DISSOLVED - AIR
(PRESSURE)
FLOTATION

by

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DECLARATION BY CANDIDATE

I, John Bratby, hereby declare that this thesis is my own work and that it has not been submitted for a degree of another University.

Signed by candidate

April, 1973.

SYNOPSIS

It appears that, in the past, little attempt has been made to establish the basic parameters influencing the flotation process, or their inter-relationships, on a quantitative basis for the purpose of developing rational design procedures.

This thesis investigates those parameters influencing solids removal and thickening by dissolved-air (pressure) flotation and the criteria pertaining to the design of flotation tanks for solids removal and thickening.

A review of pertinent literature on flotation is presented. The historical development of flotation is given together with the theory of bubble-particle adhesion. Four common methods of flotation are compared and their inherent advantages and disadvantages are discussed. Dissolved-air (pressure) flotation, the method used in this investigation, is considered in greater detail by presenting established practical considerations for design.

A pressure-saturation unit was developed in which all the compressed air supplied is dissolved into the saturator feed by continously recycling the air through a sparger, i.e. there is no air wastage. The air circulating pump operates within the pressurized circuit so that it pumps only against the hydrostatic head of water in the saturator.

A vertical flow type flotation unit was developed which provides a counter current flow regime and a stable interface between the clear liquid and rising agglomerates. The accumulated agglomerates discharge freely and without mechanical assistance at the top of the unit.

The relationship between the mass of air precipitated at atmospheric pressure per unit volume of saturator

feed, a_p, and the (gauge) pressure in the saturator, P_g, was found to be not linear, as suggested by Henry's Law, but of the form

$$a_p = K_2(P_g)^{K_1}$$

where K_1 and K_2 are constants.

This deviation from Henry's Law is due to the rate of air precipitation which is slow at low pressure differences and fast at higher pressure differences. Operating the saturator at low (gauge) pressures is therefore less efficient than operation at high pressures.

Bubble-particle adhesion was found to be very pH dependent, more so than colloid destabilization. For this reason the standard jar test is not a suitable method for determining coagulant type, dosage and pH when metal coagulants are added for flotation. The optimum conditions of coagulant type, dosage and pH for destabilization may differ from those for bubble-particle adhesion.

Sulphate anions were found to be detrimental to the flotation process, as applied to the algal wastewater used in these investigations. Aluminium sulphate and ferric sulphate did not produce an effluent quality of high standard whereas the effluent when ferric chloride was used had a low turbidity.

Flocculation is not a necessary pre-requisite for flotation. With efficient mixing, a 14 sec reaction time after coagulant addition was all that was required prior to flotation.

The relationship between the air/solids ratio, a_s, and the limiting downflow rate, v_L, beyond which solids are drawn down with the effluent was found to be of the form

$$v_L = K_2(a_s)^{K_1}$$

where K_1 and K_2 are constants.

The float solids concentration, C_F, was found to be dependent on three parameters:

(a) Solids loading rate, Q_s;

(b) Air/solids ratio, a_s;

(c) Depth of float above water level, d_W.

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CHAPTER 1 - INTRODUCTION

Suspended solids, in the effluent from wastewater treatment plants, can result in many deleterious effects if discharged to rivers, for example:

(a) The solids settle on the bed of the river and form a silt which smothers plant and animal life and

(b) the solids may have an oxygen demand reducing the oxygen level in the river.

The unit processes which remove suspended solids from the wastewater may be given the collective name "solids-liquid seperation" processes. As an example, the activated sludge process has, as an intergral part of the process, a solids-liquid separation stage where the flocculated microorganisms are separated from the stabilised liquid fraction. The flocculated microorganisms are accumulated as a sludge and the stabilized liquid fraction is passed on for further treatment. Inefficient solids-liquid separation in this process produces a wastage of energy input to the plant since a considerable portion of the pollution load is contained in the solid fraction. Furthermore, inefficient solids-liquid separation at this stage may cause overloading in subsequent (polishing) solids-liquid separation processes, in, for example, sand filtration before discharge to the receiving stream.

Solids-liquid separation processes which have the objective of removing the "bulk" of the solid material from waters and wastewaters (rather than "polishing" a relatively solids-free effluent) utilise the density difference between the solid and liquid phases. Two processes making use of this method of separation are sedimentation and flotation. In sedimentation, particles settle to the bottom of the tank to accumulate as a sludge layer where thickening takes place by consolidation. Since wastewater particles are not appreciably greater in density than the liquid, the rate of settlement is usually slow and, consequently, the area requirements of sedimentation tanks are relatively large.

In flotation, separation of the solid particles from the liquid is brought about by the attachment of gas (usually air) bubbles to the particles. The bubble-particle agglomerates have a density lower than the liquid and rise to the surface where they accumulate as a float. The rate of rise of the bubble-particle agglomerates may be varied by adjusting the amount of air supplied to the particles. The rate of rise may be in the order of five times the rate of settlement during sedimentation. Consequently, the area requirements of flotation tanks are comparatively small.

In sedimentation, the inter-relationships of the process variables are fairly well established. In flotation, this is not so. The objectives of this research were, therefore, to identify the parameters which influence the flotation process and determine their inter-relationships.

Specifically, the objectives were to:

(1) develop an efficient dissolved-air (pressure) flotation system;

(2) determine the parameters influencing bubble-particle attachment; and

(3) identify the parameters and establish their inter-relationships with respect to (a) solids removal by flotation and (b) thickening of the floated solids.

In Chapters 2 to 5 (inclusive) a review of the pertinent literature on flotation is presented. Chapter 2 comprises a brief historical introduction to flotation. Chapter 3 deals with the theory of flotation based mainly on the work of Derjaguin. In Chapter 4, a comparison is made between the four most common methods of flotation and the advantages and disadvantages of each are briefly discussed. Chapter 5 presents certain practical considerations when using dissolved-air (pressure) flotation, which is the method used in these investigations.

Chapter 6 presents the methods, results and conclusions gained from the experimental investigations. The first section of this chapter deals with the flotation system developed as part of this investigation; the second with the efficiency of the saturation unit; the third with the effect of pH and anions on bubble-particle adhesion; the fourth with the effect of reaction time and degree of pre-flocculation on flotation; the fifth with the effect of the cross sectional area of the upper chamber on the limiting downflow rate; the sixth with the effect of the air/solids ratio on the limiting downflow rate; and the seventh with those parameters influencing the float solids concentration.

An eighth section presents an illustrative design example for a flotation tank, using the data obtained in these investigations. The ninth, and final section deals with forseeable future investigations into flotation.

CHAPTER 2 - HISTORICAL INTRODUCTION

(See reference 1).

Over 2000 years ago, the ancient Greeks, in order to separate desired minerals from crushed ores, (apparantly) used a flotation process whereby the crushed ore was fed on to a water surface. The mineral particles were retained at the surface by surface tension forces and the undesired (gangue) material sank to the bottom. This process became known as skin flotation.

In 1860, the William Haynes' process consisted of triturating dry-ground ore with an oily substance and then discharging the mass into water. The desired substances, combined with the oil, rose to the surface leaving the gangue material to sink to the bottom. This process became known as bulk-oil flotation.

In 1901 to 1902, Potter and Delprat developed a process of flotation where gas was used as the buoyant medium. One of the early applications of this process was to feed dry ore containing about 20 per cent of zinc into a hot solution of sulphuric acid. The acid, attacking the ore, liberated bubbles of gas which, attached to particles of zinc, buoyed them to the surface where they were skimmed off and recovered.

Elmore in 1904, suggested the use of electrolysis in water to produce gas bubbles but, at that time, this method was not applied commercially. This idea was later developed into a process known today as electrolytic flotation. Also in 1904, Elmore invented a process whereby air bubbles were generated by applying a vacuum to a saturated liquid whereby the air precipitated out of solution in the form of minute bubbles. This process is now a well recognised one known as dissolved-air (vacuum) flotation.

In 1906, Sulman et al developed a process whereby air bubbles were generated in the liquid by means of the shearing action of revolving impellers. Later, Callow (1914) introduced the air bubbles through submerged porous diffusers. These two methods are known as dispersed-air flotation, and today, they are used widely in the mining industry.

The development of all the above processes was confined to the problem of separating the desired minerals from the gangue material in the mining industry. Later developments in the mining industry appeared to concentrate mainly on the problem of enhancing the selectivity of the bubbles for the desired mineral particles. This led to the development of specific chemical additives (conditioning chemicals) which are called Frothers, Collectors, Activators and Depressants. The functions of these chemicals are as follows(2,3,4,5):

Frothers promote stable bubbles at the liquid surface. That is, they prevent the bursting of the bubbles which would result in the attached mineral particles being lost.

Collectors alter the surface properties of the mineral particles so that they are preferentially accepted by bubbles. For specific minerals there are, necessarily, specific collectors.

Activators enhance the effect of collectors.

Depressants serve to depress the attachment of the undesired gangue material to the bubbles.

During the late 1920's, flotation began to be used outside the mining industry. The Scandinavian paper industry used a dispersed air flotation technique for recovering paper fibres from paper mill wastes(6).

In the late 1930's(6), flotation began to be used in treating wastewaters. Here, the selective recovery of material is not the criterion. The problem is merely to remove as much of the particulate material as possible. Because selective flotation is not necessary or indeed, desirable, treatment of wastewaters by flotation in this respect is easier than its application in the mining industry. However, there is a restriction on the use of chemicals which must not impart toxicity, taste, colour or odour to the water. This thesis will concentrate solely on the process of flotation as applied to wastewater treatment.

CHAPTER 3 - THEORY OF FLOTATION

The theory of flotation applied to water and wastewater treatment requires the study of three phases : gas, liquid and solid(7). The gas phase is usually air, the liquid phase water and the solid phase particulate matter. Prior to flotation, waste water particles are usually flocculated to a certain degree. Flocculation may occur by natural biological activity (such as in activated sludge) or it may be induced by the addition of coagulant chemicals. When the (flocculated) wastewater particles are introduced to the air bubbles (by whatever means they are generated) then bubble-particle attachment may occur by three mechanisms(7):

(1) Trapping of air bubbles in the irregularities of a floc structure;

(2) Trapping of air bubbles within the floc structure as the flocs form;

(3) Adhesion of the air bubbles to the particles.

Mechanisms (1) and (2) arise merely from physical capture and therefore depend on the process of flocculation. Mechanism (3), the phenomenon of adhesion, merits elaboration:

There is necessarily a sequence of events resulting in adhesion. Initially there must be an encounter between a rising bubble and a suspended particle. When contact has occured, the next requirement for adhesion is that the film of water between the bubble and particle must be removed so that the air takes the place of the liquid at the particle surface. One might be tempted to say that due to the upward inertia of the bubble the liquid film is "squeezed" aside, resulting in contact, whereupon adhesion takes place. However, the following experiment, conducted by Evans in 1954(8) shows that the situation is not so simple. A vertical tube, connected to an air-pump, was immersed in water and an air bubble formed at the end of the tube. Immediately above the bubble a rotating silica disc was situated, also immersed in the water. The periphery passed over the bubble at a distance just sufficient to prevent the bubble being dislodged by the drag of the water carried with the disc.

The speed of rotation of the disc and its position were adjustable. If the disc was treated at a spot on the periphery such that it became hydrophobic, then the instant that the hydrophobic spot passed over the bubble, the latter dislodged itself from the end of the tube and adhered to the disc at that point. From this, it seems that under favourable circumstances, there is a direct attraction between a bubble and the solid particle surface.

The source of the attraction between bubbles and particles arises from electrical phenomena. Bubbles, in water, carry a negative surface charge. This is due to the orientation of the water molecule dipoles at the air-water interface(9). The negative (oxygen) end of the dipole generally predominates on the air side of the interface while the positive (hydrogen) end exists as a diffuse layer within the water. Due to the surface charge carried by bubbles, they carry layers around them analogous to the electrical double layer surrounding colloids.

In Derjaguin's theoretical treatment of bubble-particle adhesion(10,11), the space around the bubble is divided into three zones (see Figure 1). The forces acting in each zone are then considered. The theory assumes the presence of ions in solution, either naturally present, or arising from the introduction of (say)

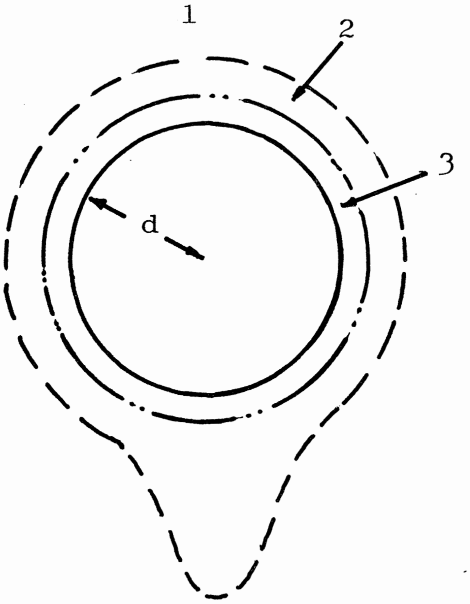


FIGURE 1: Designation of Zones Around a Bubble.

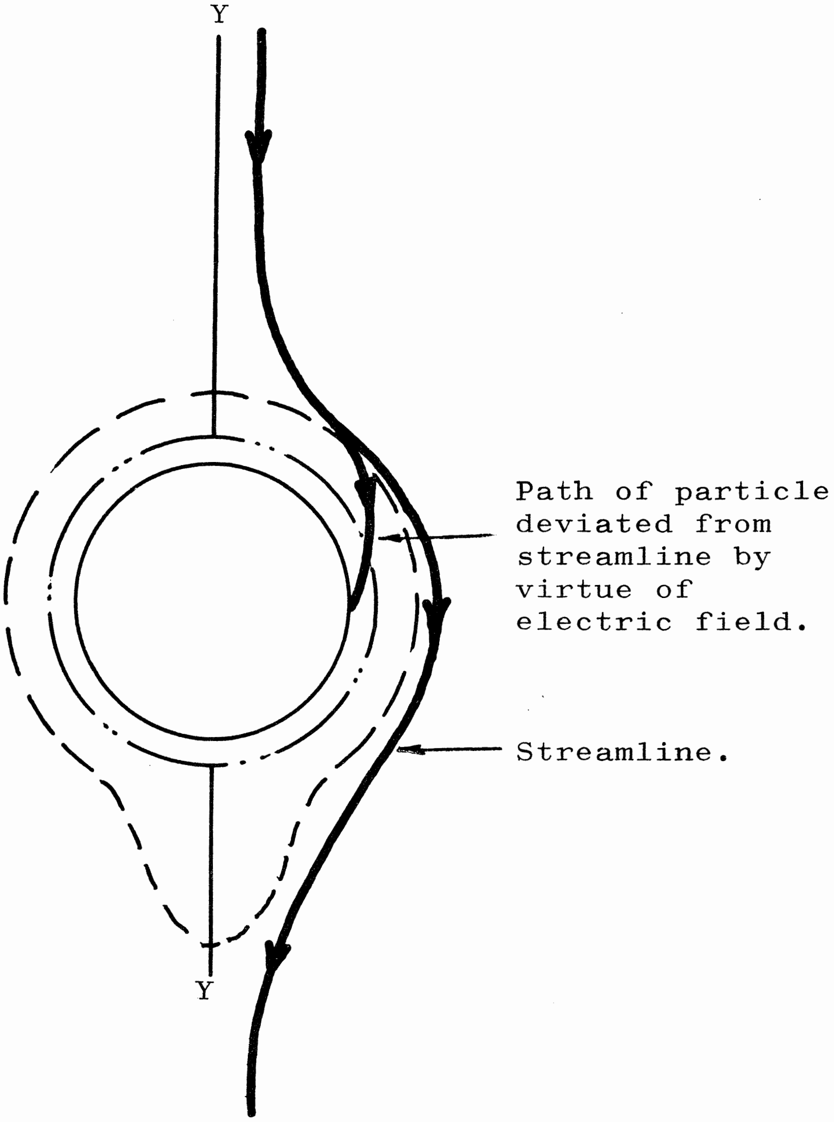


FIGURE 2: Effect of Diffusio-Phoretic Layer on the Path of a Particle in a Streamline.

metal coagulants.

Zone 1 constitutes the bulk of the liquid (outside zones 2 and 3). The forces acting in zone 1 are of a hydrodynamic origin. Whether a particle in the path of the bubble would strike the bubble or be swept past with the streamlines depends on the relative effects of inertia and viscous drag. If hydrodynamic forces only are present, the particles below a critical size can never approach close enough to the bubble for adhesion to occur since they would always be swept past in the streamlines: It has been shown mathematically(1) that unless a particle is located exactly in the path of the centre of the bubble (on axis YY of Figure (2)), then it can never come in contact with the bubble if hydrodynamic forces only are acting.

Zone 2 constitutes a diffusio-phoretic layer. Small particles reaching the outer boundary of zone 2 come under the influence of a diffusio-phoretic force (see Figure 2). This force arises as follows: As the liquid flows around the bubble a tangential current is set up just beyond the surface and a layer of ions is swept from the front of the bubble to the rear. The front of the bubble becomes undersaturated with respect to the concentration of ions in solution and the rear supersaturated. A situation thereby arises whereby ions are introduced to the front of the bubble from the liquid and ejected from the rear of the bubble into the liquid. The overall effect of this is to set up an electric field in which the ions move in a manner similar to the phenomenon of electrophoresis which gives rise to the diffusio-phoretic forces. Small particles, less than 10 micrometers, in the presence of these forces move out of the streamlines produced by the bubble's motion towards the bubble's surface.

Once a particle has been deviated from its streamline by the diffusio-phoretic forces (through zone 2), it eventually reaches the outer boundary of zone 3. In zone 3 there exists an electro-phoretic layer which induces a force between particle and bubble of the same nature as that existing between colloids, that is, an attractive or repulsive force generated from the interaction of electrical double layers.

With hydrophobic particles, attachment occurs spontaneously unless there exists a strong repulsion due to electrical double layers of like sign(10). With hydrophilic particles, strongly bound water layers may prevent bubble-particle attachment unless there exists a strong attraction due to electrical double layers of opposite sign. (See Appendix I).

The above theory by Derjaguin is applicable to particles of radius less than 10 micrometers - this being the thickness of the diffusio-phoretic force layer around the bubble. For particles of larger size Derjaguin extended his treatment which is summarized as follows:

When a large particle penetrates the diffusio-phoretic layer, ions are desorbed from the surface of the particle and are diffused to the surface of the bubble, where they are subsequently adsorbed. The part of the bubble accepting these ions thereby acquires a charge which corresponds to that of the ions most readily diffused. An equal and opposite charge is set up on that portion of the particle where desorption has taken place. An estimate of the electric field set up by this process has been given a value in the order of thousands of volts per centimetre The considerable electric forces arising from this situation results in a subsequent thinning of the water layers separating the particle and bubble surface, in two ways:

(1) The "spot"of charges on the bubble surface attract the "spot" of opposite charges on the particle hence squeezing aside the water layer.

(2) The "spot" of charges acquired by the bubble, being of mutual like sign, experience mutual repulsion which causes a state of radial tension concentrated at that spot. This radial tension is transmitted to the film of adjacent liquid thereby tending to make it thinner.

When the above phenomena have finally brought about contact between a bubble and a particle, that is, when the separating film of liquid between them has ruptured and broken away, the interfacial tension forces related to the solid-liquid and liquid-air interfaces are now brought into play, as shown in Figure 3.

Algebraically, the equilibrium between the interfacial tension forces shown in Figure 3 may be expressed as

$$T_{SA} = T_{SL} + T_{LA} \cos \theta$$

where

T_{SA} = interfacial tension force between the solid and air.

T_{SL} = interfacial tension force between the solid and liquid.

T_{LA} = interfacial tension force between the liquid and air.

S = area of contact of the air phase with the solid.

θ = contact angle, i.e., the angle formed between the air-liquid interface and the solid-liquid interface at the point where the three phases make contact.

If T_{SA} is equal to or less than T_{SL}, the contact angle is zero or cannot exist, in which case a liquid film prohibits adhesion between the solid and air phases. When

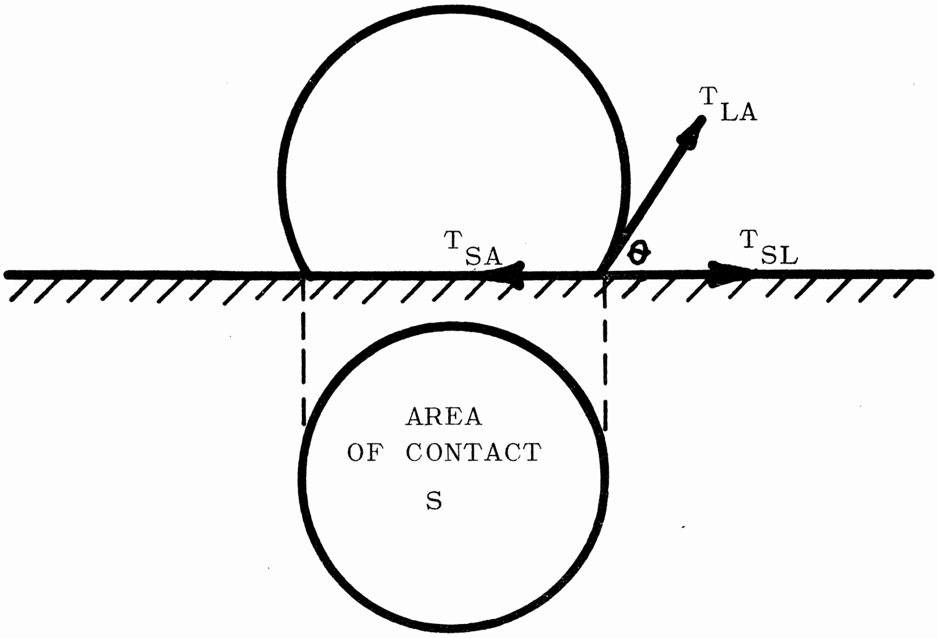


FIGURE 3: Conditions of Attachment of a Bubble to a Solid Surface.

