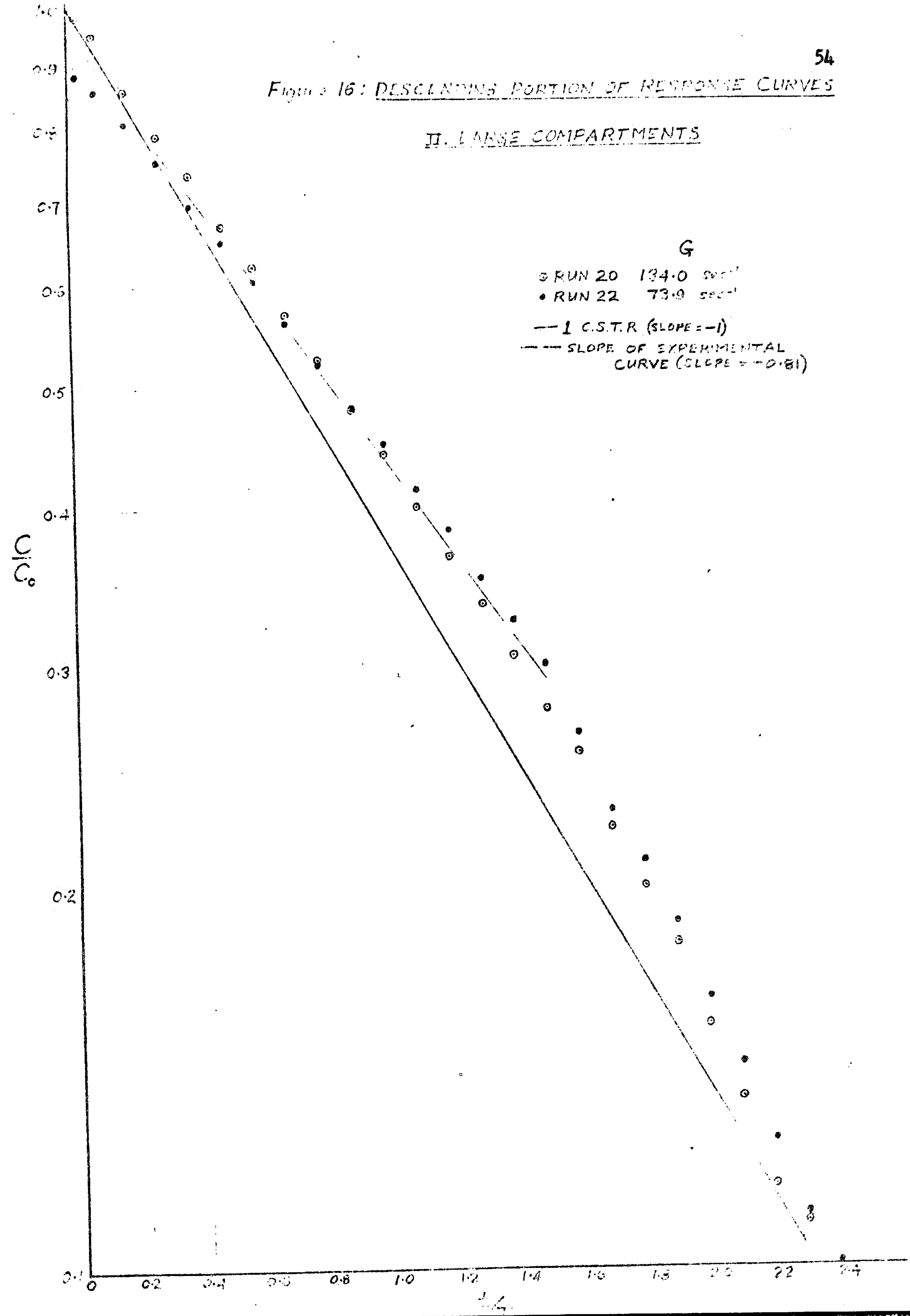
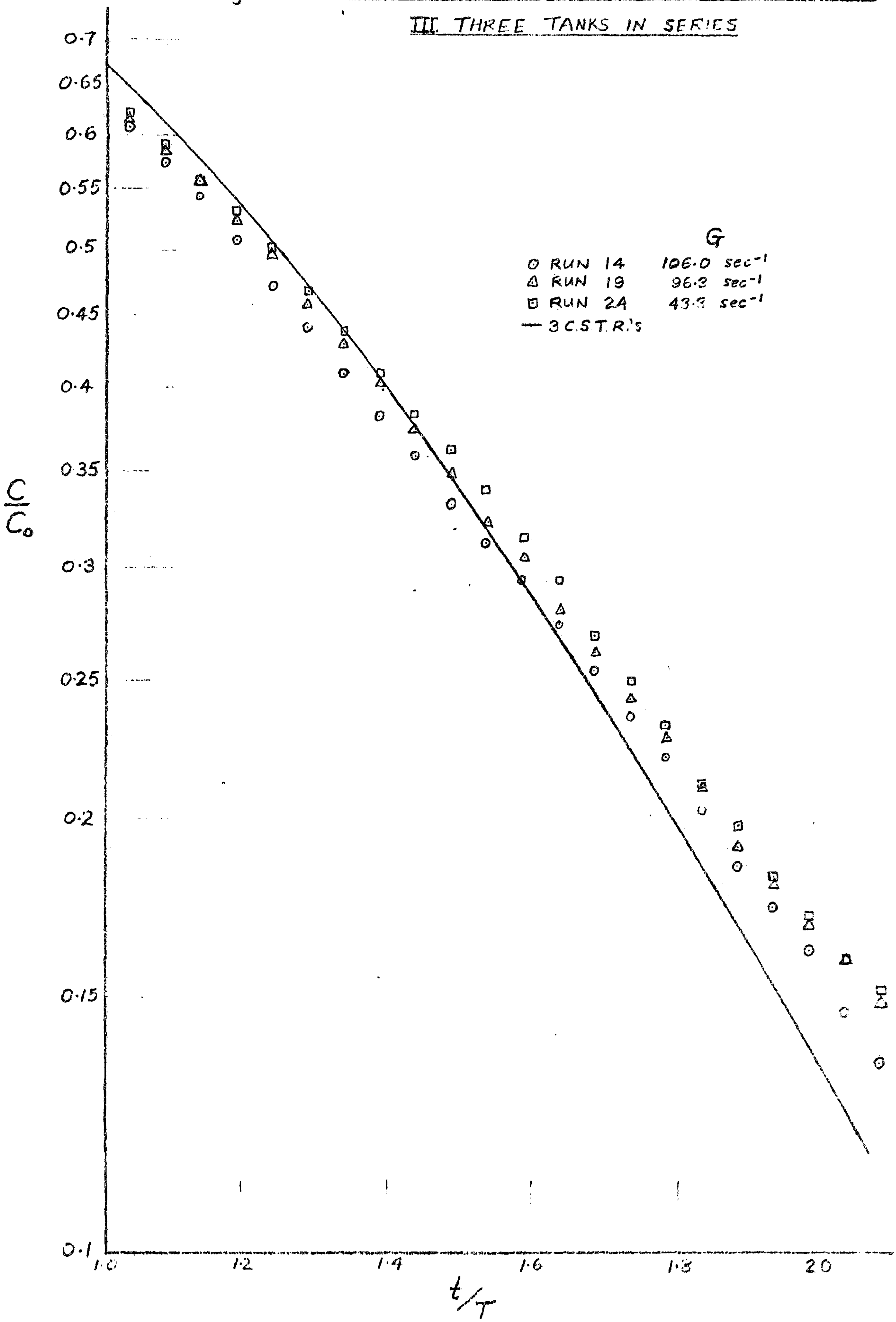


Figure 17 : DESCENDING PORTION OF RESPONSE CURVES

III. THREE TANKS IN SERIES



arrangements contain deadwater regions, greater mixing is achieved in the single tank than in the large compartment. In other words, with the same energy input per unit volume, better mixing is obtained in the one small vessel using one mixer than in the vessel three times as long using three similar mixers.