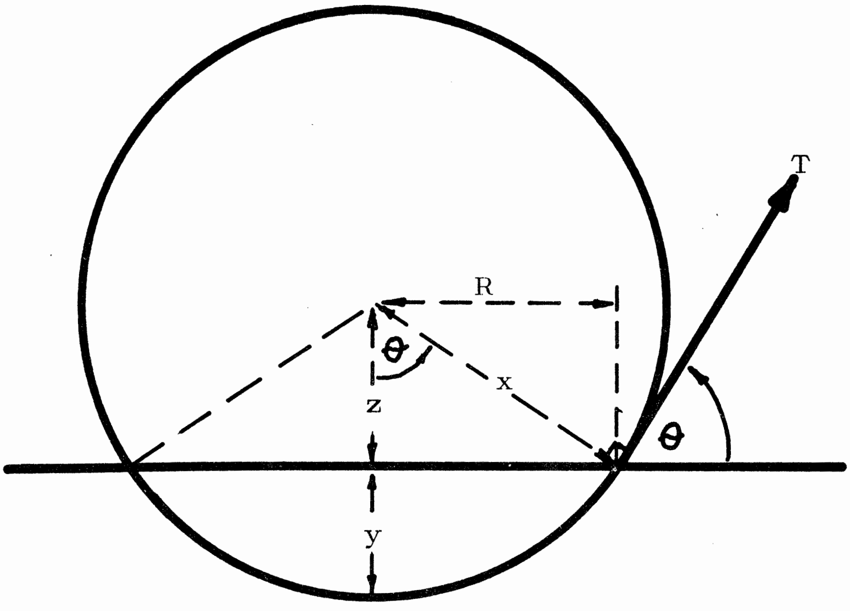


FIGURE 4

T_{SA} is greater than T_{SL}, the contact angle is greater than zero and the air bubble can adhere to the solid particle. The tenacity of adhesion of the air bubble to the solid may be deduced to be directly proportional to the contact angle:

From Figure 4:

$$z = \frac{R}{\tan \theta} \quad ; \quad x = \frac{R}{\sin \theta} \quad ; \quad y = x - z \quad .$$

$$\therefore \quad y = \frac{R(1-\cos \theta)}{\sin \theta}$$

Area of contact = $S = \pi R^2$

$$\therefore \ R = \left(\frac{S}{\pi}\right)^{\frac{1}{2}}$$

Since the interfacial tension force, T, is defined as force per unit length, the force tending to resist dislodgement of the bubble from the solid is given by

$$F = 2\pi RT \sin \theta$$

$$= 2(\pi S)^{\frac{1}{2}} T \sin \theta$$

If the bubble is dislodged from the solid, then the average force needed to accomplish this is given by

$$F_{average} = \tfrac{1}{2}\, 2(\pi S)^{\frac{1}{2}} T \sin \theta$$

The work done in dislodging the bubble = $F_{average} . y$

$$= (\pi S)^{\frac{1}{2}} T \sin \theta . y$$

$$= (\pi S)^{\frac{1}{2}} T \sin \theta \left(\frac{S}{\pi}\right)^{\frac{1}{2}} \frac{(1-\cos \theta)}{\sin \theta}$$

$$= ST\ (1-\cos \theta)$$

Hence the work done in pulling the bubble from the surface is directly proportional to (1-cos θ).

For the range of θ commonly encountered in flotation(3), the curve of θ versus (1-cos θ) is substantially a straight line. The work required to pull a bubble from a surface, therefore, is directly proportional to the contact angle. From this, one may say that the larger the contact angle, the larger is the adhesional force tending to hold the bubble to the particle, or the greater is the tendency for the bubble to adhere to the particle.

Certain conclusions may be drawn from the above discussions:

(1) It would seem that particles may be floated most efficiently when the electrical double layers surrounding bubbles and particles are of opposite sign. Furthermore, adhesion would occur more spontaneously when the particles are least wetted. In water and wastewater treatment both these requirements may be met by the use of metal coagulants. However, the optimum conditions of coagulant type, dosage and pH for flotation will be more stringent or, at best, coincide with those conditions for optimum colloid destabilization since one is dealing with electrical double layers surrounding particles and bubbles rather than considering particles only.

(2) The larger the contact angle between the solid, liquid and air phases, the greater is the efficiency of bubble-particle attachment. It has been stated(13) that "the purpose of having air bubbles as small as possible is to increase the contact angle ...... If the contact angle is large, there will be a tendency for the bubble to cover more unit area per unit bubble surface area and thereby increase the contact strength between the two". The first part of the statement, as it stands, appears to be rather obscure. Under normal circumstances, the contact angle formed between three phases is a

function solely of the three phases, apparantly independent of the bubble size. The second part of the statement, however, can be explained. From Figure 3, one can visualise that the larger the contact angle, θ, the larger is the area of contact, S, between bubble and particle. The way in which this advantage of a large contact angle may be brought about by reducing the bubbles' size, however, is not clear. There are, however, definite advantages of having smaller sized bubbles. The first is the relatively low degree of turbulence induced by smaller bubbles. Their orderly, slower rate of rise enhances bubble-particle attachment and the reduced buoyancy force of <u>each</u> bubble reduces the tendency of the bubble to tear away from the particle. This does not necessarily mean, however, that the rate of rise of bubble-particle agglomerates is reduced since there may be several bubbles attached to each particle. This last statement leads to a second advantage of small bubbles. For a given volume of air introduced to a liquid then the smaller the bubbles' size, the greater will be the total interfacial area between air and liquid. This results in an increased contact opportunity between bubbles and particles which, consequently, will result in more efficient use of the introduced air. "Smaller bubbles have a better opportunity to become attached to particulate matter and therefore are likely to result in higher solids removal"(1).

CHAPTER 4 - METHODS OF FLOTATION

In wastewater treatment there are four methods of flotation which may be used. They are dispersed-air flotation; electrolytic flotation; dissolved-air (vacuum) flotation and dissolved-air (pressure) flotation.

The differences between these methods are the ways in which the bubbles are generated in the liquid. Each method will be briefly described in turn together with their inherent advantages or disadvantages.

Dispersed-air flotation

In this method, high speed impellers, immersed in the liquid, are employed. Air is introduced to the liquid either by injecting compressed air near the impeller, or it arises from the suction (vortex) action of the impellers. The shearing action of the impeller disperses the air into small bubbles. The liquid, together with the generated bubbles, is then passed to the flotation chamber where the bubbles become attached to wastewater particles.

An alternative method of dispersed-air flotation is to generate bubbles by introducing air through porous diffusers. However, this has the disadvantage that the diffusers tend to clog rapidly(12).

A disadvantage of the dispersed-air technique is the relatively large size of the bubbles produced compared to other methods. From Chapter 3, this results in comparatively inefficient bubble-particle attachment. (In order to obtain finer bubbles, soap or detergents are sometimes added). A further disadvantage of the dispersed-air technique using impellers is that the process tends to generate a comparatively high degree of turbulence(13) which impairs bubble particle attachment and bubble-particle agglomerate growth.

Electrolytic Flotation

In electrolytic flotation, bubbles of gas are generated in the liquid by passing a direct current between two electrodes immersed in the liquid. Under the normal process of electrolysis in water(14), bubbles of oxygen are formed at the anode due to the discharge of OH^- ions ($4OH^- \longrightarrow 2H_2O + O_2 + 4e^-$). At the cathode, bubbles of hydrogen are formed. The anode may be made of a material which does not erode (such as platinum) or it may be made, to advantage in some cases, of a material which emits certain metal ions into solution. When metal coagulants are necessary for flotation, the dosage of coagulant may be significantly reduced or eliminated by installing "sacrificial" iron or aluminium anodes(15). Iron or aluminium ions are liberated from the anode into the wastewater stream together with the generated gas bubbles. The cost of the anodes, which need to be replaced periodically, may in certain cases be less expensive than using coagulant chemicals. With this method, no anions (such as $SO_4^=$ or Cl^-) are introduced which in some cases, may have a detrimental effect on bubble-particle adhesion. Furthermore, sacrificial anodes are preferable when it is necessary to prevent the dissolved solids concentration from becoming too high.

A possible disadvantage of this method is that normal wastewaters may not be sufficiently electrically conductive and, consequently, the inorganic salts concentration may have to be raised by the addition of salts. A proposed method of increasing the conductivity(16) was to mix the wastewater with sea water. With this method, bubbles of chlorine are generated at the anode and bubbles of hydrogen at the cathode.

A further disadvantage of this process is that the electrical power requirements to produce sufficient gas for flotation may be excessive compared to other methods of flotation. A possible application is to use this technique in a supplementary capacity to other methods of flotation. The prime objective would be to reduce the dosage of metal coagulants for flotation.

Dissolved-air (vacuum) flotation

With this method of flotation, air is dissolved into the wastewater stream at atmospheric pressure and then passed to a vacuum chamber where the air, now super saturated, precipitates out of solution(7). This method has the advantage over dispersed-air flotation in that the bubbles produced are of very small size. Due to the presence of particulate matter in the wastewater, the bubbles tend to precipitate on the particulate nuclei provided. This is a more positive means of bubble-particle attachment than the former processes since precipitation occurs within the body of the wastewater. The process, therefore, does not rely to the same extent on bubble-particle contact prior to adhesion.

The degree of turbulence induced by the precipitating bubbles is less than that in the dispersed-air technique and hence there is less tendency to disrupt any floc particles present.

The major disadvantages of this process are, firstly, the expense necessary for constructing the vacuum chambers, and secondly, the maximum practical pressure drop possible is only in the order of one-atmosphere(7). This small pressure drop may, in some cases, provide insufficient precipitated air for flotation of all the wastewater particles.

Dissolved-air (pressure) flotation.

The process which has gained the widest acceptance in wastewater treatment is dissolved-air (pressure) flotation(7). In this method, air is dissolved into the wastewater (or a portion of the wastewater, or a portion of the effluent from the treatment plant) under pressures higher than atmospheric pressure. The saturated (pressurized) stream is then mixed with the wastewater stream and passed to a flotation chamber where the pressure is essentially atmospheric. At this lower pressure, the air, which is now supersaturated, precipitates out of solution in the form of minute bubbles. Since the bubbles are generated in a manner similar to that in the "vacuum" method, the advantages described for the "vacuum" method are also inherent in the "pressure" method. However, with this method, the flotation chambers are of a simpler design since no vacuum has to be applied — "pressure" flotation is carried out under atmospheric conditions.

With dissolved-air (pressure) flotation a saturation unit is necessary to dissolve the air into the water under pressure up to saturation. However, in most cases, only a comparatively small pressurized stream is required since there is (within reason) no limit to the pressures which may be used. Consequently, the amount of air precipitating out of solution may be varied at will, depending on the conditions, by simply adjusting the pressure under which the air is saturated. Pressurization of the whole wastewater stream is therefore unnecessary and indeed, undesirable. Consequently, the saturation unit is usually small compared to the wastewater flow and therefore need not be a major factor in the capital cost of the plant.

Due to the inherent advantages of enhanced control and simplicity of operation in the dissolved-air (pressure) flotation process, this was the method used for the investigations to be described subsequently. Consequently, the remainder of this literature review will concentrate solely on the dissolved-air (pressure) flotation technique.

CHAPTER 5 - PRACTICAL CONSIDERATIONS FOR DISSOLVED-AIR (PRESSURE) FLOTATION

(See reference 7):

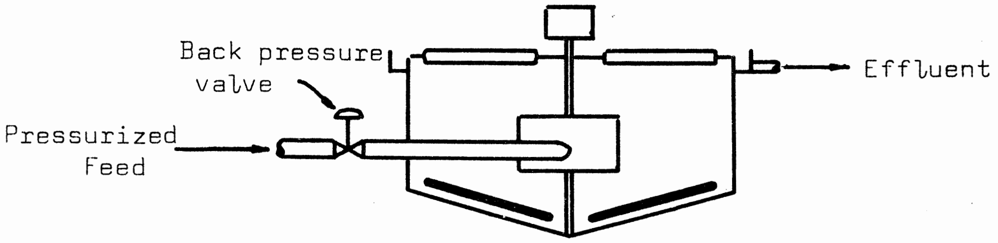
The process of dissolving air into water under pressure may be classified into two general headings : Total Pressurization and Partial Pressurization. (See Figure 5).

Total Pressurization

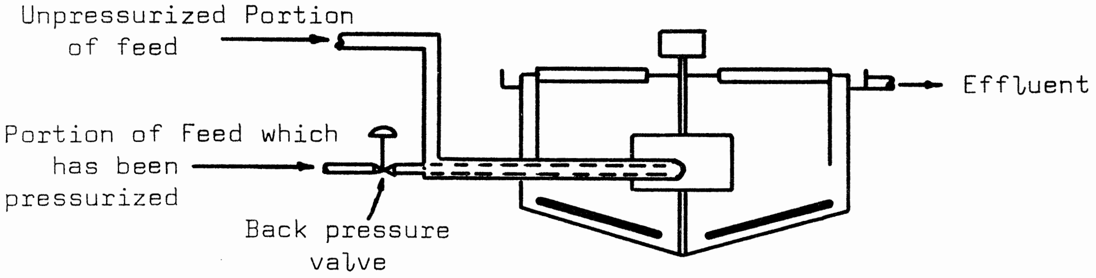
With total pressurization, the whole wastewater flow is pressurized (and saturated) to obtain the required mass of precipitated air. A disadvantage of this method is that if flocculation of the particles prior to flotation is considered important for efficient solids removal, then flocculation should take place prior to introduction in the flotation tank. The flocculated particles, therefore, are best fed to the flotation tank by gravity since pumping into the pressurized saturation unit would destroy the flocs. Further, if wide variations in wastewater flow are anticipated then total pressurization is impractical.

Partial Pressurization

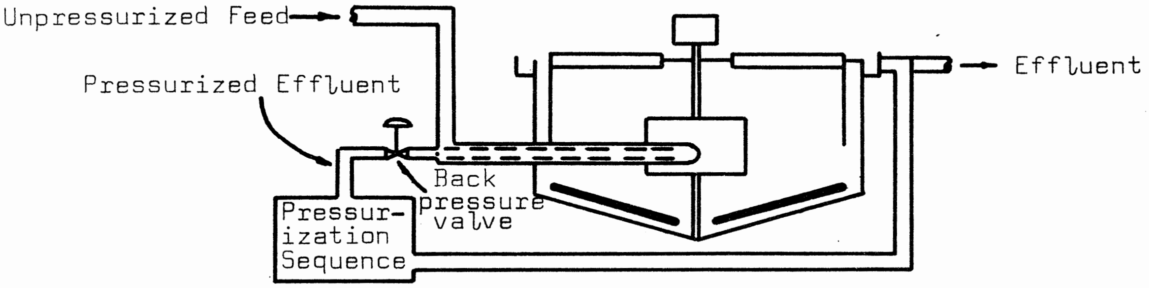
With partial pressurization, either a part of the waste water flow is pressurized and saturated or, alternatively, a part of the clarified effluent is pressurized and saturated. In either case, the pressure-saturated stream is blended into the normal wastewater stream in the flotation tank (see later). Recycling a portion of the effluent to the pressurized saturation unit has the disadvantage that the total hydraulic flow into the flotation tank is increased and thus the size of the flotation tank would have to be increased. Usually, however, the recycle ratio is not great and thence may not be of major consequence (it depends on the amount of precipitated air required and the pressure at which the air is saturated).



Total Pressurization



Partial Pressurization of Feed



Partial Pressurization of Effluent

FIGURE 5: Methods Employed for Partial and Total Pressurization (see ref.7).

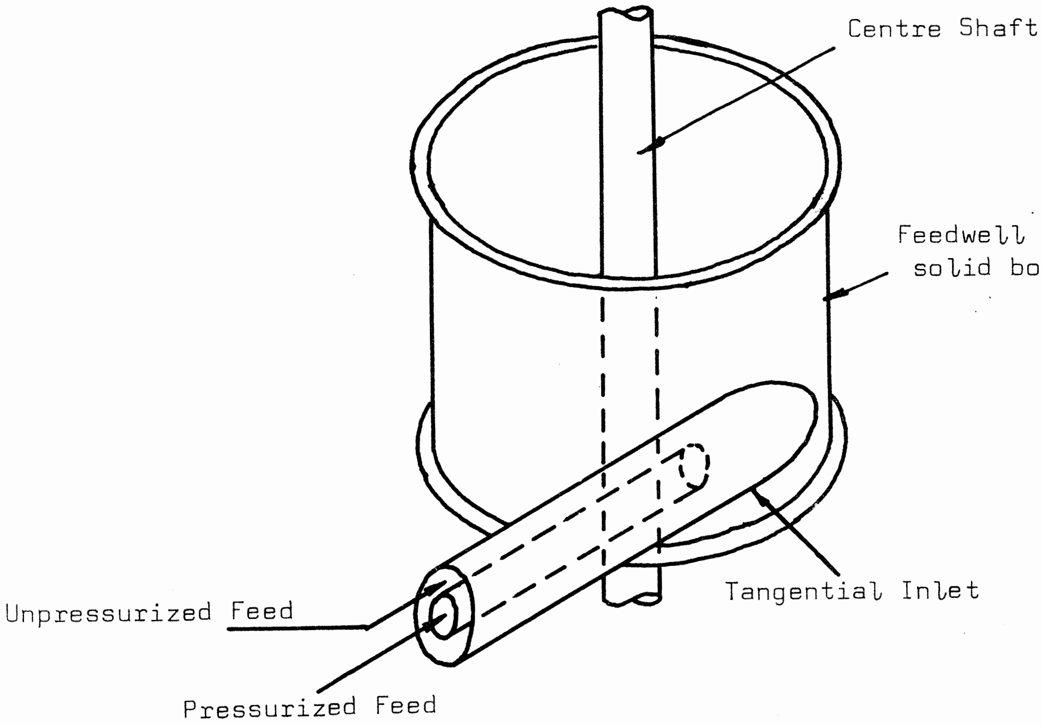


FIGURE 6: Partial Pressurization Feedwell (see ref 7).

With partial pressurization, greater amounts of air need to be dissolved into the pressure-saturated stream to maintain the required mass of precipitated air. Consequently, with partial pressurization, higher pressures need to be used than with total pressurization.

In partial pressurization, it has been suggested that the blending of the pressure-saturated stream with the normal wastewater stream is most efficient when concentric piping is used (see Figure 6). The pipe carrying the normal wastewater stream surrounds the pipe carrying the pressure-saturated stream with the velocity in the latter exceeding the velocity of the former by a ratio of 3:2. This results in the pressure-saturated stream diffusing with a jet stream pattern into the normal wastewater stream with a high degree of mixing. The pipe carrying the pressure-saturated stream is stopped short from the outlet of the pipe carrying the normal wastewater stream at such a distance that most of the blending actually takes place within the pipe.

The back pressure valves shown in Figure 5 maintain a constant discharge pressure in the pressure-saturation unit. If a constant hydraulic flow is anticipated then ordinary gate or needle valves may be used. For varying flows, however, an automatic (spring loaded) back pressure valve is required.

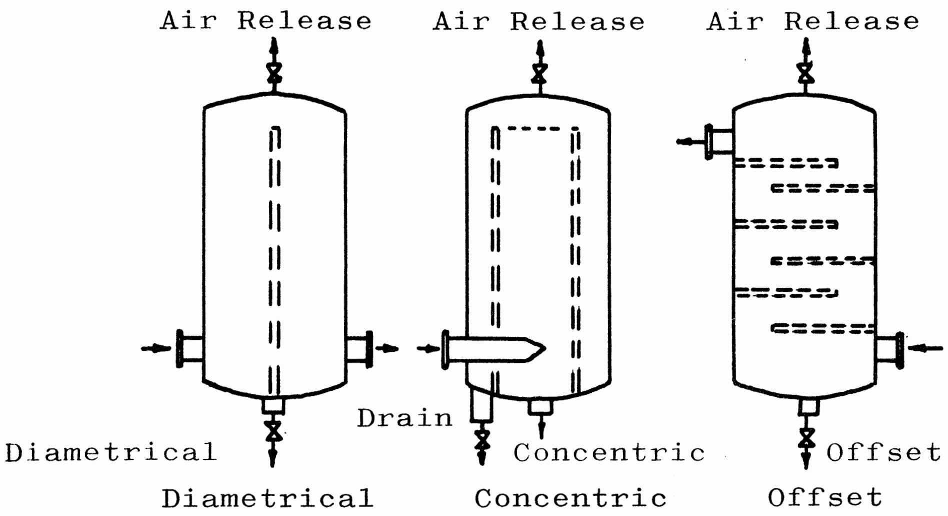
If a central-feed type flotation tank is proposed, then a circular feedwell may be provided (see Figure 6). The pipe carrying the influent (blended) stream is arranged tangentially to the feedwell. This is in order to utilise the full volume of the feedwell and to ensure that the rising bubble-particle agglomerates are allowed to proceed with minimum turbulence into the flotation tank. This prevents any tendency for the agglomerates formed to break up.

Air Dissolution

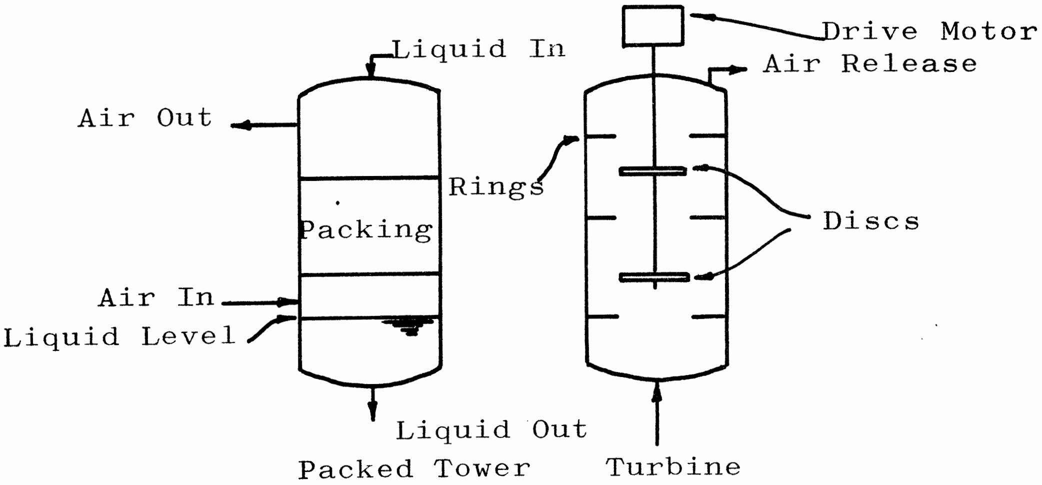
A common method of introducing air into liquid under pressure is to inject it into the suction side of a centrifugal pump. The impeller of the pump churns the air into the liquid and the mixture is passed to a retention tank (see later). Injecting air into the suction side of a pump has certain disadvantages. Firstly, air up to only approximately 6 per cent by volume may be injected safely before air-binding occurs in the pump. Consequently, strict control over the rate of air flow must be maintained and may prove difficult if one has a varying liquid flow. Secondly, air injection tends to lower the discharge rate and pressure of the pump.

At any particular pressure, the quantity of air necessary for saturation - in excess of that which may be injected into the suction side of the pump before air-binding occurs, is injected into the discharge side of the pump. No more than 25 per cent of the air required for saturation may be injected into the discharge side.

To ensure full dissolution of the air and to remove any excess undissolved air in the liquid, the mixture of air and liquid is passed, under pressure, to a retention tank or contact vessel. If undissolved air bubbles were to enter the flotation tank it would cause undesirable turbulence which could impair bubble-particle agglomerate formation and could disrupt the accumulated agglomerate layer at the top. Various retention tank designs are illustrated in Figure 7. Those designs employing baffles are such that the air-liquid solution is agitated as much as possible. This aids air dissolution and enhances the tendency for undissolved air bubbles to make contact, grow in size and rise to the top of the retention tank where the air is released by an automatic air-release bleed valve.



Static Designs

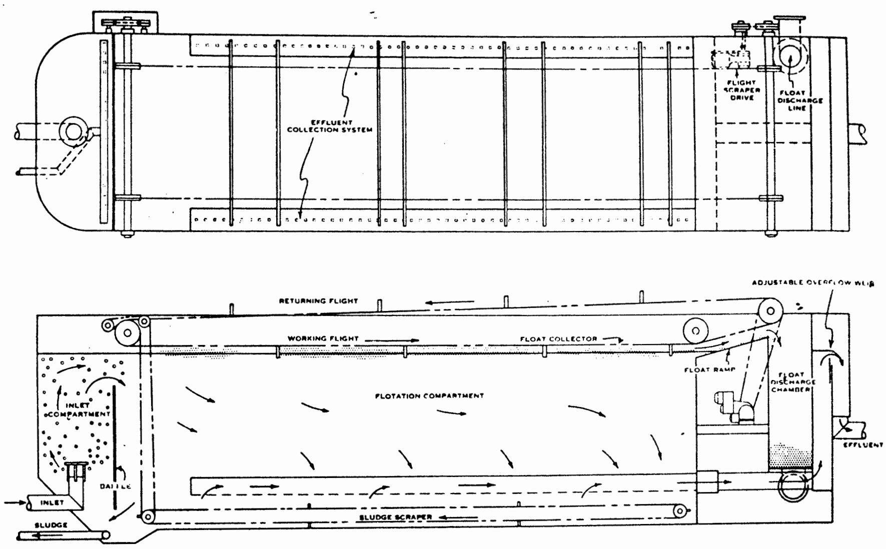


Dynamic Designs

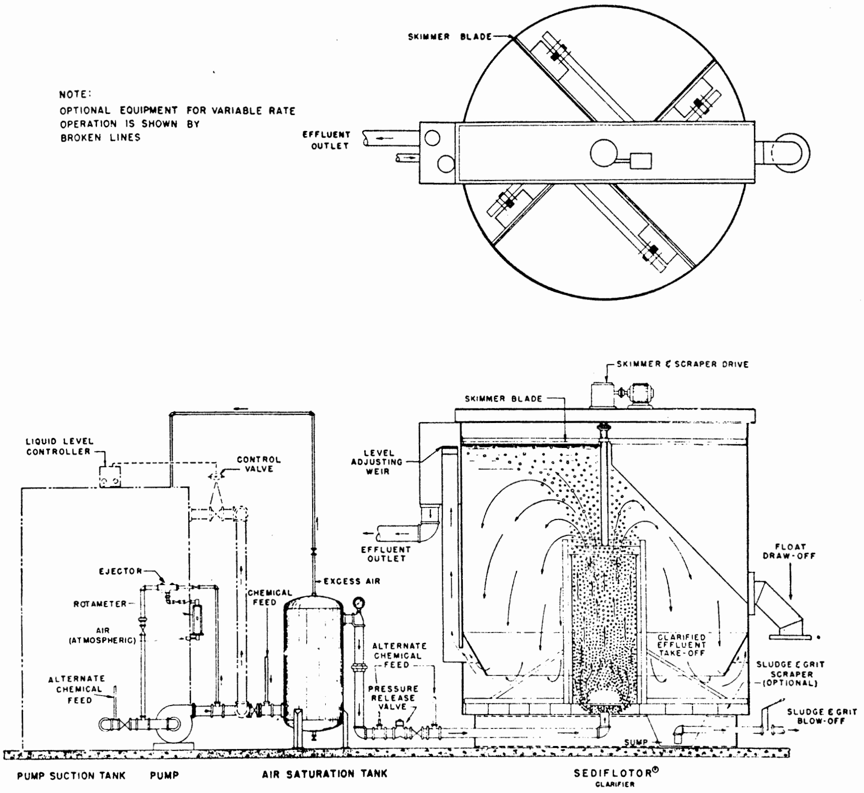
FIGURE 7: Retention Tank Arrangements (see ref.7).

Flotation Tanks

Dissolved-air (pressure) flotation tanks have been either of rectangular section, where the hydraulic flow is horizontal, or of circular section, where the flow is vertical (see Figure 8)(17). In either case, facilities are usually provided whereby solids which are non-floatable (such as grit etc) and settle out to the bottom of the tank, may be removed by sludge scraper mechanisms. A disadvantage of having rectangular tanks is that bubble-particle attachment must take place within the inlet compartment - no subsequent opportunity for attachment is provided. With vertical type flotation tanks, however, the "counter-current" flow regime provides an enhanced opportunity for contact.



A rectangular pressurized flotation unit.



Arrangement for pressurized flotation using a circular unit.

FIGURE 8: Examples of Dissolved-Air (Pressure) Flotation Tanks. (See reference 17).

CHAPTER 6 - EXPERIMENTAL INVESTIGATIONS

FLOTATION SYSTEM

The flotation system developed during these investigations is shown in Figure 9. It consists essentially of (1) a saturation unit and (2) a flotation unit. The saturation unit produces an air-saturated water feed under pressure, which, when introduced to the flotation unit, forms a precipitated mass of air bubbles. These bubbles attach themselves to the particulate constituents of a wastewater to form buoyant bubble-particle agglomerates which rise and are thereby separated from the wastewater stream.

The remainder of the apparatus consists of a flocculation chamber designed to provide a high degree of flocculation to the wastewater prior to its introduction to the flotation unit; rotameters to monitor the wastewater and saturator feed rates; and peristaltic pumps to introduce the coagulant chemicals to the wastewater.

Saturation Unit

From literature(7) it appears that the most common method of saturating water under pressure with air is to inject the air, with the water, into the suction side of a centrifugal pump and pass the mixture, under pressure, to a retention tank. The pump impeller "churns" the air into fine bubbles under pressure. The retention tank provides time for the air to dissolve into the water and allows excess undissolved air to escape.

This method has certain disadvantages(7):

(1) Air binding occurs in the pump if more than 6 per cent by volume of air is introduced.

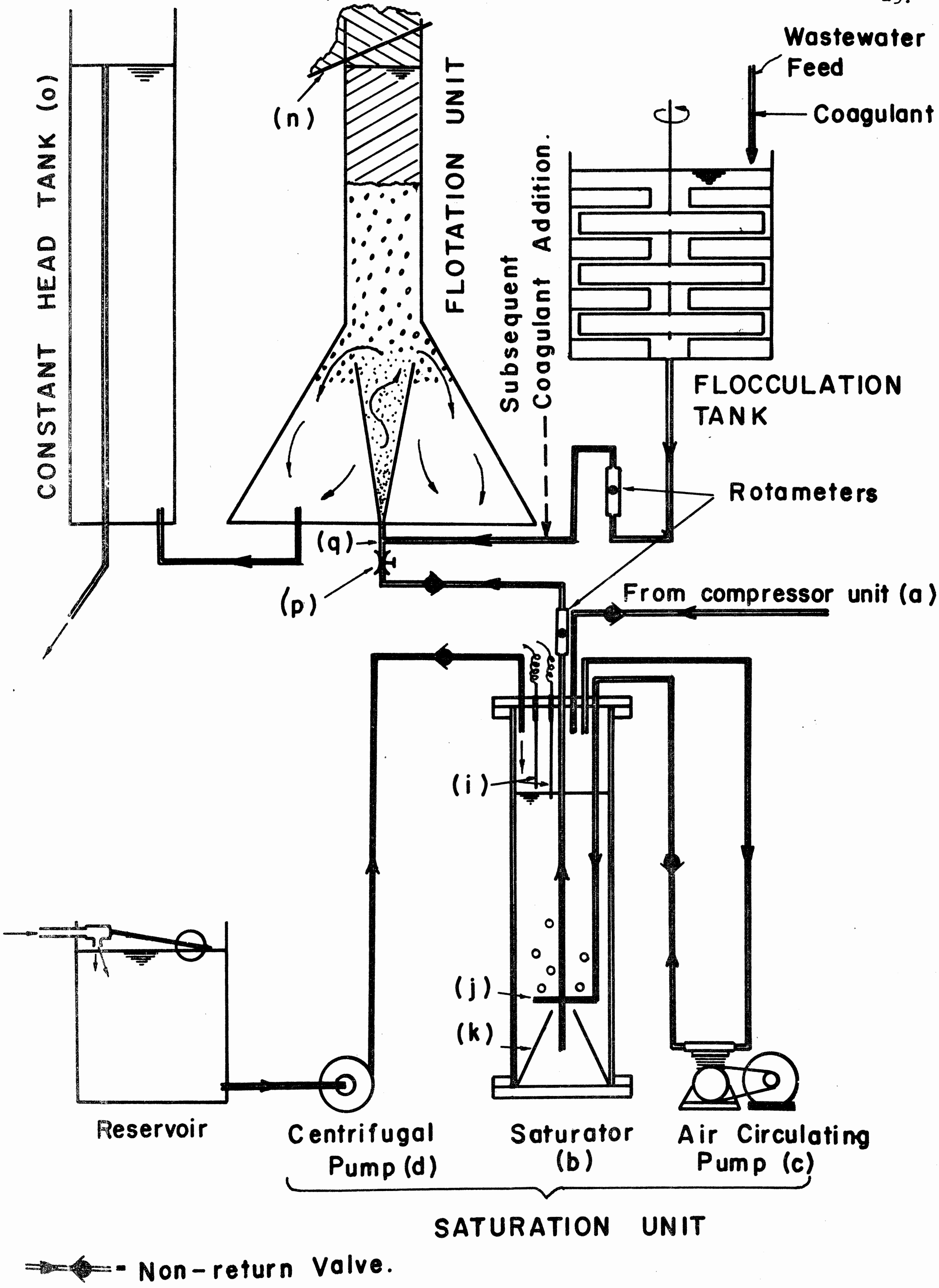
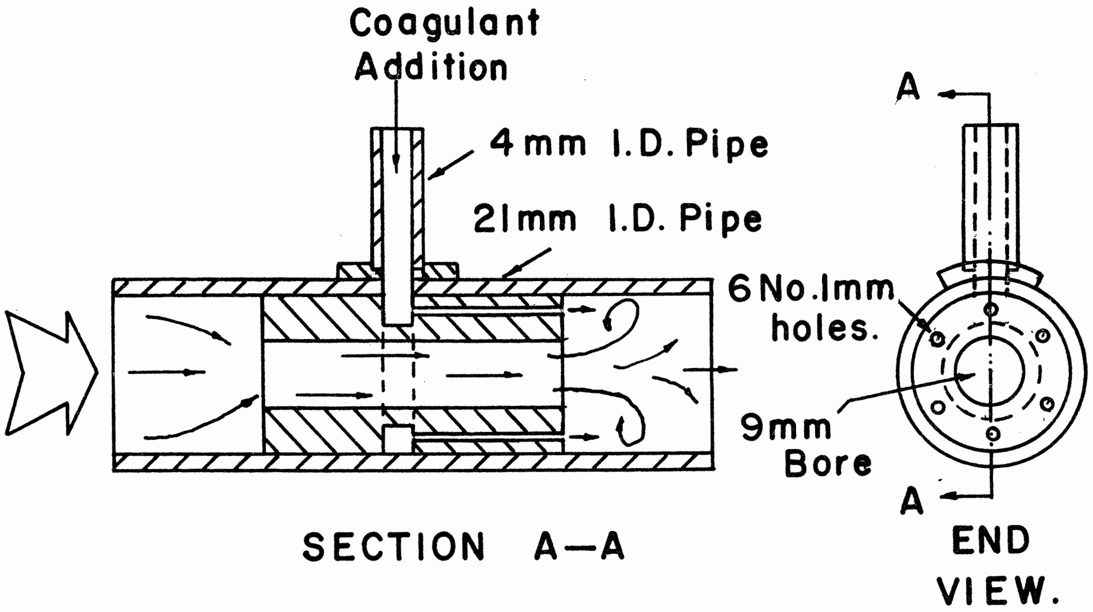


FIGURE 9 : General layout of Flotation system.



SCALE : Full size.

FIGURE 9(a) : Rapid Mixing Unit for Coagulant Addition to Waste Water Stream - from reference 19. (see p.43).

(2) Air injection lowers the discharge rate and pressure of the pump.

The saturation unit shown in Figure 9, is free from the above disadvantages. It consists of a standard compressor unit (a) with compressor, pressure cylinder and automatic pre-set (pressure) on-off unit; a saturator (b) in the form of a pressure cylinder; an air-circulating pump (c) and a centrifugal water pump (d).

The saturator is shown in detail in Figure 10. Compressed air is supplied to the saturator, at point (e) above the water level, by the compressor unit. The pressure in the saturator is that pre-set in the compressor unit, the pressure being indicated by the pressure gauge (f). A pre-set pressure release valve (g) is fitted for safety. A centrifugal pump (d) supplies water to the saturator through point (h) and two electrodes (i), one 14cm long and the other 16cm long, control the water level in the saturator. The electrodes are connected to an automatic "on-off" switch : when the water level drops below the longer electrode the pump is switched into operation and continues pumping until the water level reaches the shorter electrode, when the pump switches off. To prevent water flowing back from the saturator to the pump when it is not in operation a non-return valve is fitted to the feed line.

Air is bubbled through the water by means of an air-circulating pump which withdraws air from above the water level and re-introduces it via a non-return valve through a circular sparger (j) 15cm from the bottom of the saturator. The air-circulating pump operates within the pressurized circuit so that the volumetric rate of air bubbled through the water is independent of the pressure in the saturator - the air-circulating pump operates only against the hydrostatic head of the water in the saturator. The air supplied by the compressor unit, therefore, is only that dissolved and removed in the saturator feed. Normally, no air escapes from

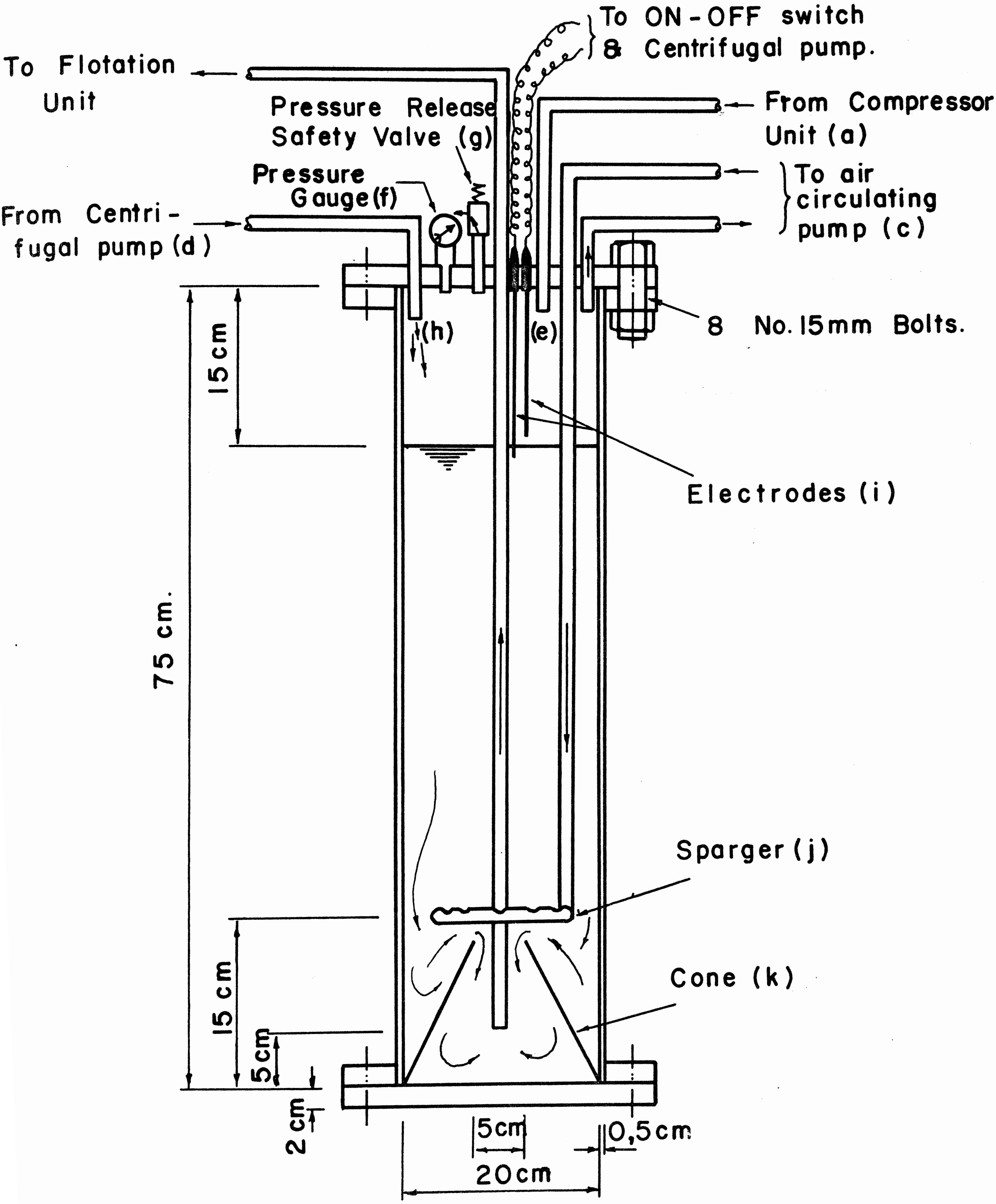


FIGURE 10: The Saturator.

the pressure release valve (g).

Undissolved air if allowed to escape to the flotation unit will give rise to undesirable turbulence and possible disruption of the float solids(7). For this reason, the volume of liquid beneath the sparger is maintained in a relatively quiescent condition by means of a truncated cone (k) placed on the floor of the saturator. The larger diameter of the cone is the same as that of the saturator and the smaller diameter, being uppermost, is 5cm. The considerable turbulence above the sparger may cause undissolved air bubbles to enter the cone but the quiescent conditions in the cone allow these bubbles to rise and pass out.

Pressure-saturated water is drawn from the saturator by a pipe reaching to within the cone and 5cm from the bottom. This pipe is connected, via a non-return valve and rotameter, to an 8mm needle valve at the inlet to the flotation unit.

In all the experiments of this investigation the saturator feed source was tap water. In the course of the tests under different loadings the effluent solids from the flotation unit varied from a maximum to a minimum and consequently if a part of the effluent was recycled as a saturator feed, an uncertain concentration of solids would be introduced with the raw influent.

Flotation Unit

The flotation unit is shown in Figure 11. It consists essentially of a lower chamber (l), of rectangular cross section, whose area increases with depth and a detachable upper chamber (m), of square cross section whose area is constant throughout its depth.

The shape of the unit evolved as follows: Preliminary experiments using a cylindrical chamber, with the discharge point of the saturator feed midway between the water

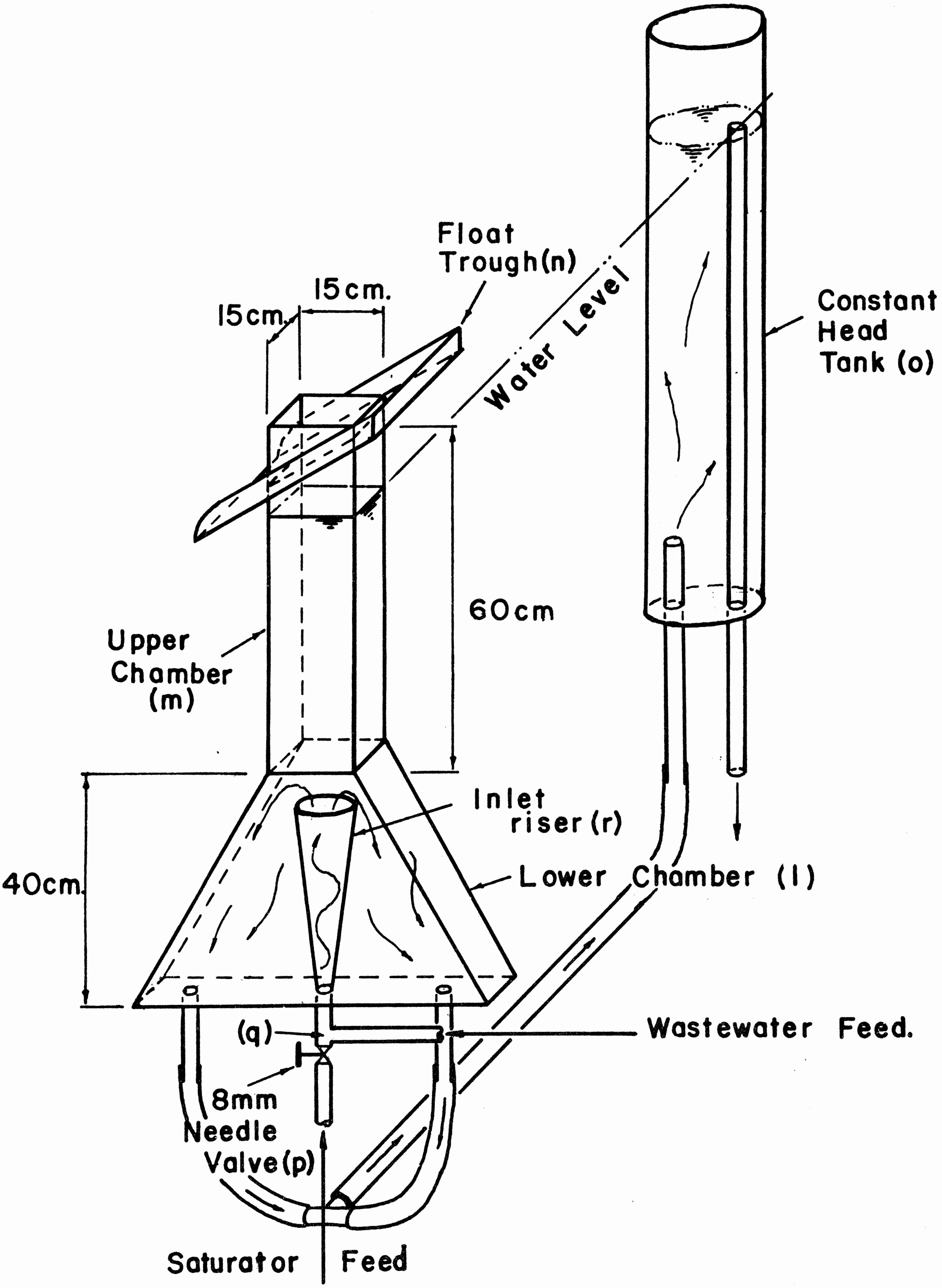


FIGURE 11: The Flotation Unit.

surface and the bottom of the chamber, indicated that at low feed rates, a stable interface between the rising bubbles and clear liquid formed at the discharge point. As the feed rate was increased the interface remained stationary until the downflow velocity slightly exceeded the rise velocity of the bubble mass whereupon the interface shifted suddenly to the bottom (discharge point) of the chamber. That is, the maximum hydraulic feed rate into the unit for a clear (bubble free) discharge is determined by the upward velocity of the bubbles (or bubble-particle agglomerates) and the cross-sectional area of the chamber, the latter fixing the downward velocity at the cross section where the drawoff is located. Consequently, in the flotation unit shown in Figure 11, the maximum hydraulic flow rate giving a clear water effluent is determined by the cross sectional area at the bottom. At intermediate flow rates, the interface forms at some level in the lower chamber, above the bottom, where the upward velocity of the bubble-particle agglomerate mass is equal to the downward velocity at that point.