Figure 17 illustrates that complete mixing does not occur when three tanks in series are used. Although the descending portion of the theoretical curve is not a straight line when plotted on semi-logarithmic graph paper, it is evident that for any value of t/T, the slope of the experimental curves at that point is less in magnitude than the slope of the theoretical curve.

In summary, it can be concluded that complete mixing does not occur in the experimental flocculation unit over the range of energy input used in this study, nor does a variation in energy input cause a corresponding change in mixing levels in any of the three physical configurations of the experimental flocculation unit.

The velocity gradients used in the previous graphs were calculated from equations (19) and (2). Camp (1955) presented procedures in the practical application of the theory of the physical process of floc formation to the design of flocculation basins. He derived the following expression for the value of the dissipation function, W, for all the paddles in a tank:

$$W = \frac{239\ C_D\ (1-k_s)^3\ S^3}{V} \cdot \sum Ar^3 \qquad (24)$$

where W is the dissipation function or the work of shear in a fluid per unit volume per unit time,

C_D is the drag coefficient,

k_s is the ratio of the rotating velocity of the fluid to the velocity of the blades,

S is the speed of the shaft (revolutions per second),

V is the volume of liquid in the tank or series of tanks,

A is the cross-sectional area of the paddles,

r is the distance between the paddle blade and the center of the shaft, and

$\sum Ar^3$ is the sum of the values of Ar^3 for all rotors.

As before, G is defined by equation (2).

$$G = \sqrt{\frac{W}{\mu}} \tag{2}$$

The theoretical values of G were calculated for the experimental runs using equations (24) and (2) assuming no motion of the fluid induced by the paddles ($k_s = 0$). For these calculations, the baffled chambers were considered as three tanks in series and G was calculated for each tank. The values of C_D were taken from Binder (1962). The theoretical values of G are listed in Table 2 together with the measured, or observed, values of G and their ratios.

The two corresponding values of G for each tank are plotted against each other in Figure 18. In effect, a curve drawn through

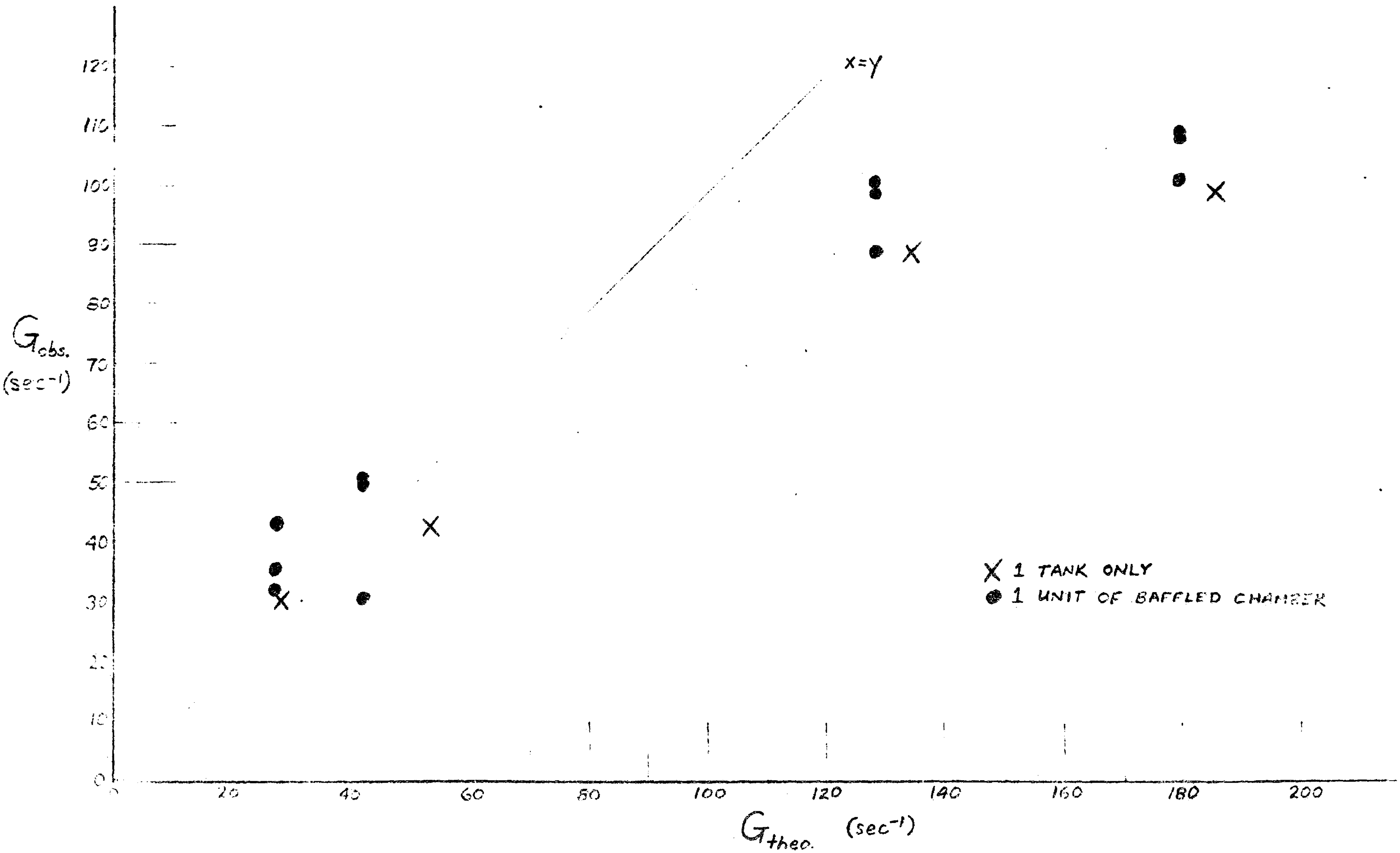
Table 2: COMPARISON OF MEASURED AND THEORETICAL G

	Run	Measured G ($G_{obs.}$) ($sec.^{-1}$)	Theoretical G (G_{theo}) ($sec.^{-1}$)	$\frac{G_{obs.}}{G_{theo.}}$
1 tank:	9	99.2	185.6	0.54
	18	88.8	133.8	0.66
	26	30.1	27.9	1.08
	27	42.4	53.0	0.80
3 tanks in series:	14	108.0	175.9	0.61
		108.7		0.62
		101.4		0.58
	19	98.7	128.3	0.77
		101.2		0.79
		89.1		0.70
	24	49.9	41.6	1.20
		49.6		1.19
		30.3		0.73
	25	35.6	27.3	1.30
		42.8		1.57
		31.4		1.15

Notes: 1. $G_{obs.}$ calculated from equations (20) and (2).
2. $G_{theo.}$ calculated from equations (24) and (2).