The shape of the flotation unit provides three favourable features:

(1) It allows a stable interface to form below the point of discharge into the unit. This induces a countercurrent movement of liquid and bubble-particle agglomerates thereby providing increased opportunity for contact between bubbles and unattached particles — and agglomerate growth.

(2) In operation, the rolling effect of the agglomerates up the inclined sides of the unit, assisted in forming larger agglomerates whose rise velocity is greatly increased in the upper chamber.

(3) The rising agglomerates are concentrated into the comparatively small area of the upper chamber

where they accumulate, thicken and discharge freely at the top into a trough (n) without mechanical assistance.

A constant head tank (o), whose outlet pipe is adjustable, regulates the water level in the unit.

An 8mm needle valve (p) maintains a back pressure on the feed line from the saturator and regulates the saturator feed rate. Immediately downstream of the needle valve the wastewater feed is introduced (q) to mix with the saturator feed. The mixture enters the unit via an inlet riser (r).

EFFICIENCY OF THE SATURATION UNIT

The saturation unit serves to saturate air into water at pressures above atmospheric. When the saturator feed is introduced to the flotation unit it experiences a pressure drop from that in the saturator to essentially atmospheric pressure. This creates an unstable supersaturated condition whereupon the air precipitates out of solution in the form of very small bubbles.

One of the problems when investigating flotation phenomena is the measurement of the mass of air precipitating in the flotation unit. It is not practical to measure directly the air precipitated in the flotation unit because the precipitated air cannot be isolated from the float. A relative measure was obtained by determining the mass of air precipitated per unit volume of saturator feed, a_p, in the monitoring unit shown in Figure 12. It was connected in parallel to the flotation unit and received a constant saturator feed of 1 l/min which was regulated with an 8mm needle valve. The total saturator feed was therefore the sum of the feed rates to the monitoring unit (1 l/min) and to the flotation unit. a_p was determined by recording

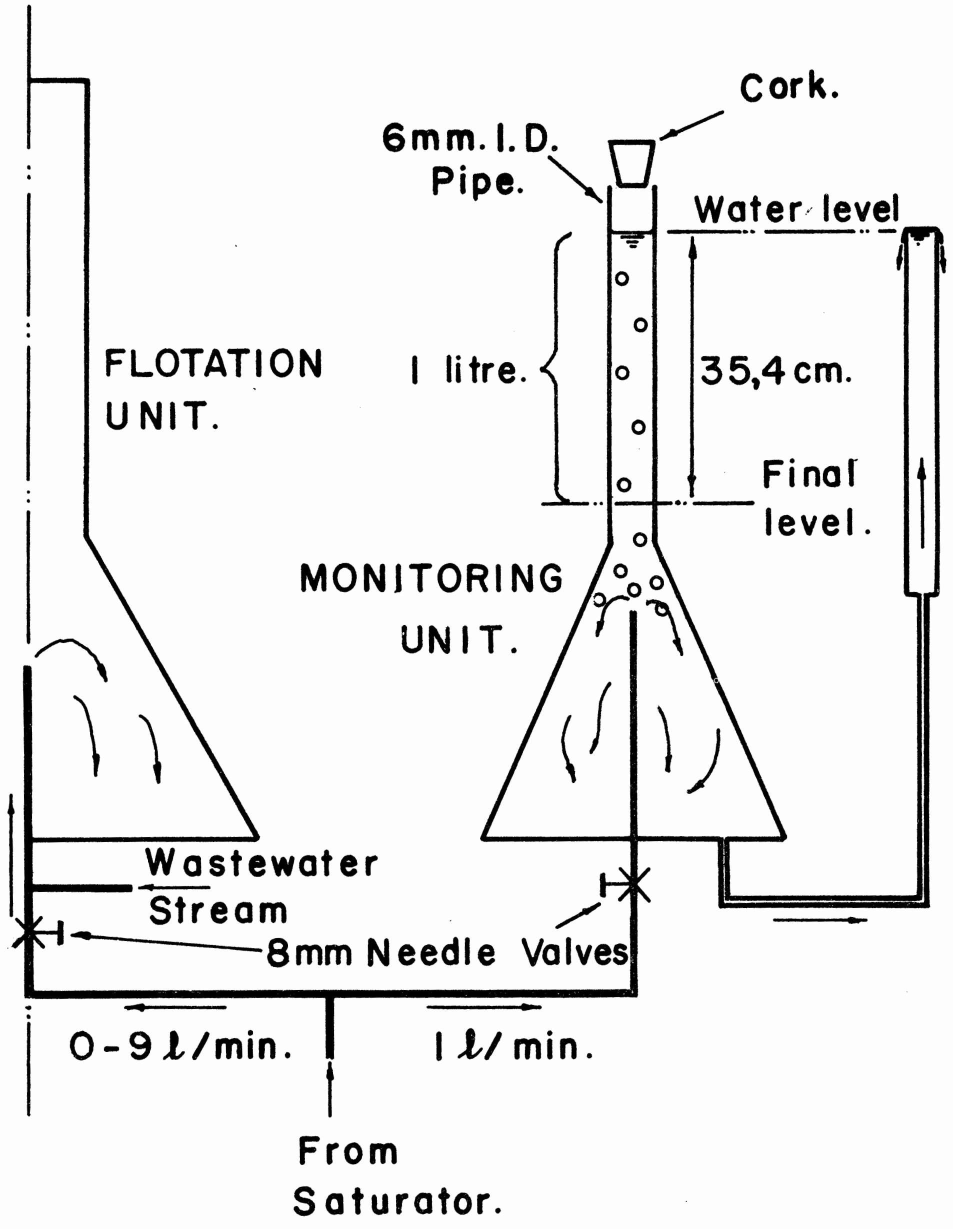


FIGURE 12: The Monitoring Unit.

the time taken for the precipitated air to displace one litre of water in the monitoring unit. From the time taken, the volume of saturator feed passed through could be determined and by correcting for hydrostatic head, the mass of air released per unit volume of saturator feed for the particular conditions of pressure and total saturator feed rate was computed. (*)

It was found that for any particular pressure in the saturator, a_p was constant for total saturator feed rates from 0 to 10 l/min (greater feed rates were unattainable with the 8mm needle valves used). Within this range, therefore, the water in the saturator was always fully saturated with air.

From Henry's Law, a linear relationship exists between the mass of air dissolved and the pressure, P_g, in the saturator. If, in the monitoring and flotation units, all the supersaturated air in solution is precipitated then a linear relationship should also exist between a_p and P_g, as shown by the broken line in Figure 13(a) (values taken from reference 7). This shows the values of a_p at equilibrium conditions. However, in Figures 13(a) and 13(b) it is evident that the relationship is not linear but of the form

$$a_p = K_2 (P_g)^{K_1} \qquad \text{———————} \quad (1)$$

where K_1 and K_2 are constants with empirical values $K_1 = 2{,}45$ and $K_2 = 0{,}0027$.

The mass of air precipitated is considerably less than the mass supersaturated in solution. This is particularly evident at low saturator pressures. As the saturator pressure increases the mass of air precipitated tends to approach the mass supersaturated in solution.

The reason for this is as follows: the rate at which the air precipitates out of solution is dependent (amongst other factors) on the degree of supersaturation which

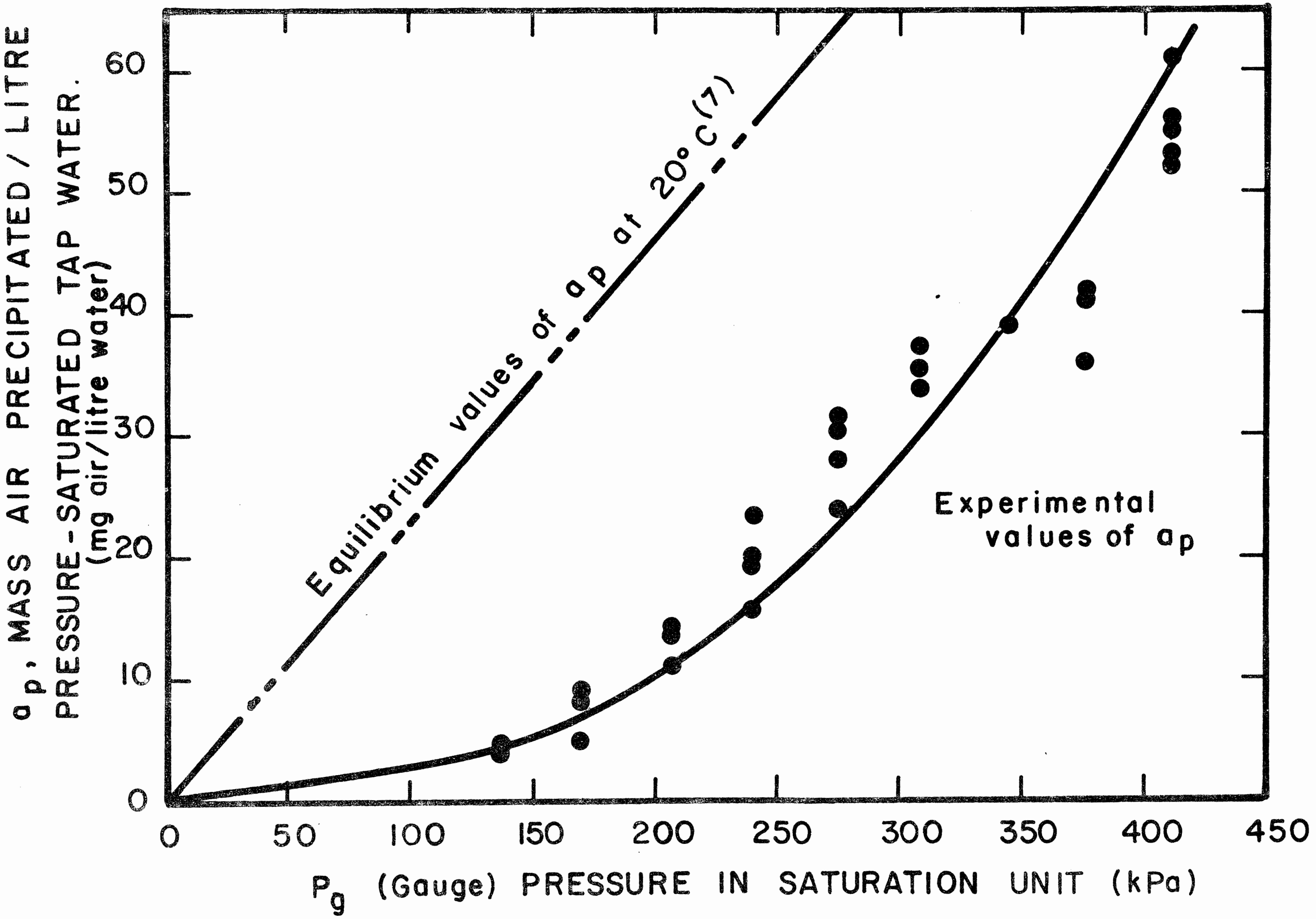


FIGURE 13(a): Mass of Air per litre of Pressure-Saturated Water Precipitated in Monitoring Unit at different Saturator Pressures (valid for Saturator feed rates from 0 to 10 l/min) (Table 1).

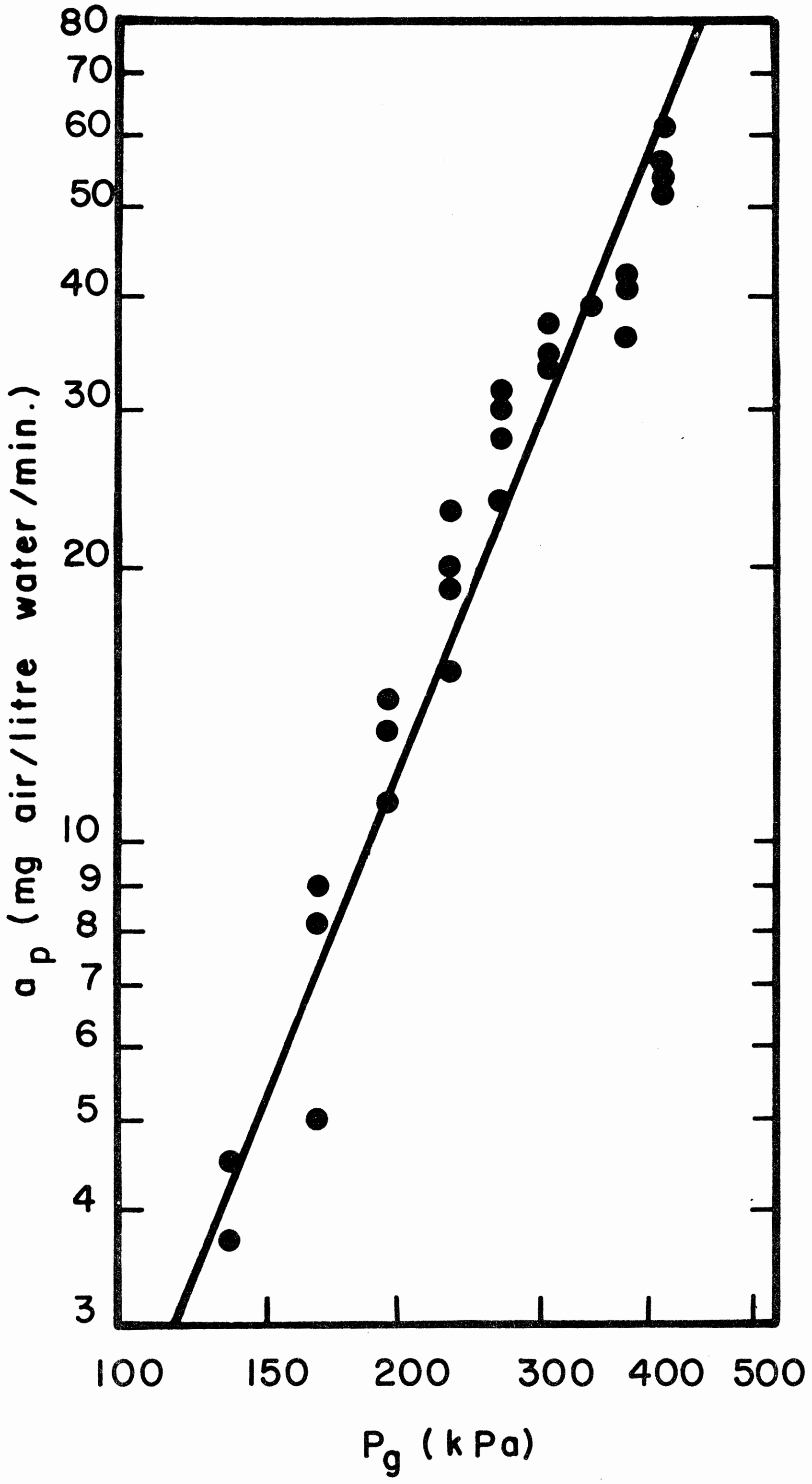


FIGURE 13(b): Relationship between a_p and P_g (Table 1).

arises from the difference between the pressure in the saturator and atmospheric pressure. At low pressure differences the degree of supersaturation is small and the rate of precipitation is relatively slow so that within the limits of the retention time in the monitoring unit, a significant fraction of the supersaturated air remains in solution. At high pressure differences the degree of supersaturation is greater, the rate of precipitation is higher and the fraction remaining in solution has less significance compared to the amount precipitated. Operating the saturator under higher pressures, therefore, would appear to give more efficient precipitation.

The mass of air precipitated in the flotation unit cannot be accurately correlated with that measured in the monitoring unit. In the flotation unit, the higher saturator feed rates used would induce a higher degree of turbulence at the needle valve which would enhance the rate of precipitation. In addition, the presence of particulate nuclei from the wastewater would also tend to enhance the rate of precipitation.

(*) On reflection, a superior method of measuring the precipitated air would be to have a clear water feed, similar to the wastewater feed, entering the monitoring unit with the saturator feed. This procedure would then take account of the dilution effect of the wastewater feed and may better simulate the conditions for precipitation in the flotation unit.

EFFECT OF pH AND ANIONS ON BUBBLE-PARTICLE ADHESION

There are three recognised mechanisms of bubble-particle attachment(7): (1) bubble capture within the surface irregularities of floc particles; (2) formation of flocs around bubbles and (3) adhesion of bubbles to particles.

A pre-requisite for all these mechanisms is destabilization. Mechanisms (1) and (2) require destabilization between particles and mechanism (3) requires a further "destabilization" between bubbles and particles(10,11). In addition to destabilization, mechanisms (1) and (2) require a certain degree of flocculation before or during flotation. Mechanism (3), in principle, requires no flocculation.

The objective of the experiments in this section was to identify which of these mechanisms dominate the flotation process.

Wastewaters may be classified into two general categories: those with natural flocculating propensity (for example activated sludge) and those to which metal coagulants must be added to induce destabilization and promote flocculation. A greater degree of control may be exercised over the optimum conditions for mechanisms (1), (2) and (3) by selecting a wastewater to which metal coagulants are added. Control is achieved by varying the coagulant type, dosage and pH since these parameters influence destabilization.

The wastewater used throughout the experimental work was algal laden oxidation pond water. Destabilization of the algae (predominantly Chlorella vulgaris) is not possible without coagulant addition. Although algal laden water was used in the investigations, the same principles should apply to any wastewater where metal coagulants are added to induce destabilization and promote flocculation.

Normally, colloid destabilization using ferric coagulants is largely independent of pH within a wide range of values. This is to be expected from the nature of the ferric complexes formed in aqueous solution(18). Ferric coagulants, therefore, are useful in determining whether or not bubble-particle destabilization is dependent on pH.

Initially, ferric chloride was used as a coagulant. The dosage and pH for optimum colloid destabilization were determined by means of the jar test, following the procedure suggested by Stumm and Morgan(18). Four batches of tests were carried out, each at a particular pH. The pH was maintained at the chosen values by the addition of hydrochloric acid (HCl) and/or sodium hydroxide ($NaOH$). The coagulant dosage for each batch was varied from 10 to 70 mg/l as Fe^{3+}. After rapid mixing, stirring and settling the supernatant turbidity of each sample was measured by means of a nephelometer. The standards used for the nephelometer were distilled water for zero and an arbitrary standard constructed from a ground perspex rod (introduced to a nephelometer tube) such that the raw algal laden waste water gave a reading of approximately 100. This corresponded to full scale deflection.

Turbidity-dosage curves are shown plotted in Figure 14(a). The optimum dosage (52 mg/l) was taken to be at point A on the lowest curve where higher dosages do not improve supernatant turbidity. From line A, Figure 14(a), the optimum pH was determined by plotting turbidity versus pH at the optimum dosage. Curve A in Figure 15 verifies that for ferric salts, colloid destabilization occurs over a wide range of pH values, from 5,0 to 7,0.

At the optimum coagulant dosage of 52 mg/l as Fe^{3+} and in the pH range, 5,0 to 7,0, the supernatant had a turbidity of 3,0 nephelometer divisions and appeared clear and "turbid-free".

To determine whether the optima found above for colloid destabilization coincided with the optima for bubble-particle attachment, tests using ferric chloride were then carried out using the flotation unit :

Before discharge into the flotation unit, efficient mixing of the coagulant into the wastewater stream was ensured by installing rapid mixing units* designed

(* See Figure 9(a)).

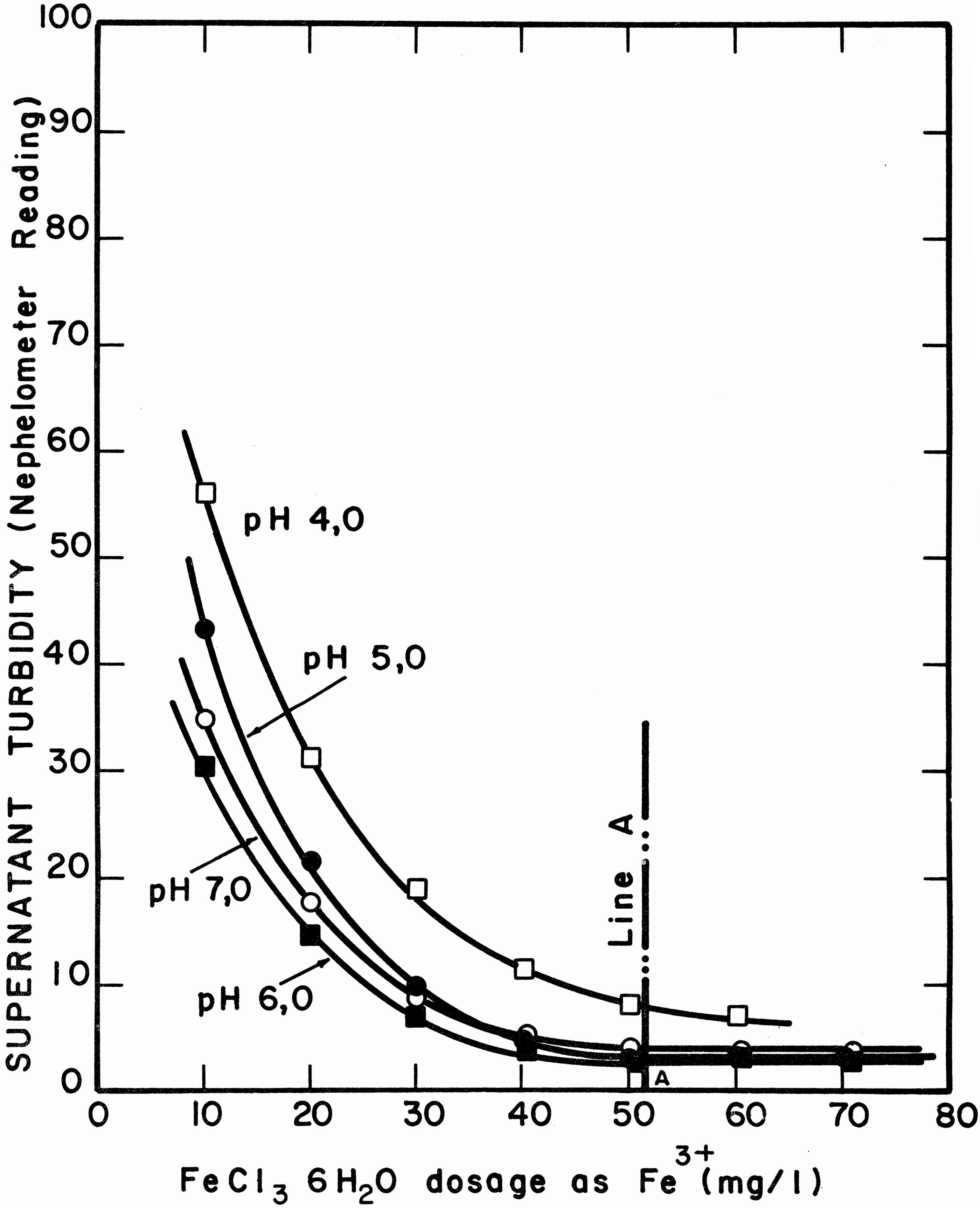


FIGURE 14(a): Turbidity-Dosage Curves Using Ferric Chloride in Jar Test (Table 2).

in accordance with the principles set out by Vrale and Jorden(19). (See Appendix II). A high degree of flocculation was induced by means of the flocculation tank shown in Figure 9. Thus the variable parameters were the same as in the jar tests, that is, the only variable parameters were coagulant dosage and pH.

The effluent quality from the flotation unit was, as before, based on turbidity using a nephelometer. A source of error was introduced by virtue of the air bubbles present within the liquid in the nephelometer tube. This was overcome by taking a 500 ml sample of the effluent from the flotation unit and violently stirring with a magnetic stirrer until the sample, when tested in the nephelometer, gave a reading which did not change with time. If such a change did occur it indicated that air bubbles were still present in the sample.

Figure 14(b) and curve B of Figure 15 show that the optimum coagulant dosage for flotation was 50 mg/l as Fe^{3+} which is not significantly different from the value found in the jar test. Unlike the results from the jar tests, however, there was now a distinct optimum pH value for flotation at pH 5,65. That is, bubble-particle attachment was significantly influenced by pH. The effluent at these optima had a turbidity of 6,0 nephelometer divisions which was only slightly higher than the optimum supernatant turbidity in the jar test.

Figure 15 illustrates the effect of pH on colloid destabilization (curve A) and bubble-particle attachment (curve B). The curves indicate that mechanism (3), that is, adhesion, is the prime mechanism promoting bubble-particle attachment; bubble capture in the surface irregularities of flocs and the formation of flocs around bubbles have lesser importance.

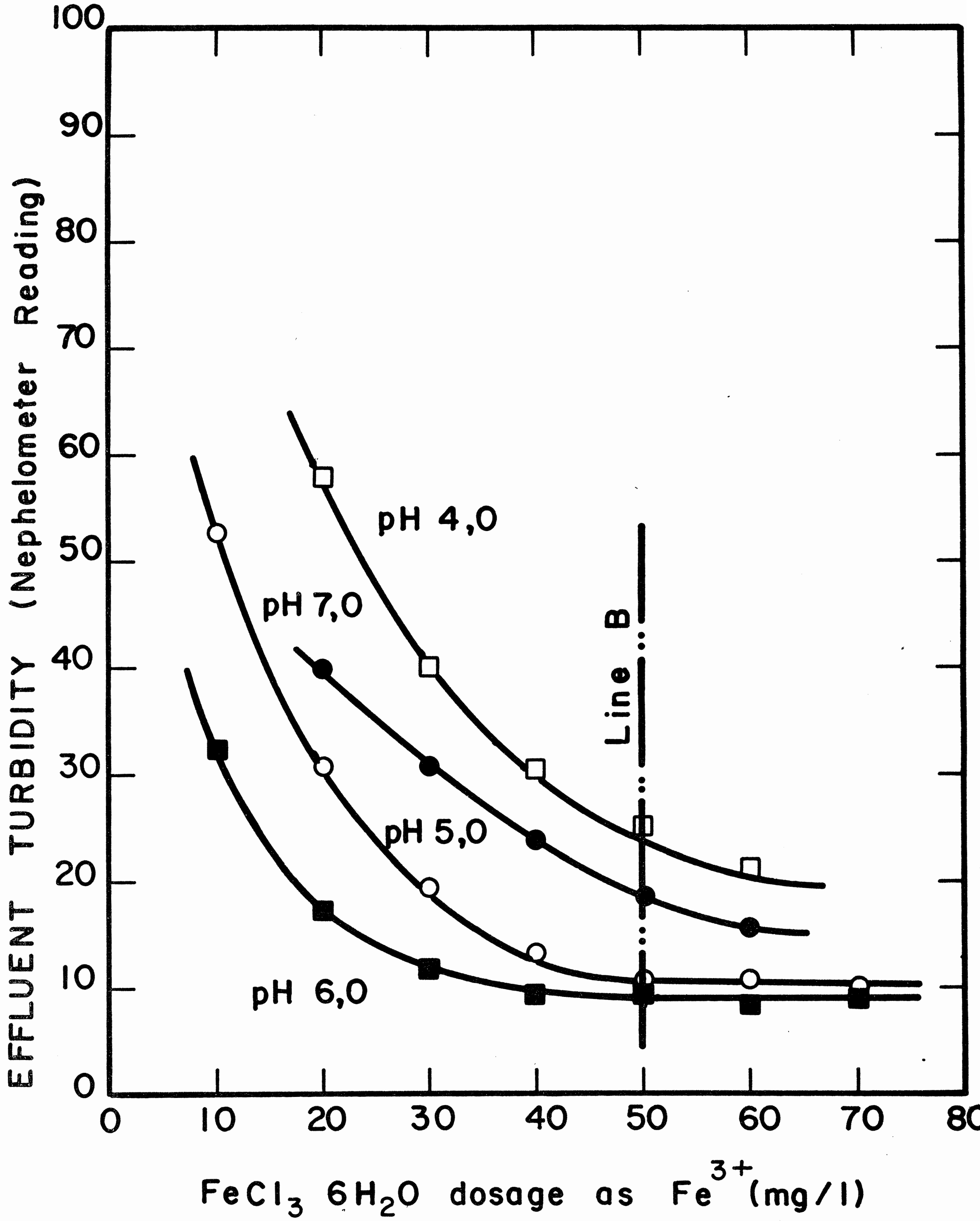


FIGURE 14(b): Turbidity-Dosage Curves Using Ferric Chloride in Flotation Unit (Table 3).

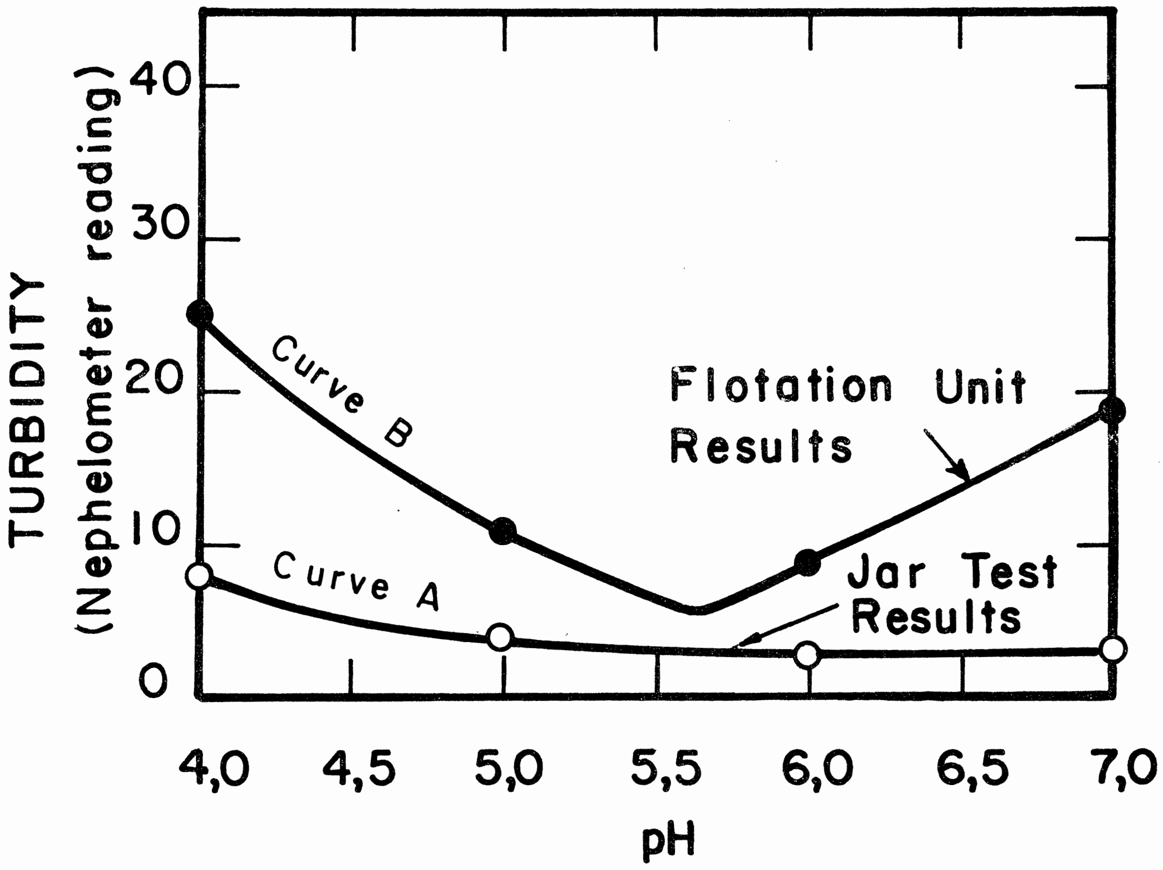


FIGURE 15: Determination of Optimum pH Using Optimum Dosage Ferric Chloride in Jar Test and Flotation Unit (values taken from Figures 14(a) and 14(b)).

These latter two mechanisms rely solely on floc formation and hence depend essentially on colloid destabilization. Since the optima for colloid destabilization are determined by the jar test, then for these two mechanisms to be prominent during flotation, the 'spread' of pH values for optimal conditions in the jar test should necessarily be duplicated during flotation. This was not the case. The distinct value of the optimum pH for flotation implies a further pH dependence for bubble-particle attachment, over and above that for colloid destabilization.

To verify the above, the jar and flotation tests were repeated using ferric sulphate. The results are shown in Figures 16(a), 16(b) and 17. From curve C, Figure 17, it is again evident that for colloid destabilization there is a wide range of pH values over which the supernatant in the jar tests has an 'optimum' turbidity (of 1,0 nephelometer divisions at the optimum dosage of 50 mg/l as Fe^{3+}). Furthermore, for flotation there is again a distinct pH value of 5,8 for optimum bubble-particle attachment, as is evident in curve D of Figure 17. This pH value is not significantly different from that for ferric chloride at pH 5,65. However, the turbidity of the effluent from the flotation unit at optimum conditions was now 11,0 compared to 6,0 when ferric chloride was used. Visually the effluent had a murky appearance.

The inferior effluent quality during flotation when using ferric sulphate suggested that with this particular wastewater, sulphate anions possibly impaired bubble-particle adhesion. To test this hypothesis both the jar and flotation tests were repeated using aluminium sulphate. The results are shown in Figures 18(a), 18(b) and 19.

In the jar test the supernatant was very clear with a turbidity of 3,0 nephelometer divisions. With flotation however, the effluent at optimum conditions, had a 'murky' appearance with a turbidity of 18,5 nephelometer

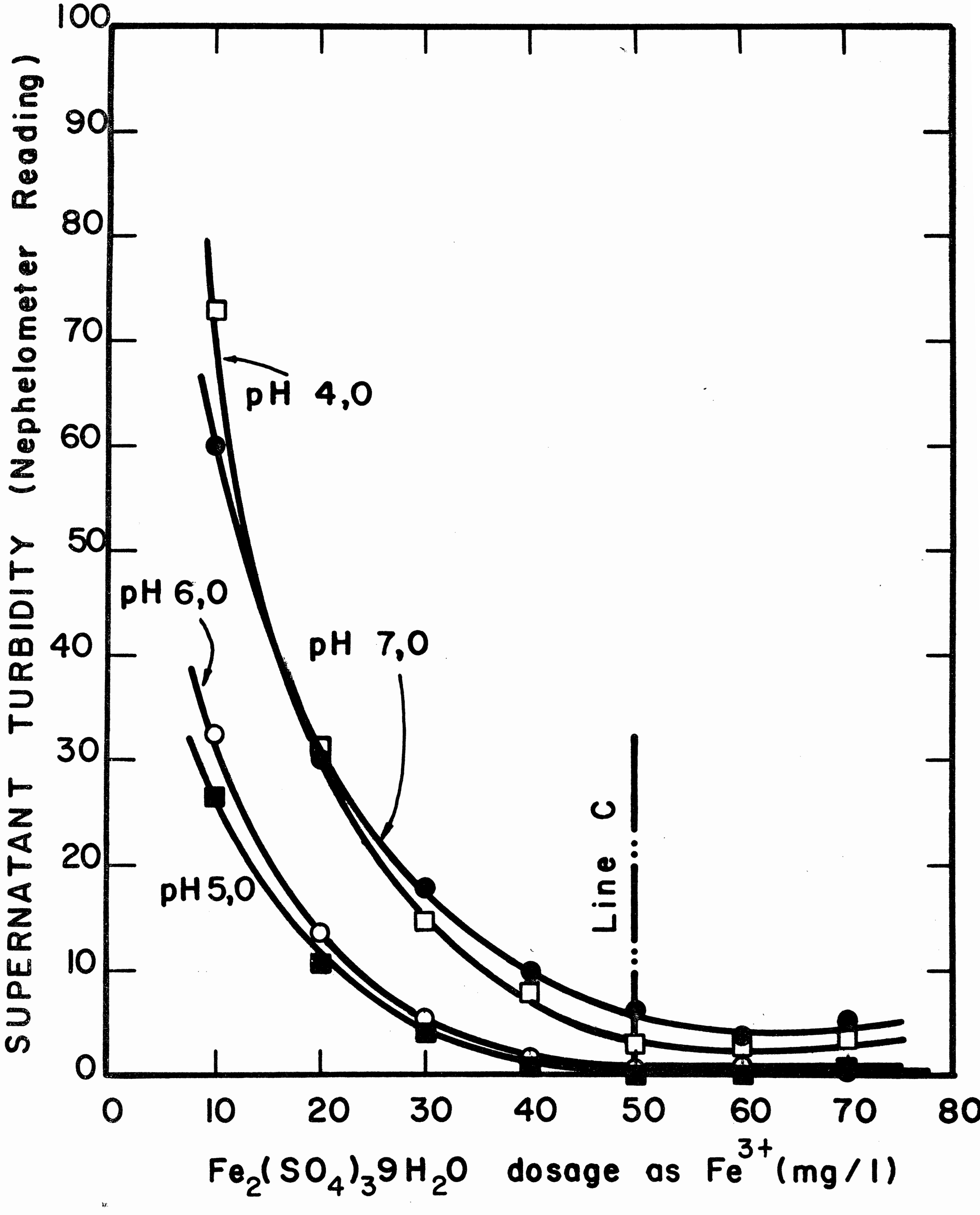


FIGURE 16(a): Turbidity - Dosage Curves Using Ferric Sulphate in Jar Test (Table 4).

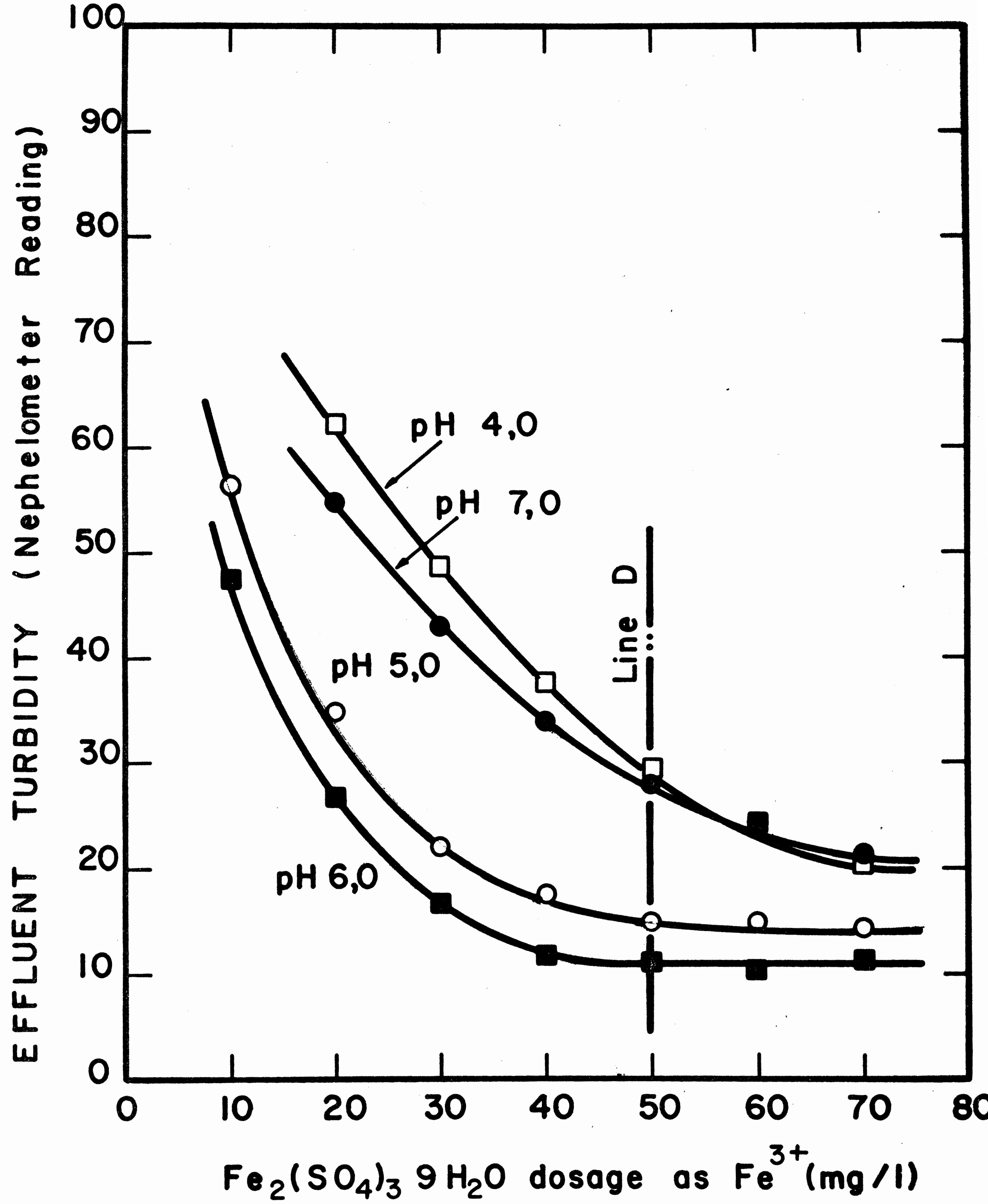


FIGURE 16(b): Turbidity - Dosage Curves Using Ferric Sulphate in Flotation Unit (Table 5).

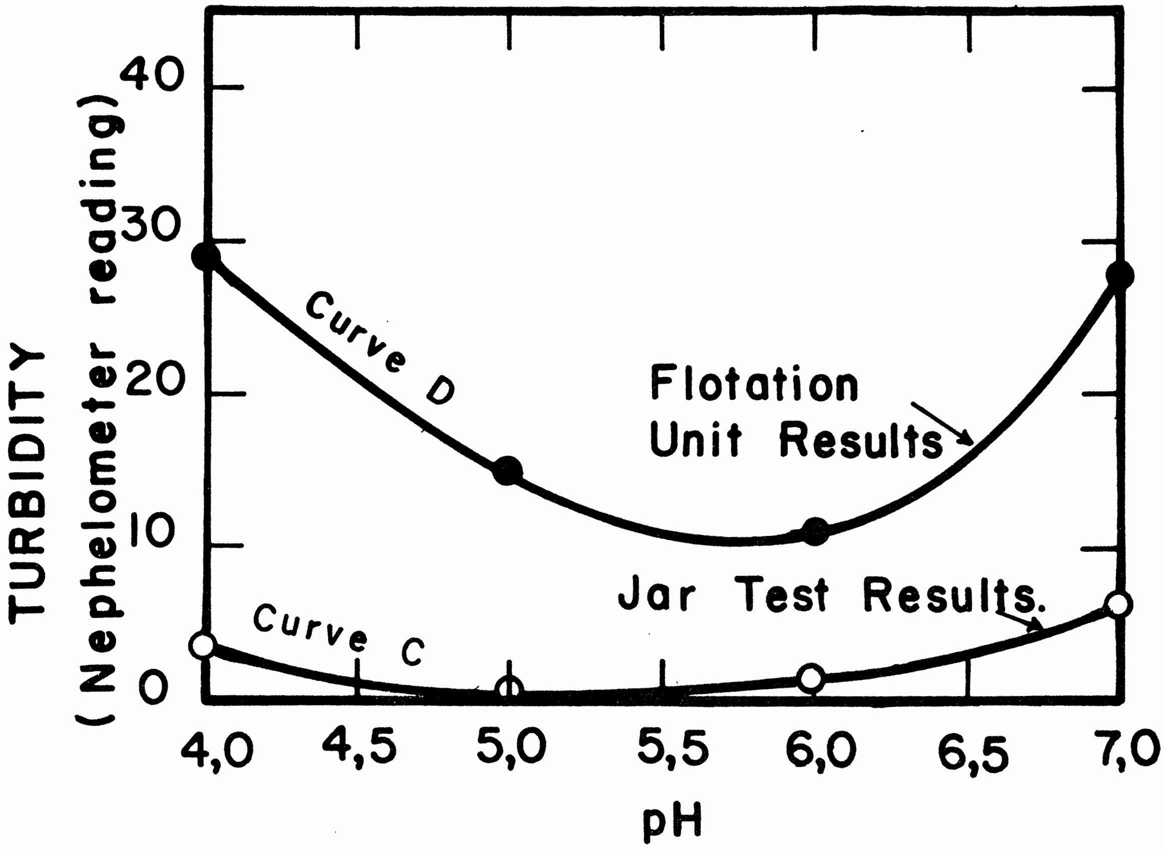


FIGURE 17: Determination of Optimum pH Using Optimum Dosage Ferric Sulphate in Jar Test and Flotation Unit (values taken from Figures 16(a) and 16(b)).