Figure 18: COMPARISON OF MEASURED & THEORETICAL ENERGY INPUTS

I: SINGLE TANK AND BAFFLED CHAMBER



the points in Figure 18 is a measure of k_s. Because k_s was assumed to be zero for $G_{theo.}$, the two values of G may be equated in the following manner:

$$G_{obs.} = (1-k_s)^{3/2} \cdot G_{theo.} \tag{25}$$

In the lower regions of energy input, $G_{theo.}$ equals $G_{obs.}$ and so k_s is approximately equal to zero. In the higher regions, $G_{theo.}$ becomes much greater than $G_{obs.}$ This implies that k_s is becoming much larger and should be equal to one if the curve extended to infinity. This increase in k_s is expected because more blades will sweep more water around with the paddles until, in the extreme case, a solid paddle will sweep all the water making k_s equal to one. As a result, the energy is not imparted to the water in such a manner as to give the best performance of mixing.

Using equation (25), k_s was calculated for several points along a curve through the points of Figure 18. These values were plotted against the corresponding values of $G_{obs.}$ in Figure 19 yielding the following relationship:

$$k_s = 0.00000025 \, G_{obs.}^{\;2.67} \tag{26}$$

Equation (26) can now be used to determine the value of k_s for each of the experimental runs involving a single tank or for any of the tanks of the baffled chamber.

Figure 19 : EVALUATION OF K_s FROM MEASURED ENERGY INPUT

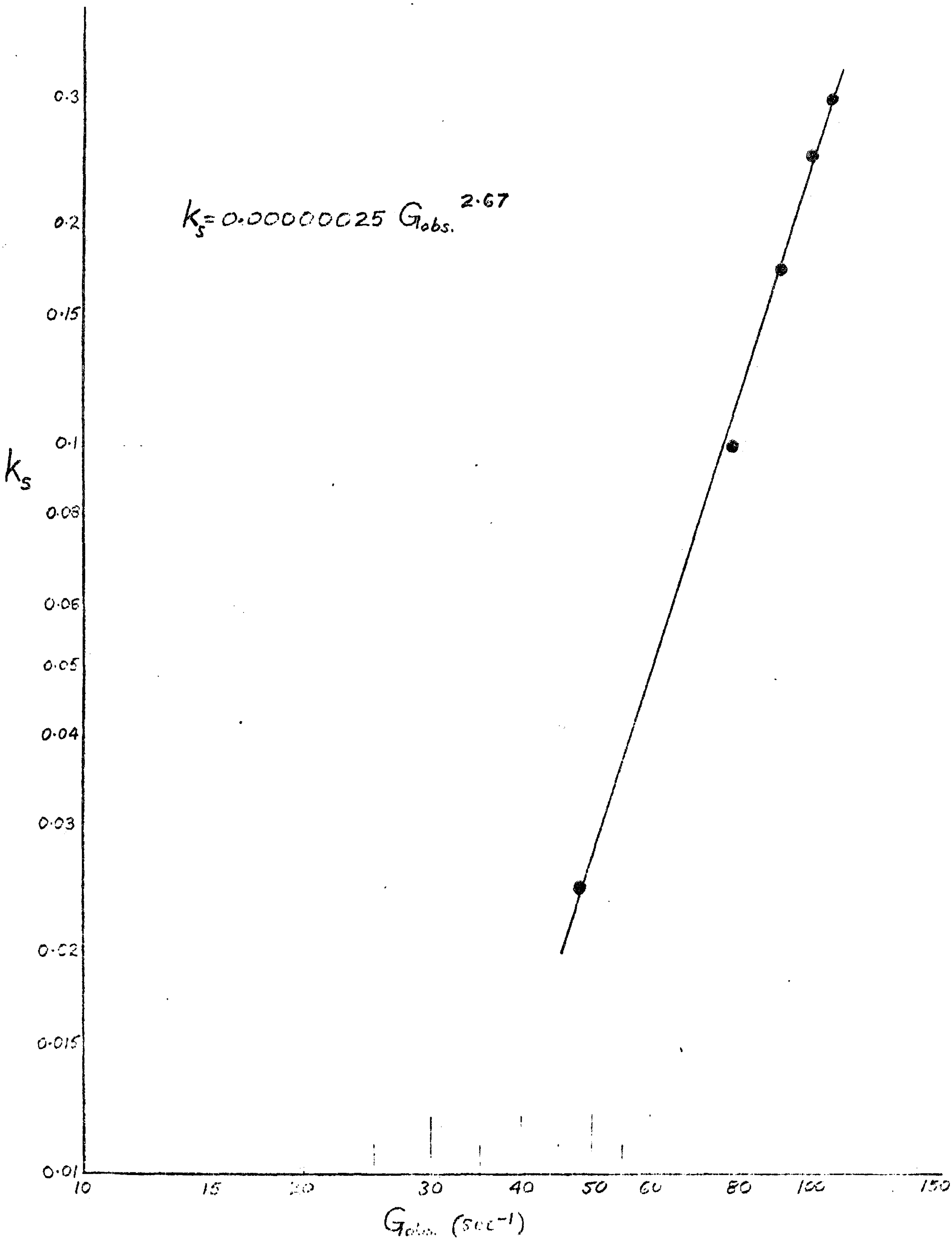
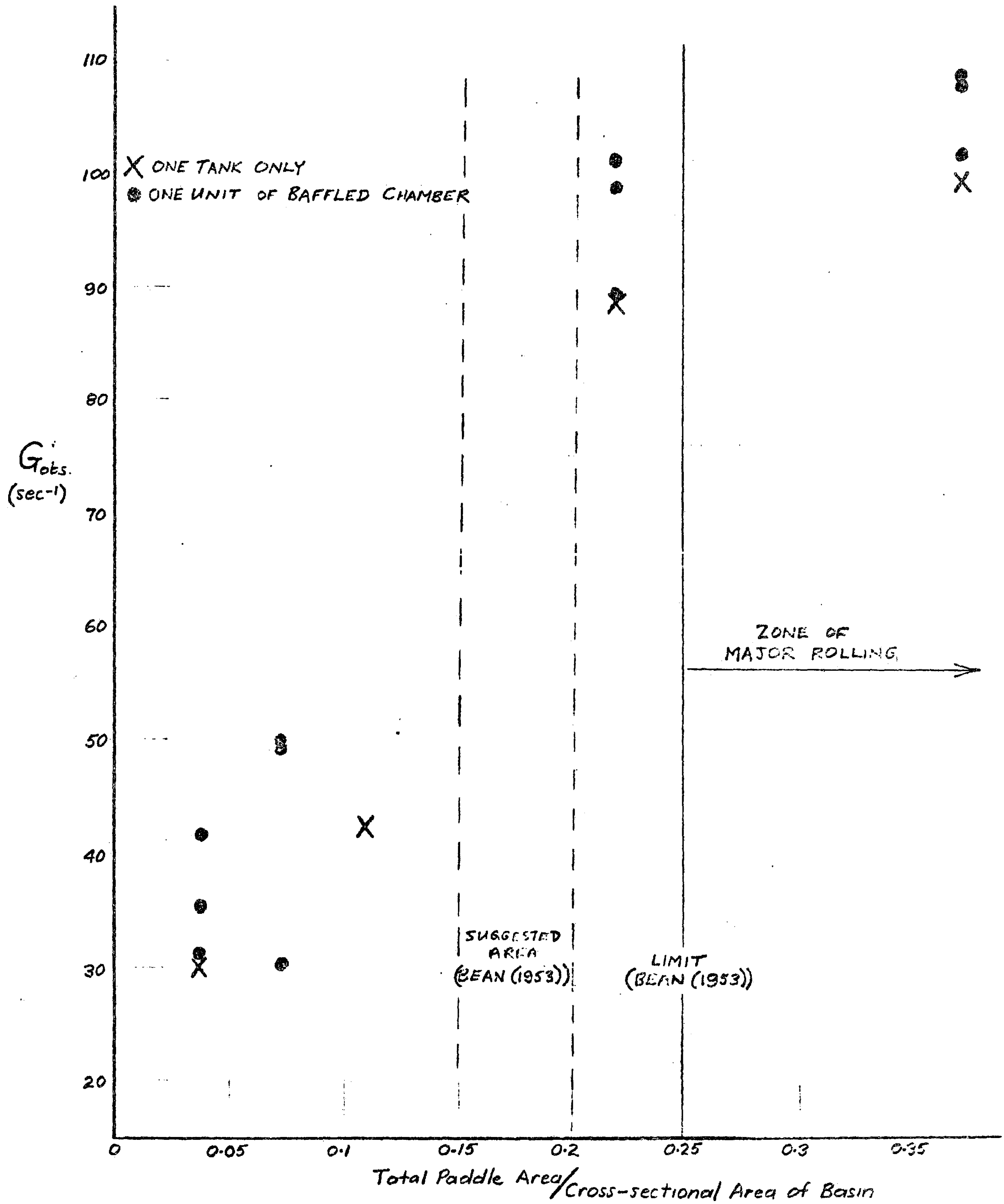


Figure 20 shows the effect of paddle area on velocity gradient. Bean (1953) stated that paddle spacing should be sufficiently wide that water in front of the paddles may readily pass around to the rear of the moving paddles. If the spacing is too small for the paddle widths, then roll must result. He suggested that the area of the paddles may generally be in the vicinity of 15 to 20 per cent of the cross-sectional area of the tank. 25 per cent will produce major rolling or rotation. Figure 20 illustrates that G is almost a linear function of the paddle area until the paddle area is 25 per cent of the cross-sectional area of the experimental basin. Increasing the paddle area beyond this value does not increase the velocity gradient to any great extent. In other words, rolling is occurring above 25 per cent.

Figure 21 shows the effect of the paddle area on the ratios of the velocity gradients. In the higher regions of paddle area, the ratio decreases as the paddle area increases. This is to be expected because of the relationships shown in Figure 18 and 20. As the paddle area increases, k_s increases and the ratio of $G_{obs.}/G_{theo.}$ decreases. Equation (25) indicates that the maximum value for this ratio should be unity, occurring when k_s is at a minimum of zero corresponding to a paddle area of zero. The maximum ratio, however, has a value of 1.57. The majority of the ratios obtained with the small paddle area have a value greater than unity. This might indicate that equation (24) is not valid in regions of low energy input.