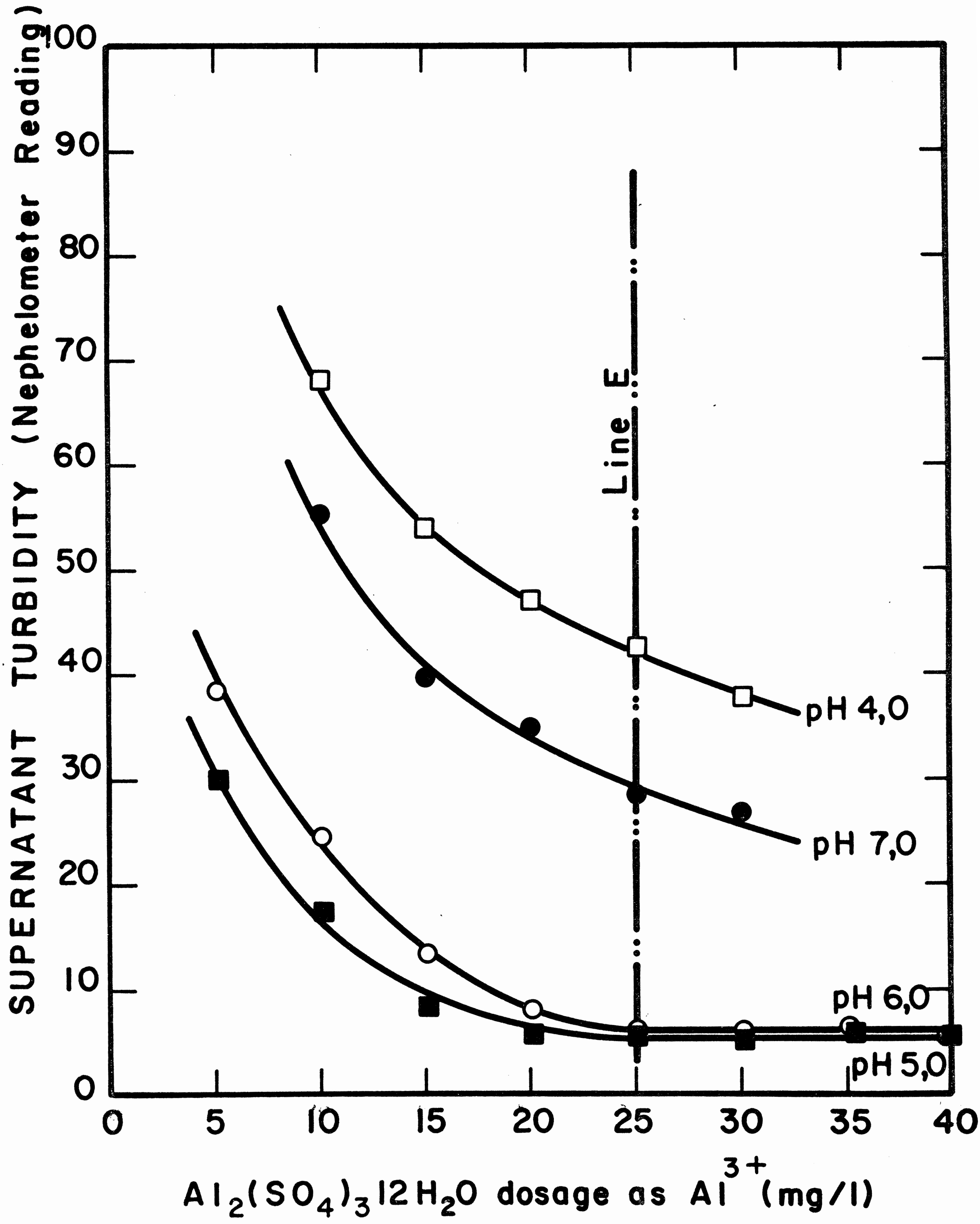


FIGURE 18(a): Turbidity - Dosage Curves Using Aluminium Sulphate in Jar Test (Table 6).

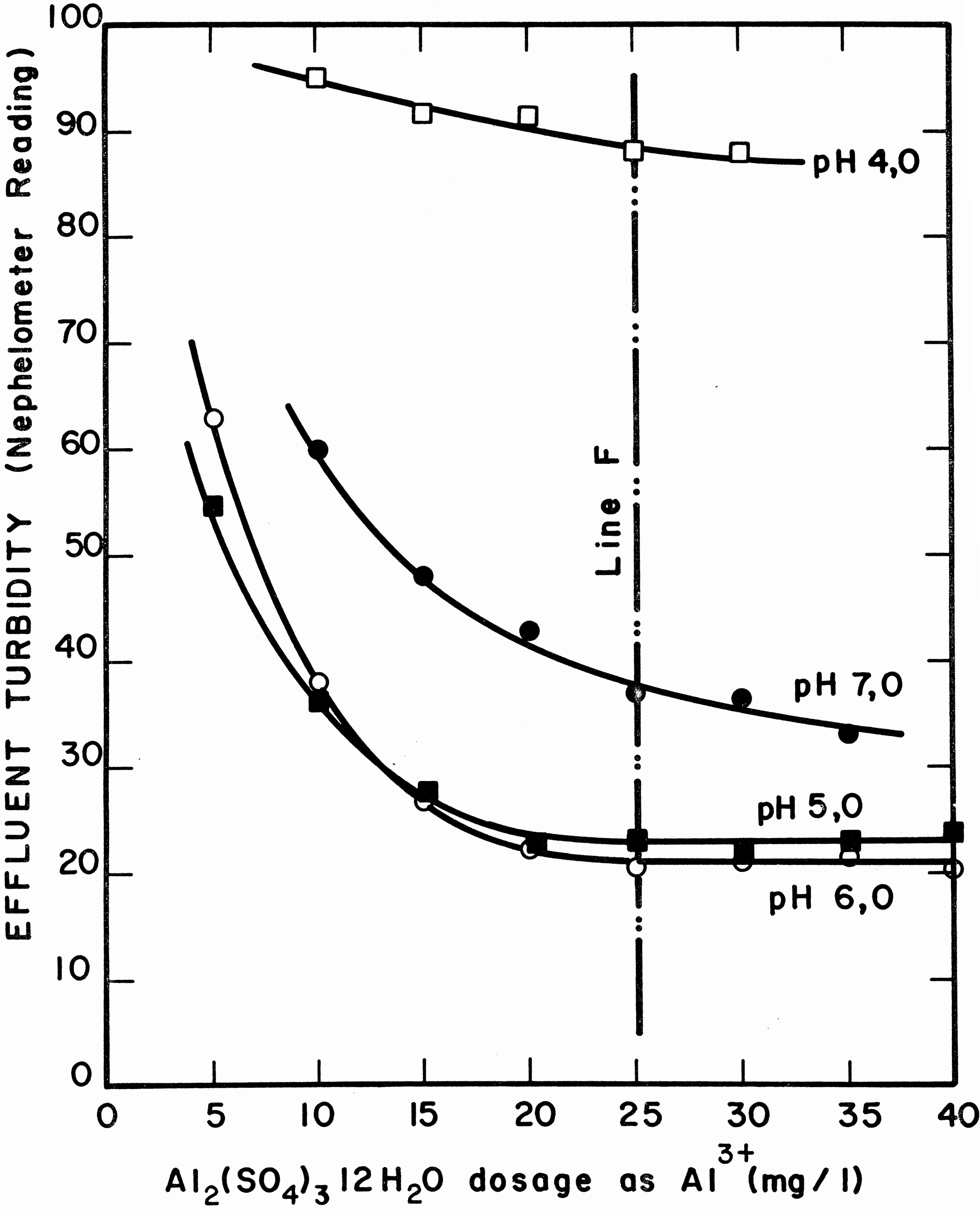


FIGURE 18(b): Turbidity - Dosage Curves Using Aluminium Sulphate in Flotation Unit (Table 7).

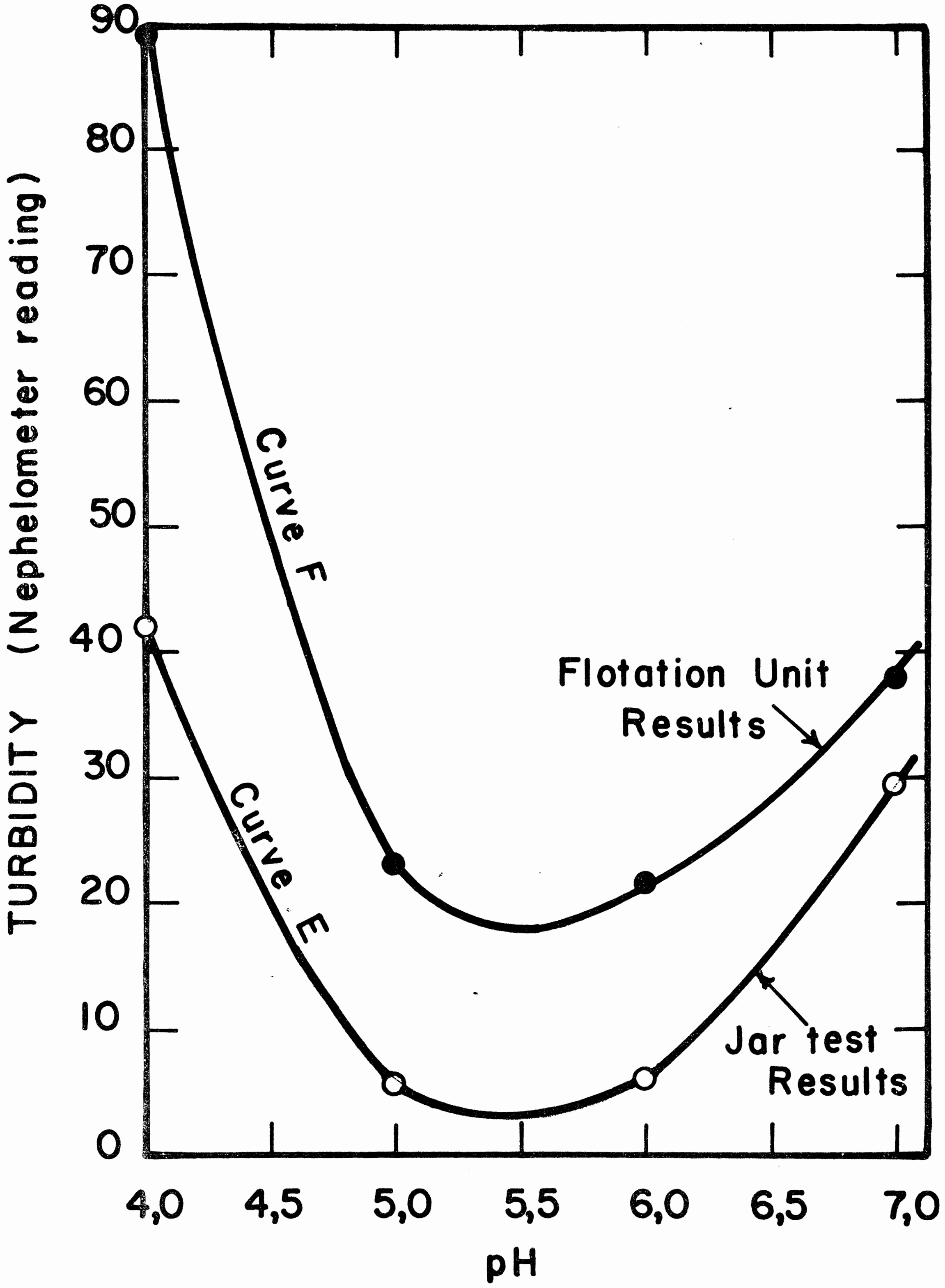


FIGURE 19: Determination of Optimum pH Using Optimum Dosage Aluminium Sulphate in Jar Test and Flotation Unit (values taken from Figures 18(a) and 18(b)).

divisions. In both tests the optimum dosage was 25 mg/l as Al^{3+} at an optimum pH of 5,5. It seems evident, therefore, that sulphate anions were inhibiting bubble-particle adhesion.

No satisfactory explanation can be advanced for the detrimental effect of sulphate anions. It is possible that, although more heavily hydrated than chloride ions, sulphate ions tend to collect more readily at solid-liquid and air-liquid interfaces due to specific chemical processes(9). By this means, the negative charge carried by $SO_4^{=}$ could impair bubble-particle adhesion.

The influence of pH on flotation is not so apparant when aluminium sulphate is used (see Figure 19). From the nature of the aluminium complexes formed in aqueous solution(18), the optimum pH for colloid destabilization has, in any case, a distinct value.

As the optima for flotation may differ from those for colloid destabilization, the jar test is not a reliable method for determining the optimum conditions of coagulant type, dosage and pH for flotation.

In the subsequent investigations reported, ferric chloride was the sole coagulant used. The dosage was maintained at the optimum value of 52 mg/l as Fe^{3+} and the pH at 5,65. These optimal values were checked periodically by running a set of flotation tests as described above.

EFFECT OF REACTION TIME AND DEGREE OF PRE-FLOCCULATION ON FLOTATION

The flotation experiments described in the preceeding section were carried out with pre-flocculation. That is, the destabilized wastewater stream (after coagulant addition) was passed through a flocculation tank (see Figure 9) so that a high degree of flocculation took place prior to air precipitation and adhesion. The minimum turbidity achieved at optimum conditions was 6,0 nephelometer divisions. It was considered that the particles yielding this reading consisted of ones not amenable to either destabilization or flotation - or they consisted of colloidal metal-hydroxide precipitates.

To determine whether pre-flocculation was necessary prior to flotation, the flocculation tank was removed and the coagulant introduced to the wastewater feed line immediately prior to the junction with the saturator feed. The turbidity of the effluent from the flotation unit at optimum conditions was then 9,5 nephelometer divisions, as opposed to 6,0 with the flocculation tank installed.

To ascertain whether this deterioration in effluent quality was possibly due to an insufficient degree of destabilization, the coagulant was introduced at various points further back on the wastewater feed line to give a range of "reaction times" from 1 sec to 25 sec before coming into contact with the saturator feed.

Figure 20 shows that pre-flocculation is unnecessary. Reaction times greater than 14 sec yielded an effluent of the same quality as that with the flocculation tank installed.

Subsequently, all experiments were conducted without the flocculation tank. At the maximum wastewater flow, the reaction time of the coagulant prior to flotation was 14 sec.

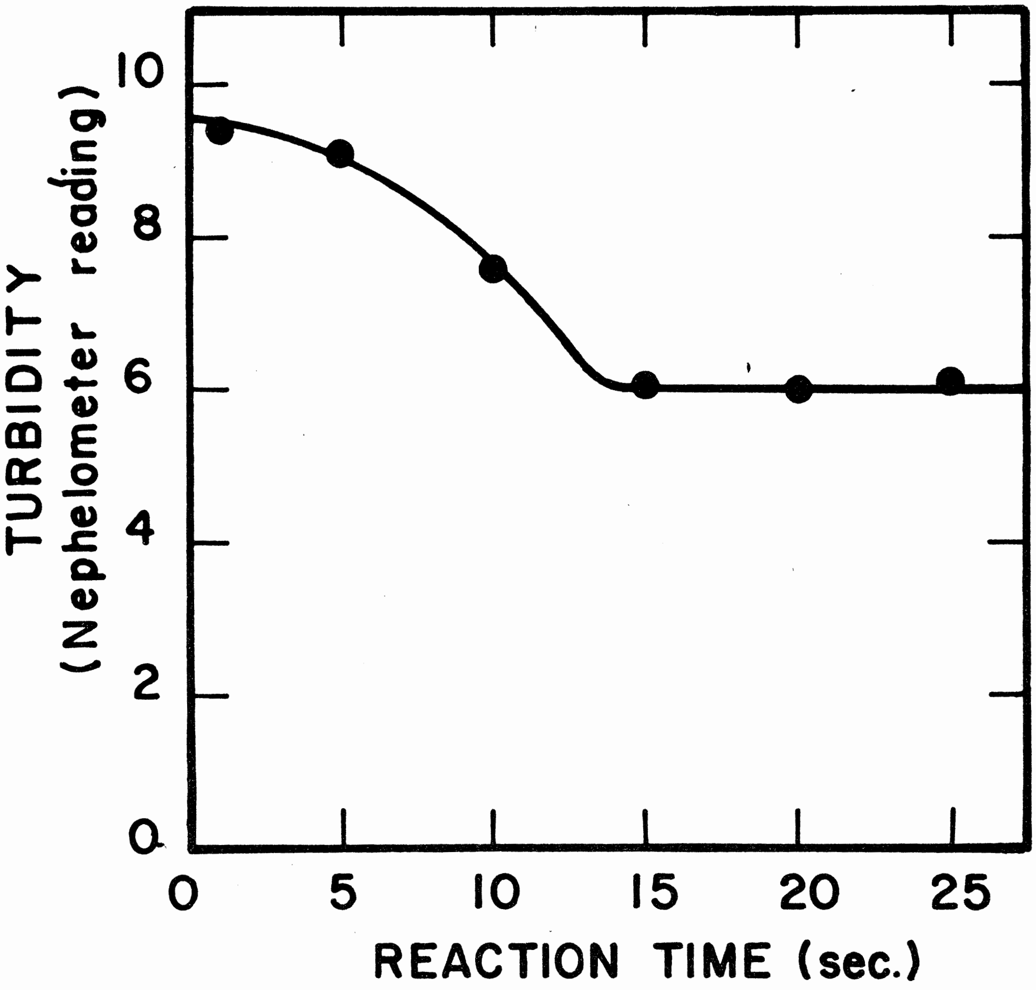


FIGURE 20: Effect of Reaction Time from Point of Coagulant Addition to Air Precipitation (Table 8).

EFFECT OF CROSS SECTIONAL AREA OF UPPER CHAMBER ON THE LIMITING DOWNFLOW RATE

Downflow rate is defined to be equal to the total hydraulic flow rate into the flotation unit divided by the cross-sectional area at the outlet of the unit (see insert, Figure 21). It is analogous to the "overflow rate" as defined in settling tank theory and has the same form of units. The unit used here for the downflow rate is $cm^3/cm^2/min$ which is, essentially, a unit of velocity.

The limiting downflow rate, v_L, is that value of the downflow rate where, for any given set of conditions, solids just commence to be drawn down with the effluent from the unit. Since from its definition, v_L has units of velocity (flow rate ÷ area), then it is dependent on the velocity of rise of the bubble-particle agglomerates. v_L is equivalent to the velocity of rise of the slowest moving bubble-particle agglomerate.

The velocity of rise of bubble-particle agglomerates is dependent on (1) the applied air/solids ratio, a_s, and (2) the relative spacing of the agglomerates.

The air/solids ratio, a_s = mass of air precipitated per unit of mass of solids. It has units mg/mg.

The relative spacing of the agglomerates determines the extent to which their velocity of rise is hindered by the downward interstitial velocity of the displaced water. This hindered effect, in turn, depends on the rate at which solids are introduced to the unit.

The latter is expressed in terms of the parameter "solids loading rate", Q_s. It is defined as Q_s = mass of wastewater solids introduced per unit time per unit area of the upper chamber and has units $mg/cm^2/min$. Q_s is dependent on (1) the wastewater solids concentration (2) the area of the upper chamber and (3) the wastewater flow rate. Increasing the wastewater flow rate into the unit, therefore, achieves both an increase in the downflow rate (=Q/A) and an increase in the solids loading rate.

The objective of these experiments was to determine whether, by increasing the solids loading rate into the unit, a "bottle-neck" situation could arise whereby solids were introduced to the unit at a rate greater than they could rise (and escape) in the upper chamber.

The experimental procedure was as follows:

A constant arbitrary air/solids ratio (of 0,0063 mg/mg) was applied throughout the experiments. This was done by maintaining a constant pressure in the saturator and a constant (arbitrary) ratio of wastewater feed to saturator feed (of 8:1).

The wastewater and saturator feeds into the unit (maintained at a constant ratio) were incrementally increased. This caused incremental increases in the downflow rate. At each downflow rate applied, the effluent quality was determined on a comparative basis with the nephelometer.

> When using the nephelometer it was found necessary to use the following testing technique : A 500 ml sample was taken of the effluent and acid was added to bring the pH value down to approximately 2,0 in order to restablize the particles. The sample was stirred violently with a magnetic stirrer to break down the bubble-particle agglomerates and liberate the air bubbles. If this procedure was not carried out it was found that not only did the flocculating particles give unreliable nephelometer readings but they also rose to the top of the nephelometer tube.

By plotting downflow rate versus effluent turbidity, the limiting downflow rate at which solids just commenced to be drawn down with the effluent was determined (see Figure 21).

The flotation unit was constructed such that the upper chamber was detachable. It was therefore possible to repeat the above experiments with upper chambers of area 225cm^2, 100cm^2 and 25cm^2 installed (see Figure 21). By this means a comparison could be made between the effluent quality at each value of the downflow rate for a range of solids loading rates. (Note: for a given downflow rate - which corresponds to a given wastewater flow rate - decreases in the area of the upper chamber effect corresponding increases in the solids loading rate).

The results of the experiments show an unexpected trend: rather than a detrimental (bottle-neck) effect when increasing the solids loading rate (by decreasing the area of the upper chamber), it appears that an advantageous effect is produced.

Figure 21 shows the following:

(1) The maximum permissible hydraulic flow rate into the unit beyond which solids are drawn down with the effluent, (determined by the limiting downflow rate, v_L) appears to be greater when the area of the upper chamber is decreased.

(2) For flow rates greater than those at the limiting downflow rates, the fraction of solids drawn down with the effluent appears to be less when the area of the upper chamber is decreased.

A possible reason for the above behaviour is as follows: By decreasing the area of the upper chamber for a given downflow rate (that is, a given wastewater feed rate), the solids loading rate is correspondingly increased.