Figure 20: EFFECT OF PADDLE AREA ON VELOCITY GRADIENT



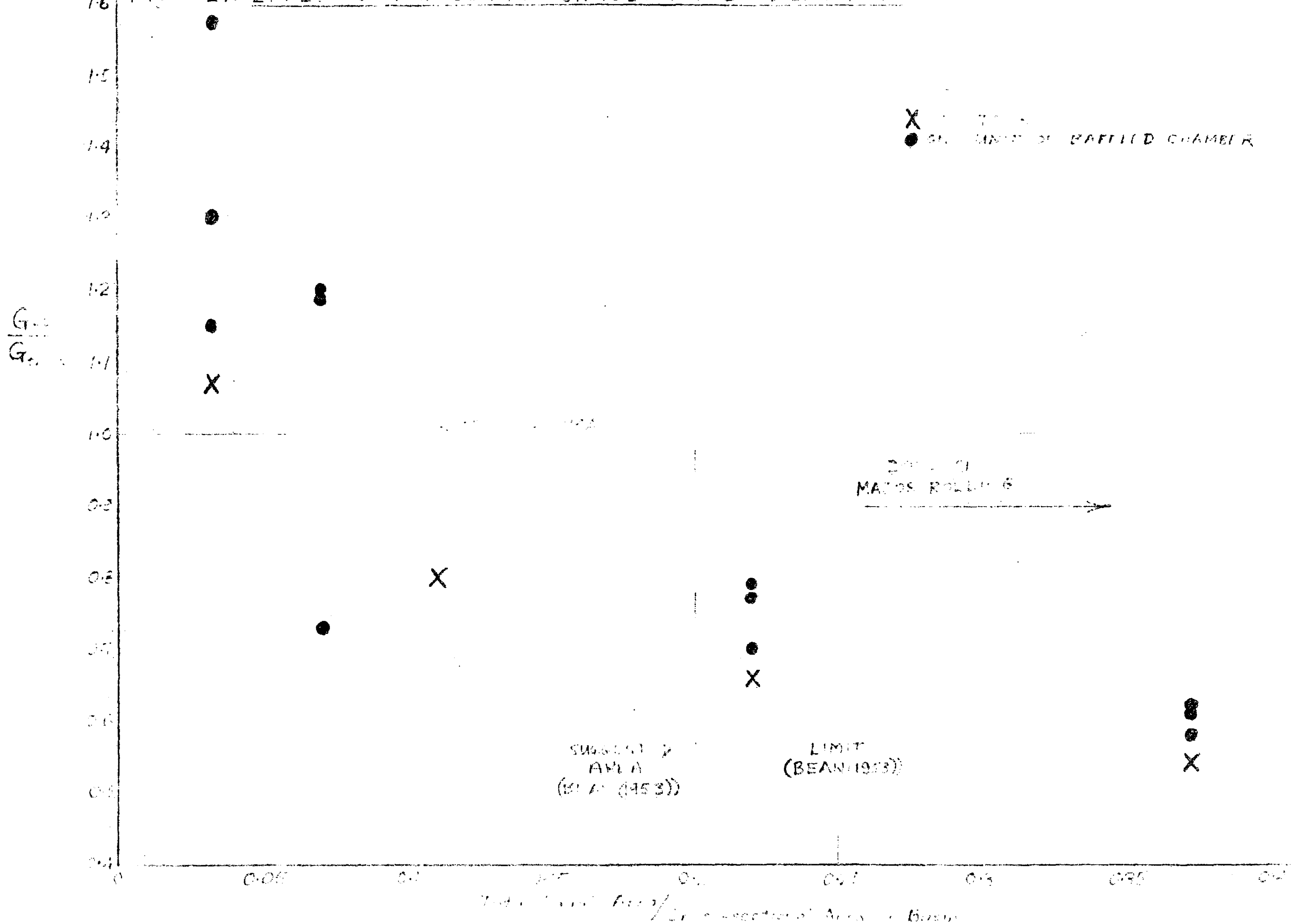


Figure 21: EFFECT OF PADDLE AREA ON VELOCITY GRADIENT RATIO

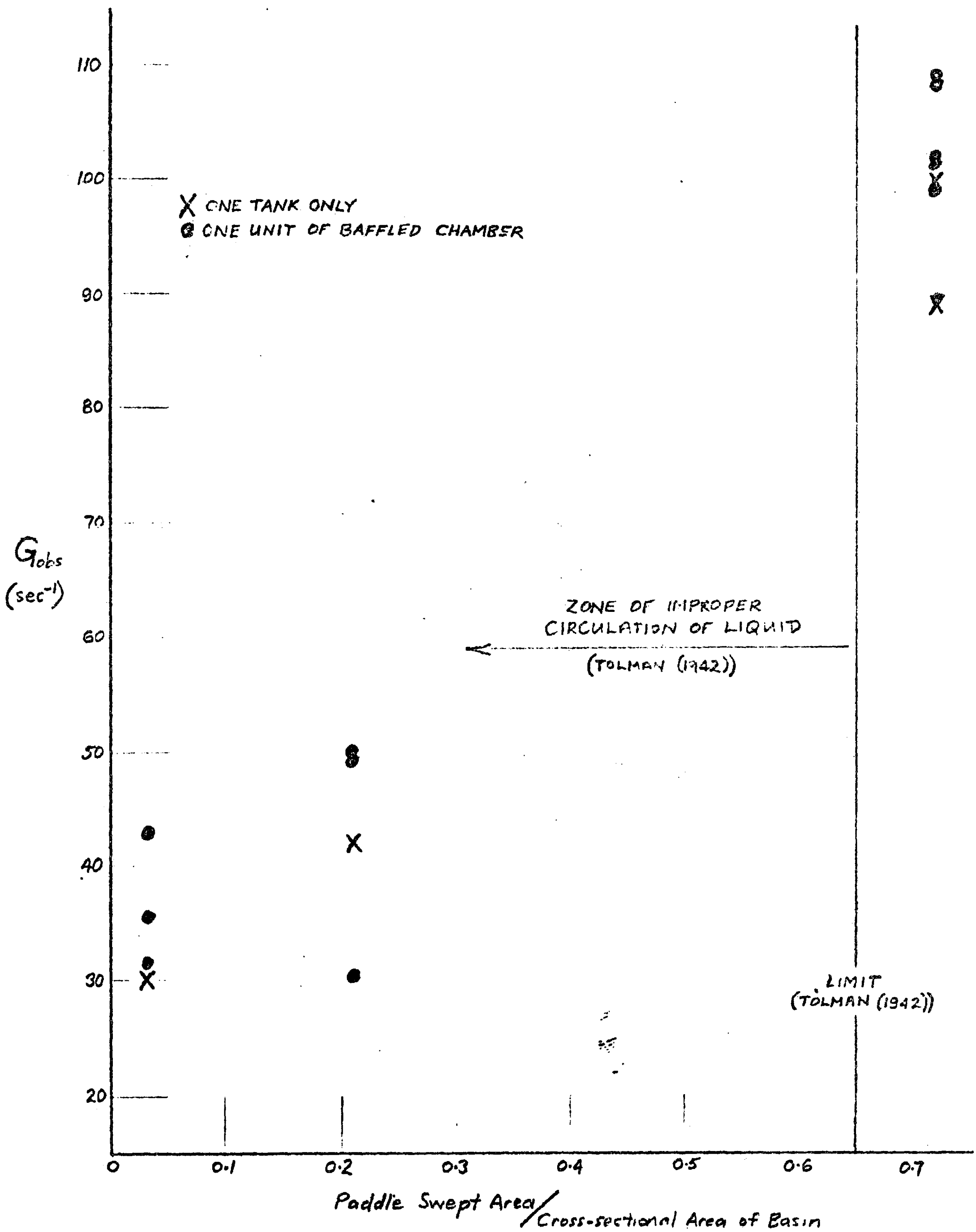
One of the important considerations in paddle wheel design and application is the clearance of the periphery of the wheel between walls, bottom, and liquid surface. With a great mass of liquid at the periphery of the wheel, a certain mixing speed is essential to cause circulation of the liquid from the center to the ouside of the wheel where it displaces the liquid at that periphery. After a series of tests, Tolman (1942) suggested that the areas swept by the paddle wheels should be not less than 65 per cent of the cross-sectional area of the basin.

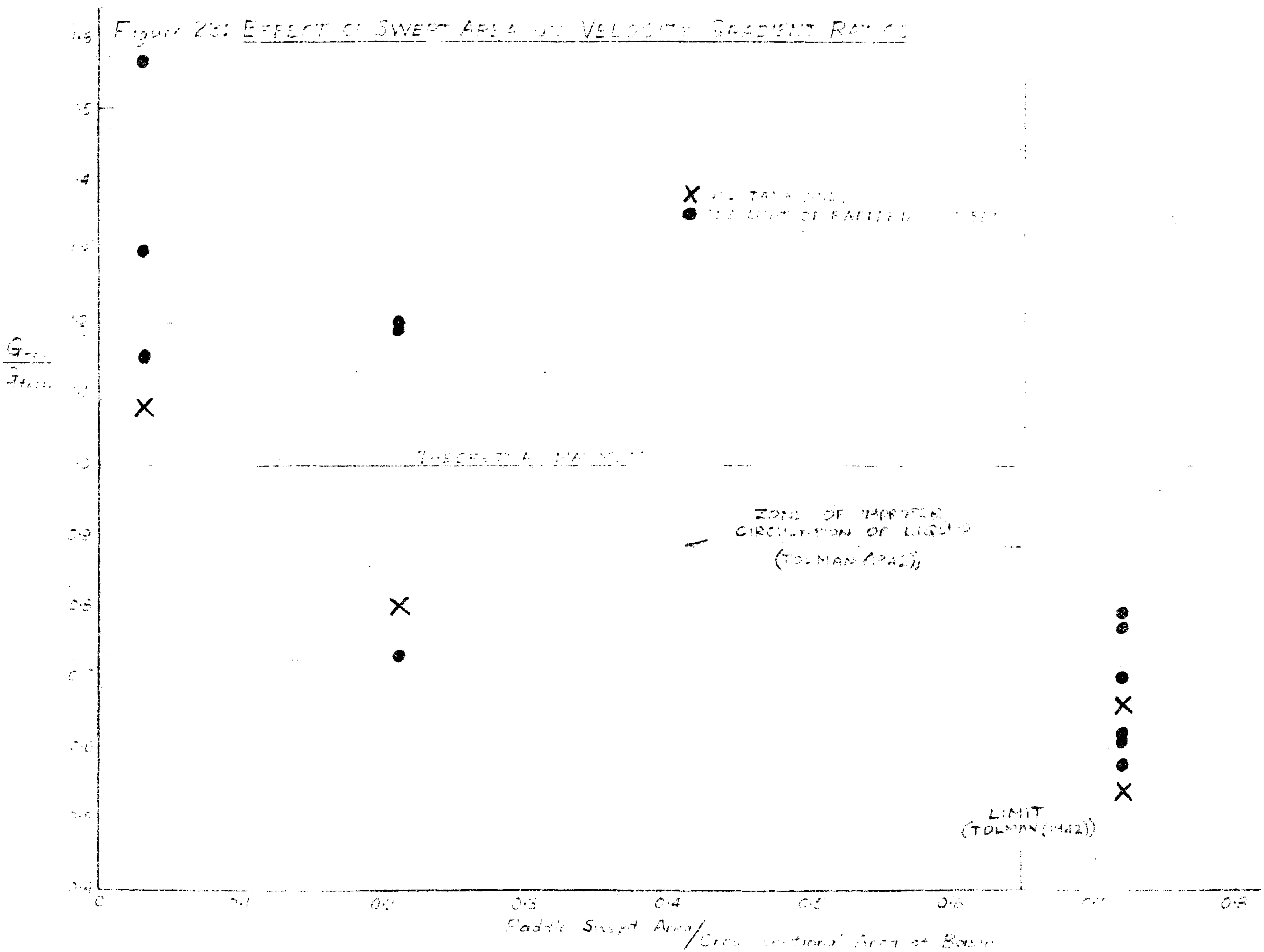
Figure 22 illustrates that G does not increase substantially with increased swept area until it is approximately 50 per cent of the cross-sectional area of the experimental basin. Above this value, G increases quickly as the swept area increases.

Figure 23 shows the effect of paddle swept area on the ratio of the velocity gradients. As in Figure 21, the ratio is above unity for low paddle swept area indicating again that euqation (24) may not be valid in regions of low energy input.

Figures 20 and 22 indicate that these design parameters of Bean and Tolman would appear to be based upon the effective conversion of energy input into mixing rather than the kinetics of flocculation. Whether these parameters are effective in flocculation of particulate systems would depend upon the kinetics of each system.

Figure 22: EFFECT OF SWEPT AREA ON VELOCITY GRADIENT




Effect of Swept Area on Velocity Gradient Ratio
LIMIT
(TOLMAN (1942))
0
0·4
0·8
Paddle Swept Area/Cross-sectional Area of Basin

CONCLUSIONS, DISCUSSION AND RECOMMENDATIONS

The studies presented in the previous chapters indicate that a mixed model should be used to adequately describe the mixing conditions in a flocculation unit. The Tanks-in-series model cannot be used because the hydraulic characteristics of a flocculation unit create short-circuiting and the existance of "deadwater regions". It may be possible to use the Dispersion model, but because of the high degree of dispersion in a flocculation unit, extremely accurate equipment would be required to measure the response curve, so that meaningful results could be obtained with the variance technique involved.

The following conclusions concerning energy and mixing in a flocculation unit can be made from the tests performed on a laboratory model with a maximum capacity of 6 liters:

1. Complete mixing does not occur in a small experimental flocculation unit over the range of energy input and the residence time used in this study.
2. Greater mixing is achieved in a baffled unit than in a similar sized unit without baffles. They reduce short-circuiting and the amount of deadwater regions.
3. With the same energy input per unit volume, greater mixing is achieved in a small unit than in a larger one using the

same residence time and no baffles in either.

4. A variation in energy input does not cause a corresponding change in mixing levels for either one small tank or three tanks in series.

5. The expression derived by Camp (1955) for determining the value of the dissipation function, or the work of shear in a fluid per unit volume per unit time, for all the paddles in a flocculation basin may not be valid for low energy inputs.

6. For higher values of energy input, the ratio of the rotating velocity of the fluid to the velocity of the blades (k_s) may be related to the measured root-mean-square velocity gradient (G) by the following relationship:

$$k_s = 0.00000025\ G^{2.67}$$

Baffling of a flow-through vessel to achieve better mixing is a fact mentioned often in mixing and flocculation literature. Argaman and Kaufman (1970) presented the most recent work describing experimental flocculation studies supporting this fact.

It is not surprising that complete mixing does not occur even at the maximum energy input used in this study. Only in the case of a perfect Continuous Stirred Tank Reactor should mixing be absolutely complete, eliminating short-circuiting and deadwater regions. Stators should reduce these deadwater regions and provide greater mixing.

The most surprising conclusion is the fact that a variation in energy input does not cause a corresponding variation in mixing levels.

The range of energy input used in this study was consistent with those used in actual practice. Argaman and Kaufman (1970) arrived at the following conclusion:

> "For a given residence time the performance (of a flocculation unit) increases almost linearly with the root-mean-square velocity gradient until a maximum value is reached beyond which any further increase results in a decrease in performance."

They are only the last of several authors to arrive at a similar conclusion. This seems to indicate that although a variation of energy input within the range used in actual practice causes a corresponding variation in flocculator performance, this change is due to shearing forces within the water causing aggregation and/or break-up of the flocs rather than to actual mixing levels reached in the vessel.

The expression derived by Camp (1955) for determining the value of the dissipation function in a flocculator is a very important one and is generally used as the basis of flocculator design. This report indicates that it may not be valid for the lower energy inputs used in this study. Because these low values are comparable to those used in actual practice, the validity of this equation should be checked for this lower range.

Horizontal-arm paddles are becoming more prominent in practice, and so experimental work should be done on these in an attempt to get a better understanding of the mixing - energy input relationship.

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APPENDIX "A"

Computer Programmes Used for Analyses

Table A1: NOMENCLATURE USED FOR COMPUTER PROGRAMMES

Programme Symbol	Meaning or Equivalent	Programme Symbol	Meaning or Equivalent
AREA	Area under dimensionless response curve	NORUN	Total number of experimental runs
C1(I)	C (experimental)	NRUN	Number of experimental run
C(I)	C (corrected)	Q	Flow rate
CO	C_o	SUMF	$\sum \frac{C}{C_o}$
DELT	Δt	T	t
DET	T	TEMP	Temperature
DCONC	Tracer concentration	T(I)	t corresponding to C
DVOL	Tracer volume	TH(I)	$\frac{t}{T}$
FACT	(n-1)!	THETA	$\frac{\Delta t}{T}$
FLAG	Time delay from flocculator effluent to recorder strip chart	UMTHF	$\sum \left(\frac{t}{T} \cdot \frac{C}{C_o} \right)$
FN1	$e^{-t/T}$	V	V
FN1(I)	$\frac{C}{C_o}$	XMEAN	Mean of dimensionless response curve
N	Number of data points per run	XN	Number of tanks-in-series