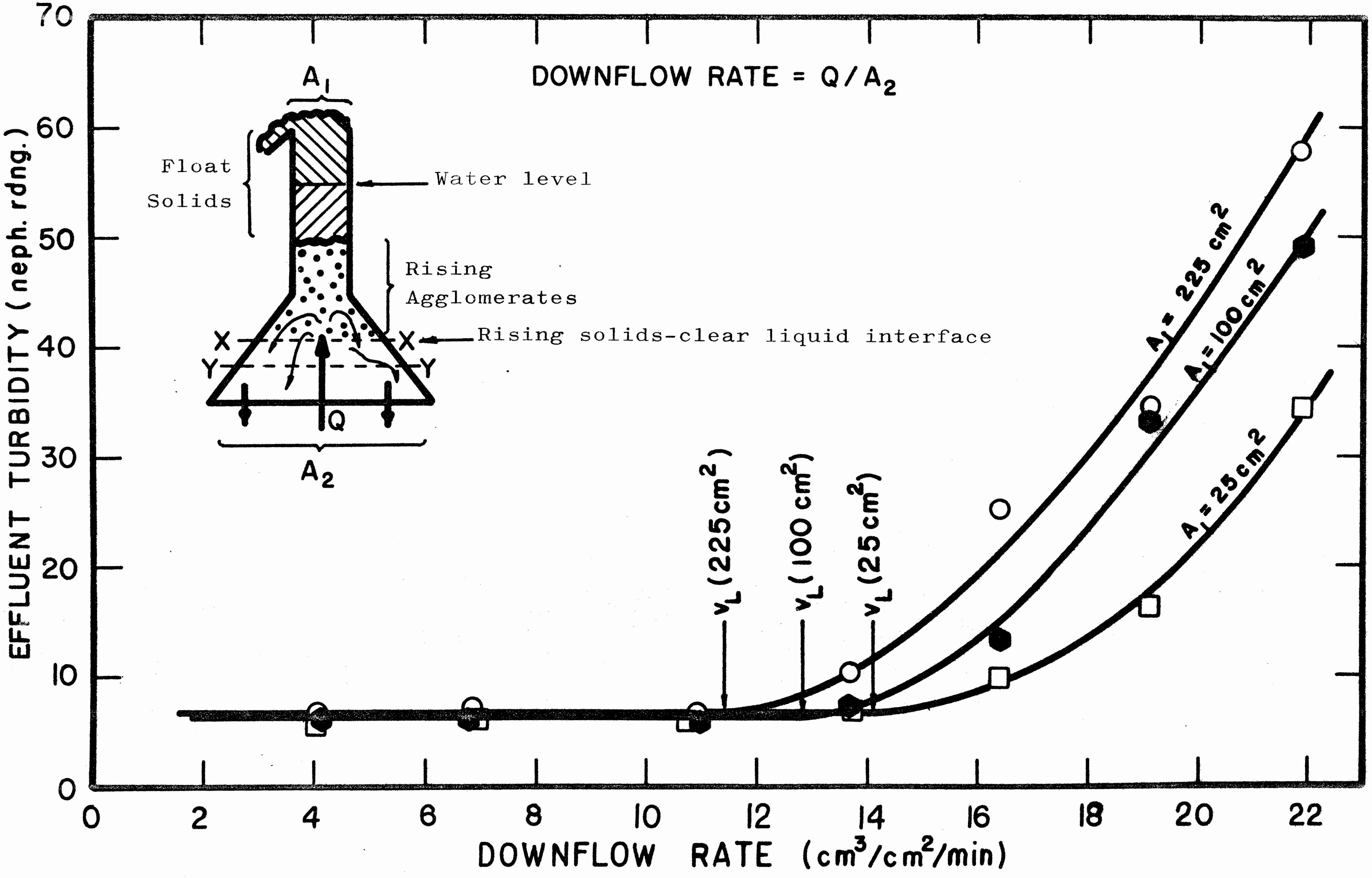


FIGURE 21: Effect of Cross Sectional Area (A_1) of Upper Chamber on Limiting Downflow Rate, v_L (Table 9).

This causes a "confinement" of the rising agglomerates in the upper chamber which gives rise to greater contact opportunity and induces progressively larger agglomerates to form. A stage is reached where the agglomerates are confined to such an extent that they mutually touch and become, in effect, a buoyant mass. This was substantiated by observing the interface between rising agglomerates and "clarified" liquid at high solid loading rates (with high wastewater feed rates and the upper chamber of area 25cm^2 installed). This interface was seen to oscillate in position - beginning a cycle at level XX, say, (see Figure 21) and falling slowly such as would occur if solids "build-up" were taking place. However, on reaching a level YY, say, the agglomerate mass "surged" upwards until level XX was again reached. At each upward surge, solids were discharged over the float weir at the top of the upper chamber. The values of the solids loading rate, Q_s, beyond which this behaviour was noted was approximately 200 mg/cm^2/min.

Design of flotation tanks for <u>clarification</u> will be based on an economic assessment of the area (A_2 of Figure 21) requirements for different limiting downflow rates at corresponding air/solid ratios (see next section). By decreasing the area (A_1) of the upper chamber it appears that the limiting downflow rate for any particular air/solids ratio may be increased. It is therefore possible that for a particular wastewater flow rate, the area, A_2, of the flotation unit may be significantly decreased by decreasing the area of the upper chamber.

The above ignores the design considerations pertaining to thickening (see later) where, by decreasing the area of the upper chamber, the resultant increase in solids loading rate will give rise to a decrease in float solids concentration. This may be overcome by including a further upper chamber whose area is determined on the basis of a desired float solids concentration. However, in some cases, it is not required to have a high float solids concentration. For example, if flotation is used with the activated sludge process, where the "activated sludge" is separated form the stabilized liquid fraction, then a fraction of the float will be returned to the reactor. In this case the float solids concentration can be low, say 2%. This concentration will still be at least twice that which can be obtained with sedimentation.

EFFECT OF AIR/SOLIDS RATIO ON LIMITING DOWNFLOW RATE

The limiting downflow rate, v_L, is dependent on the rate of rise of the bubble particle agglomerates (see previous section). The rate of rise of bubble-particle agglomerates is dependent primarily on the air/solids ratio, a_s, where a_s = mass of air precipitated per unit mass of solids. (mg/mg).

It would appear, therefore, that the parameters a_s and v_L are related. The objective of the experiments in this section was to determine if there is a relationship between a_s and v_L and the form of such a relationship.

The experimental procedure was as follows:

A value of a_s was fixed and the downflow rate was incrementally increased by increasing the waste water and saturator feed rates. These feed rates were maintained at an arbitrary ratio of 4:1 respectively so that the dilution effect of the saturator feed was always constant and, further, for any particular pressure in the saturator, the air/solids ratio was independent of the flow rate into the unit.

At each downflow rate, the effluent solids were determined and a curve was plotted of downflow rate versus effluent solids. This procedure was repeated for a number of air/solid ratios.

The air/solids ratio was varied by changing the pressure in the saturator from 140 kPa to 420 kPa in 35 kPa increments. a_s was determined for each test by means of the "monitoring" unit shown in Figure 12. It was based on a wastewater solids concentration of 450 mg/l which was found to remain approximately constant for the duration of these investigations. This was determined periodically be centrifuging samples of the wastewater plus coagulant, drying the residue and weighing.

For each test, a period of two times the retention time was allowed to elapse before a sample was taken. This period was found to be sufficient to establish stable conditions.

Solids present in the effluent were determined by two methods : (1) by the use of the nephelometer and (2) by vacuum filtration of effluent samples with high retentive filter papers (Whatmans' No. 42) and weighing the residue.

The nephelometer was used primarily because of its simplicity and swiftness and further, because absolute values of effluent solids concentration were not necessary for the strict objective of these tests. The objective was merely to detect at which stage solids commenced to be drawn down with the effluent under the imposed conditions of each test. (The testing technique described in the preceeding section was also applied when using the nephelometer for these tests).

The determination of effluent solids by vacuum filtration was carried out in addition to the nephelometer method because of the following reasons:

(a) Although quantitative determinations of effluent solids were not strictly necessary, a knowledge of their orders of magnitude under different conditions was nevertheless desirable.

(b) Due to damage, the nephelometer was found to have reduced sensitivity for these tests and it was necessary to construct a new arbitrary standard for full scale deflection as before. Repetitions of the tests with absolute solids determinations afforded a check on the results gained from using the nephelometer. Because of this reduced sensitivity an effluent quality with a nephelometer reading of 2,0 (at optimum conditions) is equivalent to a nephelometer reading of 6,0 as quoted in the predeeding sections. The corresponding suspended solids concentration at these (optimal) nephelometer readings was approximately 2 mg/l. This constituted approximately 99,5% solids removal, the wastewater solids concentration being approximately 450 mg/l.

Figures 22(a) and 22(b) show, for each air/solids ratio, the effluent solids present at each value of the downflow rate. In Figure 22(a) the effluent solids are expressed relatively in terms of turbidity (nephelometer reading). In Figure 22(b) the effluent solids concentrations are expressed in terms of mg/l. In both Figures 22(a) and 22(b) the limiting downflow rate, for each air/solids ratio, is taken to be that value of the downflow rate where further increases serve to increase the effluent solids.

Figure 23 shows that if the limiting downflow rates, v_L, taken from Figures 22(a) and 22(b) are plotted against their corresponding air/solid ratios, a_s, with both axes logarithmic, a straight line relationship results. The relationship is of the form

$$v_L = K_2 \, (a_s)^{K_1} \quad \text{———————} \quad (2)$$

where K_1 and K_2 are constants with empirical values $K_1 = 0,77$ and $K_2 = 550$.

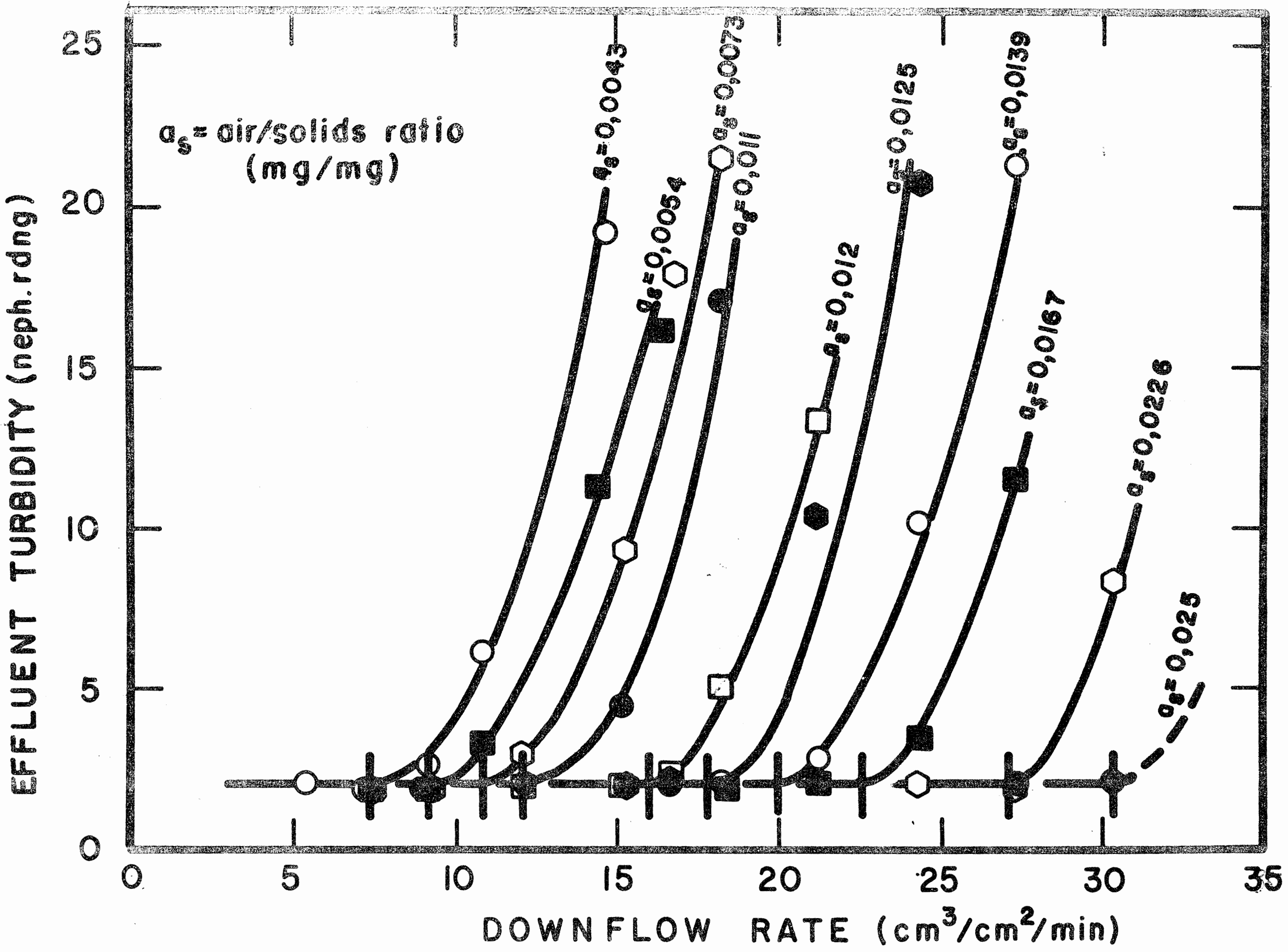


FIGURE 22(a): Determination of Limiting Downflow Rate at Different Air/Solid Ratios (Effluent solids determined by nephelometer) (Table 10).

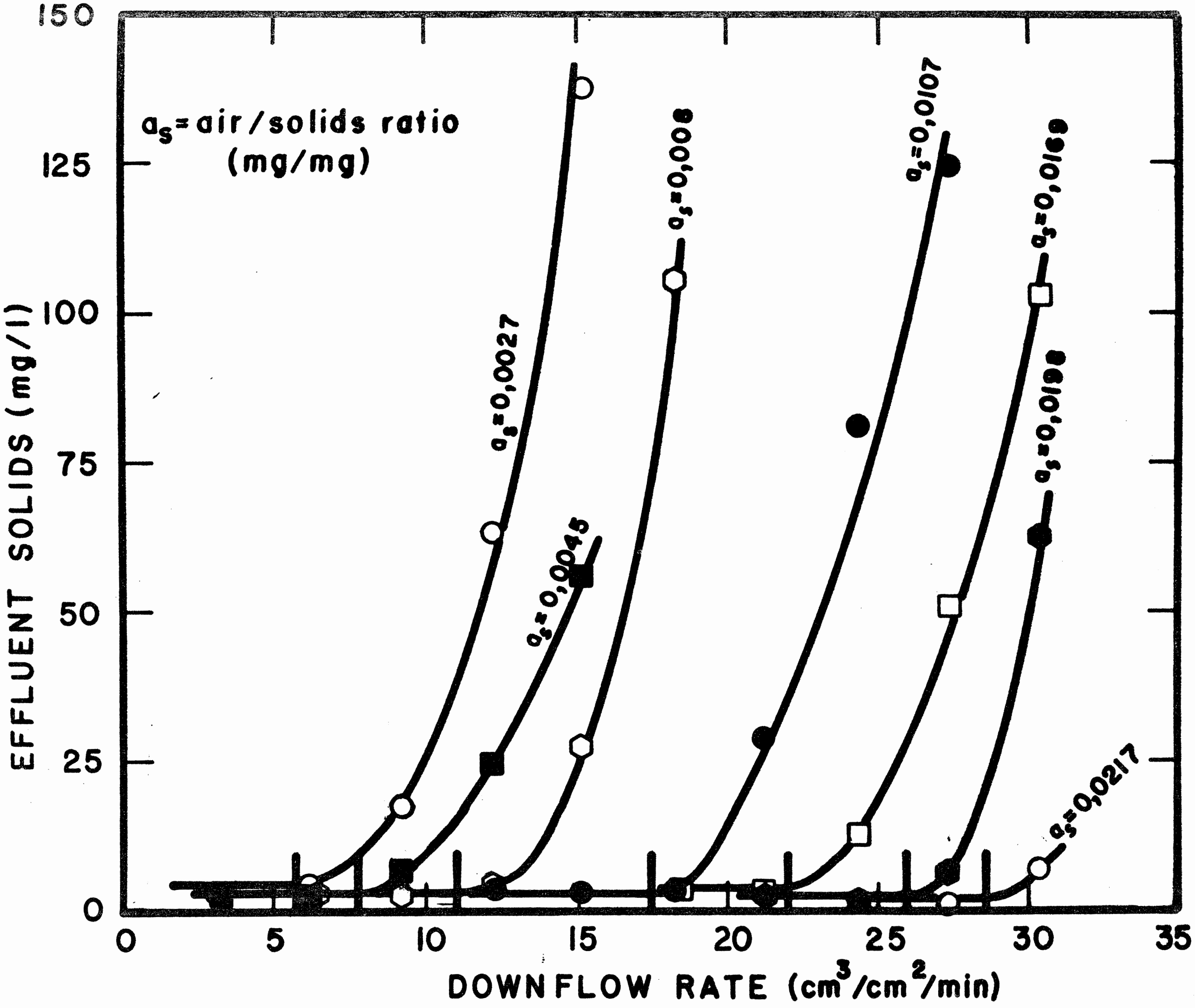


FIGURE 22(b): Determination of Limiting Downflow Rate at Different Air/Solid Ratios (Effluent solids determined by vacuum filtration and drying) (Table 11).

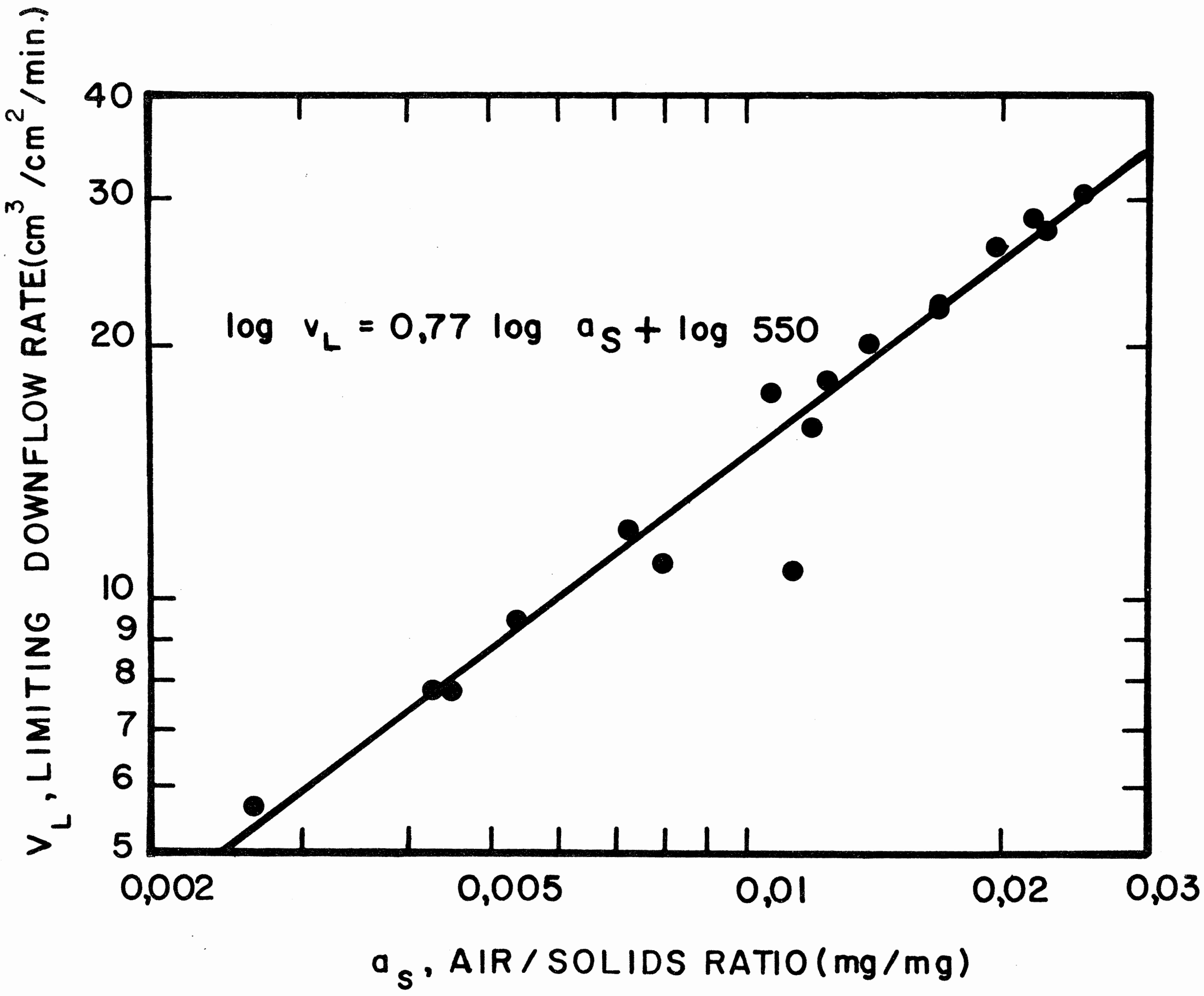


Figure 23: Relationship Between Air/Solids Ratio and Limiting Downflow Rate (values of v_L taken from Figures 22(a) and 22(b)).

Equation (2) cannot be strictly correct for it does not reflect the state where there is a zero velocity of rise ($v_L = 0$) at some finite value of a_s, that is, when the bubble-particle agglomerates are <u>just</u> buoyant.

A more realistic expression, therefore, would be to write

$$(v_L + v_S) = K_2 \, (a_s)^{K_1} \quad \text{———} \quad (3)$$

where v_S = the settling velocity of an average particle at $a_s = 0$. (i.e. at $a_s = 0$, $v_L = -v_S$ where the negative sign implies the "overflow rate" used in sedimentation theory).

From equation 3, when $v_L = 0$, the finite value of a_s is given by

$$a_s = \left(\frac{v_S}{K_2} \right)^{1/K_1} \quad \text{———} \quad (4)$$

However, under operating conditions, the value of v_S would be much smaller than v_L and equation (2) may be applied for design purposes within the general limits of a_s used in these experiments.

Figure 23 further shows that there is no significant difference between the results gained with the nephelometer and with vacuum filtration. This suggests that in tests such as these, where the <u>detection</u> of solids is the prime requisite, the use of a nephelometer or turbidimeter is warranted. The advantage of using such a measurement technique is its inherent simplicity and swiftness. The disadvantage is that the magnitude of the effluent solids is unknown.