Table A2: <u>EXPERIMENTAL RESPONSE CURVE ANALYSIS</u>

```
C
      DIMENSION T(300),C(300),FN1(300),TH(300),C1(300)
      READ (2,15)  NORUN
      DO 1000 IJK =1,NORUN
      READ (2,15) NRUN
      WRITE (3,2)     NRUN
      READ (2,6) DCONC,DVOL,V,TEMP
      READ (2,5) Q,DELT
      DET=V/Q
      READ (2,15) N
      M=N-1
C
C FLUOROMETER OUTPUT CORRECTIONS
C
      READ (2,10) (C1(I), I=1,N)
C  DELAY TO FLUOROMETER
      FLAG=56.0*3.1416*(3.0/16.0)**2*16.387/(4.0*Q)
      DO 20 I=1,N
      C(I)=C1(I)
C  FLUOROMETER CALIBRATION CORRECTION
      IF (C(I)-16.0) 64,64,63
  63  C(I)=(C(I)+5.416)/85.905
      GO TO 21
  64  C(I)=0.016*C(I)**0.996
C  TEMPERATURE CORRECTION
  21  C(I)= C(I)/EXP(0.0265*(18.0-TEMP))
  20  CONTINUE
C
C DIMENSIONLESS RESPONSE CURVE
C
  71  C0=DCONC*DVOL/V
      DO 30 I=1,N
      FN1(I)=C(I)/C0
      T(I)=FLOAT (I)*DELT-FLAG
      TH(I)=T(I)/DET
  30  CONTINUE
       THETA=DELT/DET
C
C AREA UNDER THE DIMENSIONLESS CURVE
C
  80  SUMF=0.0
      DO 50 I=2,M
      SUMF=SUMF+FN1(I)
  50  CONTINUE
       AREA=THETA/2.0*(FN1(1)+2.0*SUMF+FN1(N))
      WRITE (3,27) AREA
   55 WRITE(3,1)
      WRITE (3,2)
```

Table A2: (Continued)

```
      UMTHF=0.0
      DO 90 I=1,N
      WRITE (3,25) FN1(I),TH(I)
       UMTHF=UMTHF+(TH(I)*FN1(I))
   90 CONTINUE
      XMEAN=UMTHF/SUMF
      WRITE (3,34) XMEAN
C
    1 FORMAT (1X,29HDIMENSIONLESS RESPONSE CURVE)
    2 FORMAT  (1H1,1X, 3HRUN,I3//)
    3 FORMAT (/11X,4HC/C0,10X,5HT/DET/)
    5 FORMAT (2F10.4)
    6 FORMAT (4F10.4)
   10 FORMAT (5F15.5)
   15 FORMAT (I4)
   25 FORMAT (1X,2F15.5)
   27 FORMAT (1X,22HAREA UNDER THE CURVE =,F8.5//)
   34 FORMAT (//1X,6HMEAN =,F8.5)
 1000 CONTINUE
      CALL EXIT
      END
```

Table A3: <u>RESPONSE CURVES FOR n TANKS-IN-SERIES</u>

```
C
      WRITE (6,1)
      WRITE (6,2)
C
      DET=30.0
C
      FACT=1.0
      XN=0.0
   30 XN=XN+1.0
      IF (XN.LT.2.2) GO TO 60
      FACT=FACT*(XN-1.0)
   60 WRITE (6,3) XN
      WRITE (6,4)
      T=0.0
   10 T=T+0.5
      GO TO 7
   20 T=T+2.0
    7 THETA=T/DET
      IF (XN.LT.1.2) GO TO 40
      FN1=(XN**XN/FACT)*(THETA**(XN-1.0))*EXP(-XN*THETA)
      GO TO 50
   40 FN1=EXP(-THETA)
   50 WRITE (6,6) FN1,THETA
      IF (THETA.LT.0.51) GO TO 10
      IF (THETA.LT.2.51) GO TO 20
      IF (XN.LT.8.2) GO TO 30
      STOP
C
    1 FORMAT (3X,31HSOLUTION FOR N TANKS IN SERIES.//)
    2 FORMAT (3X,11HDET=30 MIN./)
    3 FORMAT (//3X,2HN=,F3.1)
    4 FORMAT (//10X,4HC/C0,11X,5HT/DET/)
    6 FORMAT (1X,2F15.5)
C
      END
```

APPENDIX "B"

Fluorometer Calibration Curves

Figure B1: FLUOROMETER CALIBRATION CURVE FOR HIGH VALUES

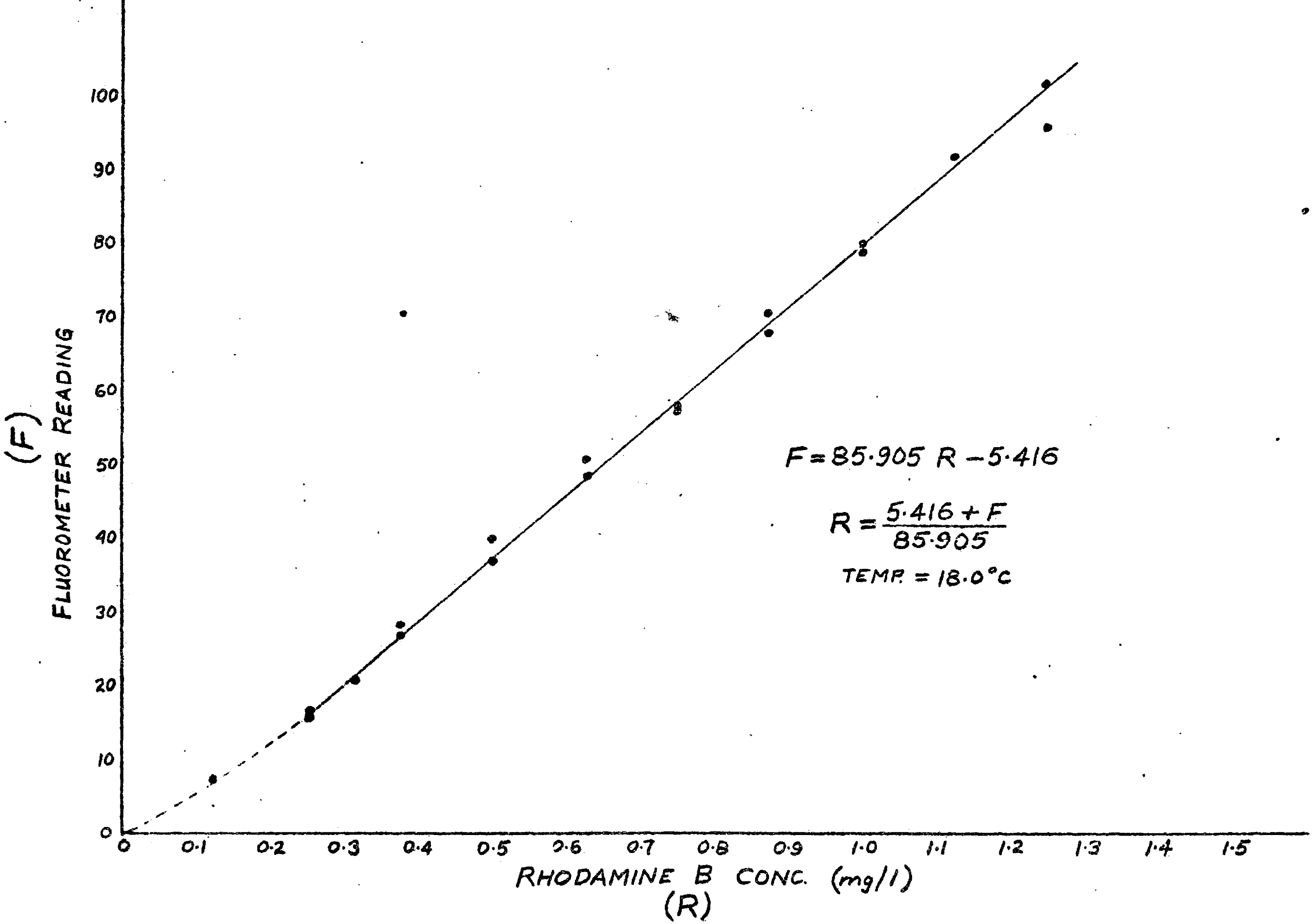
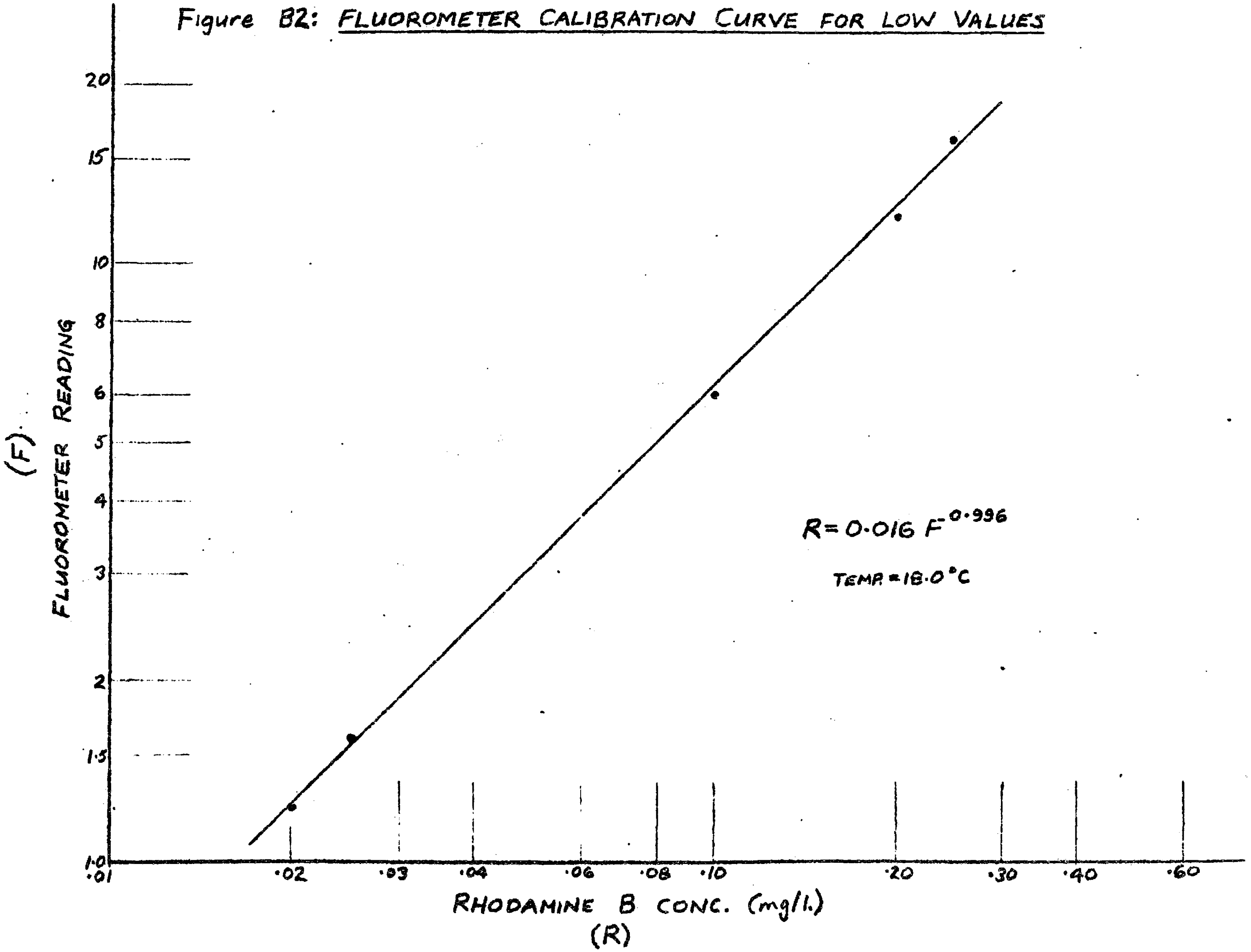


Figure B2: FLUOROMETER CALIBRATION CURVE FOR LOW VALUES



APPENDIX "C"

Nomenclature

Table C1: NOMENCLATURE

A = cross-sectional area of paddle blades

C = concentration at any time or point

C_o = initial concentration at inlet of vessel

C_D = drag coefficient

C_θ = fluorescence at any temperature, θ

C_s = fluorescence at a standard temperature, θ_s

D = longitudinal or axial dispersion coefficient

D = experimentally determined rate variable

d_1, d_2 = diameters of particles of size 1 and size 2, respectively

$\frac{du}{dz}$ = velocity gradient

f = friction factor

G = root-mean-square velocity gradient

H_{ij} = collision frequency between i-fold and j-fold particles

K = reaction constant

k = rate constant

k_s = ratio of the speed of the shaft and the velocity of the water surrounding the paddle

L = length of vessel

m = number of reactors in series

N_1^O = mass concentration of primary particles in flocculator influent

N_1^1 = mass concentration of primary particles in flocculator effluent

N_1^m = mass concentration of primary particles in effluent of m'th compartment of flocculator

Table C1: (Continued)

n = number of tanks in series

n_i = concentration of i-fold particles

n_j = concentration of j-fold particles

n_1 = number concentration of particles of size d_1

n_2 = number concentration of particles of size d_2

n_1^0 = number concentration of primary particles in flocculator influent

n_1^1 = number concentration of primary particles in flocculator effluent

R = hydraulic radius

r = distance between center of paddle blade and center of shaft

r_i = radius of an i-fold particle

r_j = radius of a j-fold particle

S = speed of mixers

T = residence time

T_p = torque of mixers

t_p = peak time of response curve

t = real time

u = average flow velocity through vessel

u_a = fluid in active flow through vessel

V = volume of vessel

V_d = "dead" volume of vessel

V_2 = volume of floc particles per unit volume of water

W = dissipation function or the work of shear in a fluid, per unit volume per unit time

Table C1: (Continued)

Greek Symbols

θ = temperature

θ_s = standard temperature

μ = mean of dimensionless response curve

μ = absolute viscosity of the fluid

ν = kinematic viscosity of the fluid

σ^2 = variance of dimensionless response curve

ϕ = fraction of primary particles adhering to floc particles