Knowledge of the relationship between the air/solids ratio and the limiting downflow rate is necessary for design purposes. In order to prevent solids being carried down with the effluent at the maximum design flow, tanks of sufficiently large area must be provided. Such a design will necessarily compare the capital cost of the flotation tank at various air/solid ratios with the running cost of providing sufficient air to maintain these air/solid ratios. The relationship between a_s and v_L also provides a means for dealing with unexpectedly high wastewater flow rates - the air/solids ratio would be increased to a value where the corresponding limiting downflow rate, v_L, is equal to the downflow rate determined by the new wastewater flow rate.

PARAMETERS INFLUENCING FLOAT SOLIDS CONCENTRATION

Wastewater particles, after attachment to bubbles, rise and enter the upper chamber of the flotation unit until they reach the water level. The bubble-particle agglomerates commence accumulating and through their buoyancy cause the upper layers to lift beyond the water level. This continues until the agglomerates have accumulated to such a degree that the upper layers reach the float weir at the top and commence overflowing. These accumulated agglomerates are termed the "float" and their concentration, when they overflow, is termed the float solids concentration, C_F.

It was felt that C_F is a function of the three parameters Q_s, a_s and d_W, i.e. $C_F = f\ (Q_s,\ a_s,\ d_W)$.

where Q_s = solids loading rate = mass of solids introduced per unit area of upper chamber per unit time (mg/cm^2/min).

a_s = air/solids ratio (mg/mg).

d_W = depth of float above water level (cm).

To obtain data on the influence of these three parameters on C_F, two parameters were fixed and the third varied. In this case, d_W and a_s were fixed and Q_s varied. Then, for the same d_W, a_s was given different values and Q_s varied until several a_s, Q_s combinations were tested for this d_W. The whole sequence was then repeated for a new value of d_W.

> Because approximately 99,5% of the wastewater solids were captured during flotation, the solids loading rate was based on the wastewater solids concentration entering the unit. This was found to remain approximately constant at 450 mg/l. The solids loading rate was varied by adjusting the wastewater flow rate entering the unit. Solids carried down with the effluent was prevented by operating within the limiting downflow rate for a particular air/solids ratio.

For $d_W = 5$cm the plots of C_F versus Q_s, for each value of a_s, are shown in Figures 24(a) to 24(i). They show that whatever the value of a_s, as Q_s increases C_F decreases.

By abstracting data from Figures 24(a) to 24(i), C_F versus a_s may be plotted for different Q_s values, as shown in Figure 25. These plots show an unexpected trend : at low Q_s, as a_s increases C_F decreases. The reason for this behaviour is not clear but it is possible that it arises from the characteristics of the float at different air/solid ratios : at low a_s the agglomerates are large with a "fluffy", loosely bound appearance; the float has a loose, porous structure, apparantly with a high degree of bound water. At high a_s the agglomerates are smaller and of denser appearance, the float has a dense, compact structure, apparantly with less bound water. The agglomerate characteristics, it is felt, affect the expulsion of the free interstitial and bound waters in different ways and results in the behavioural pattern observed.

There is felt to be little merit in repeating the series of plots shown in Figures 24(a) to 24(i) for

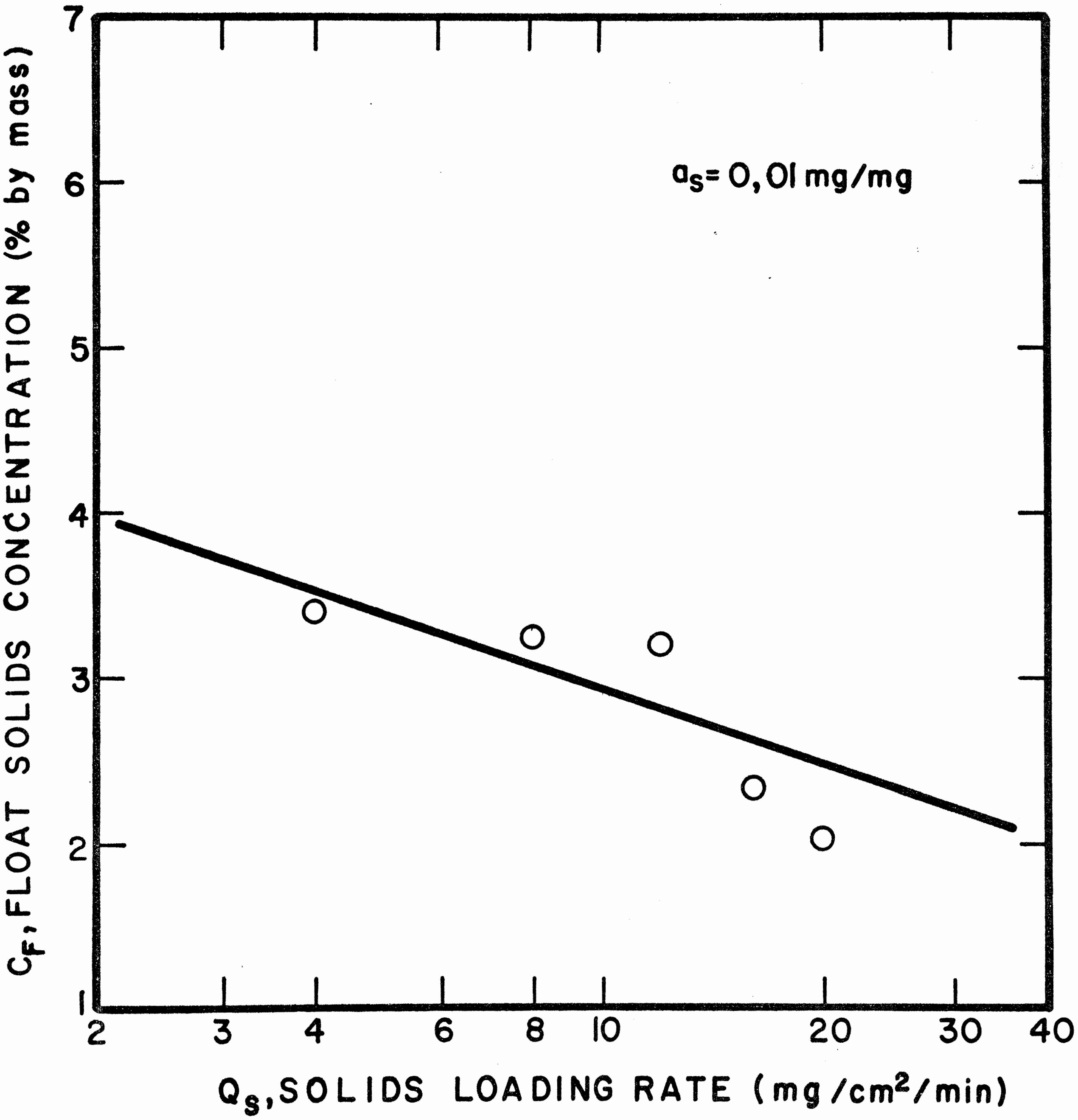


FIGURE 24(a): Relationship between Float Solids Concentration, C_F, and Solids Loading Rate, Q_s, at Air/Solids Ratio, a_s = 0,01 mg/mg. (Table 12).

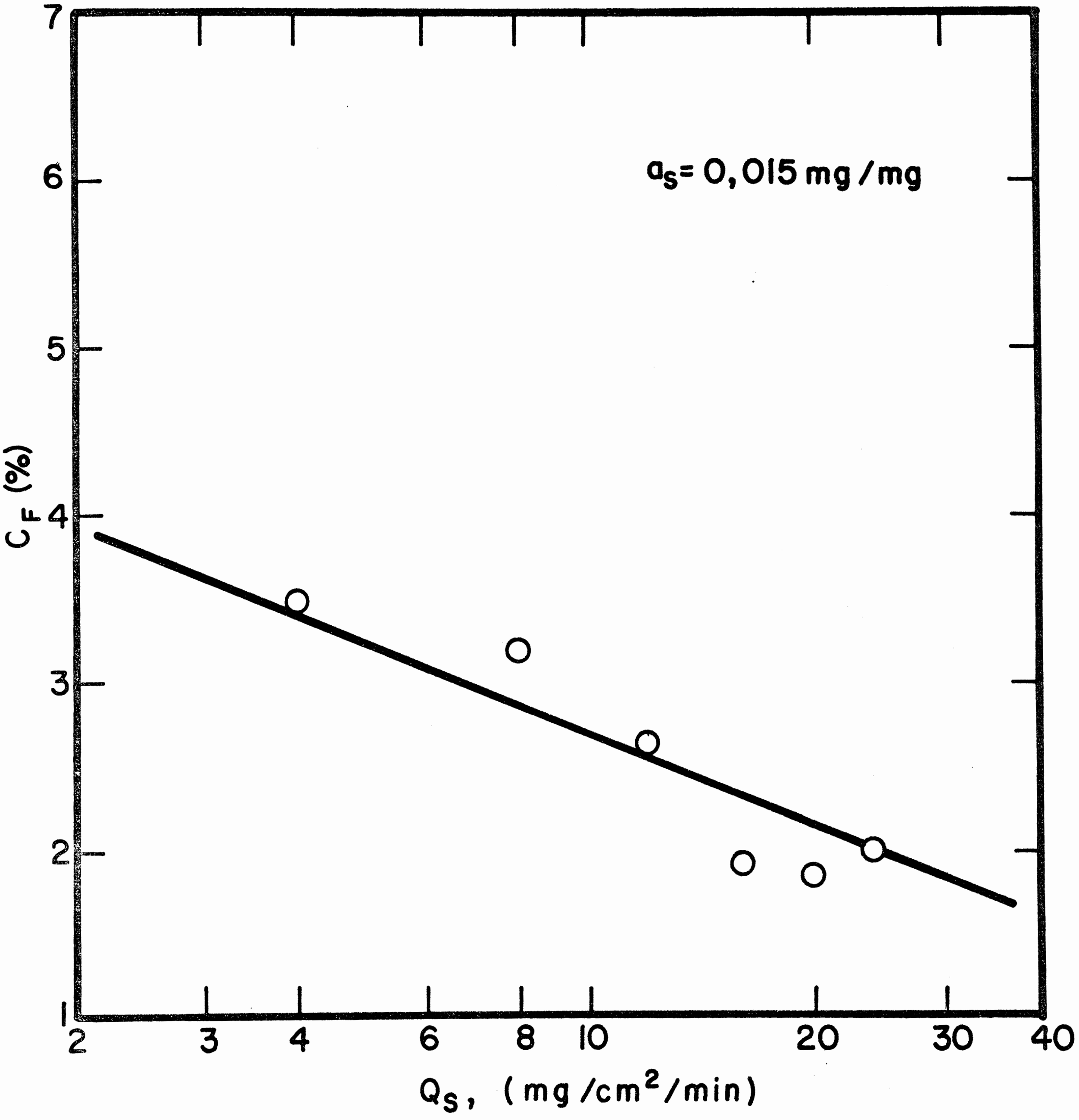


FIGURE 24(b): As for Figure 24(a) but a_s = 0,015 mg/mg (Table 13).

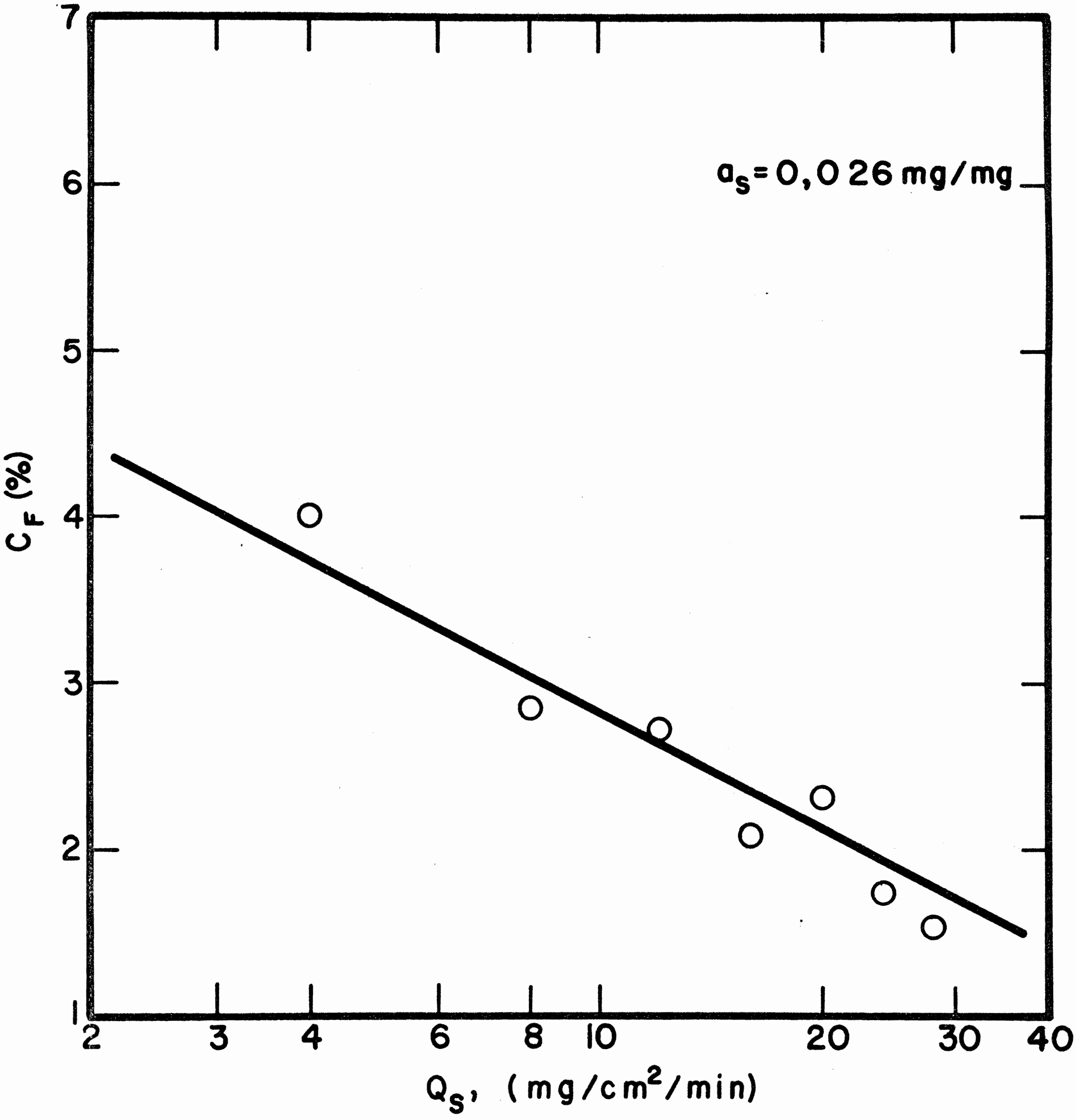


FIGURE 24(c): As for Figure 24(a) but a_s = 0,026 mg/mg (Table 14).

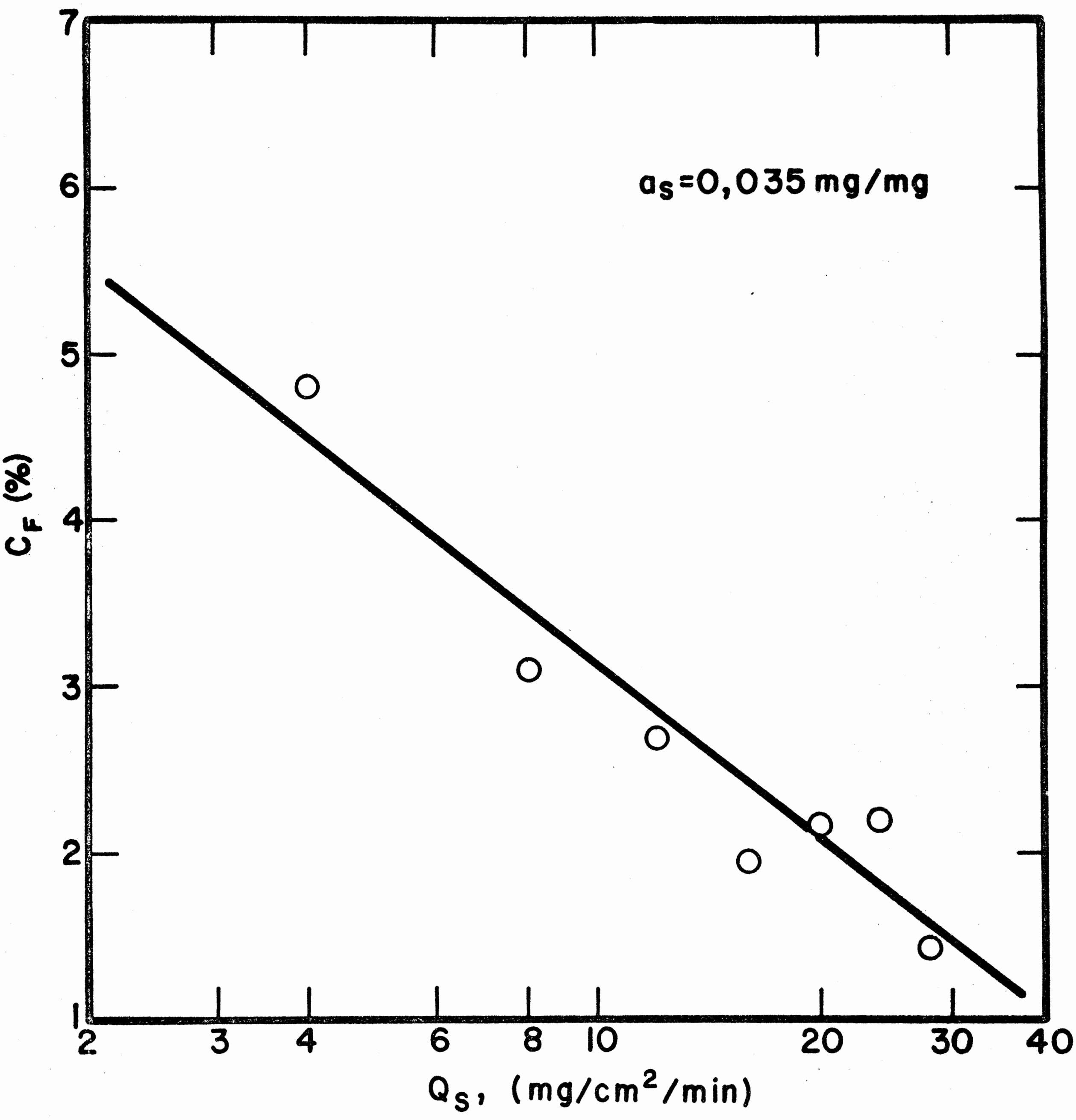


FIGURE 24(d): As for Figure 24(a) but a_s = 0,035 mg/mg (Table 15).

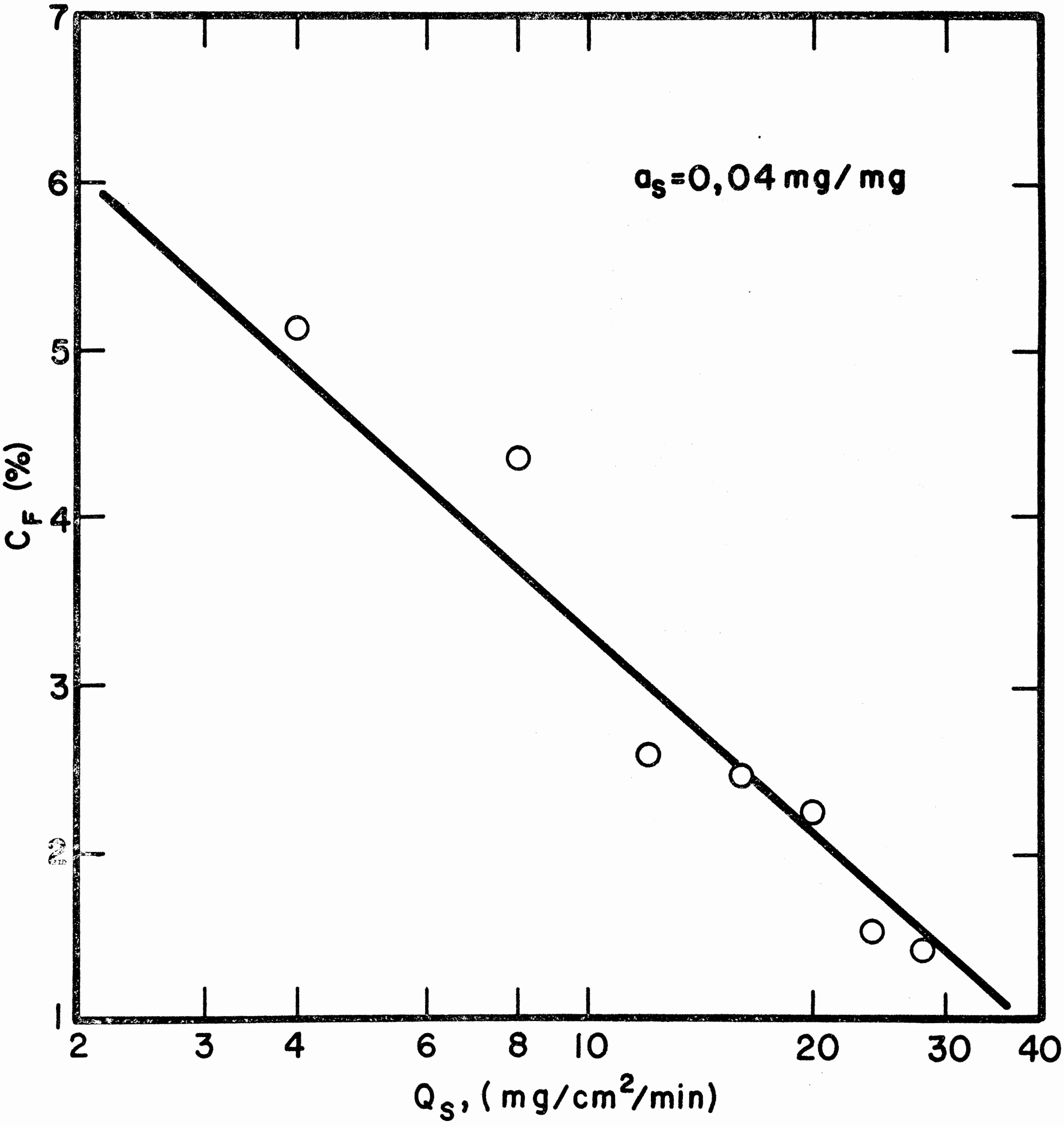


FIGURE 24(e): As for Figure 24(a) but a_s =0,04 mg/mg (Table 16).

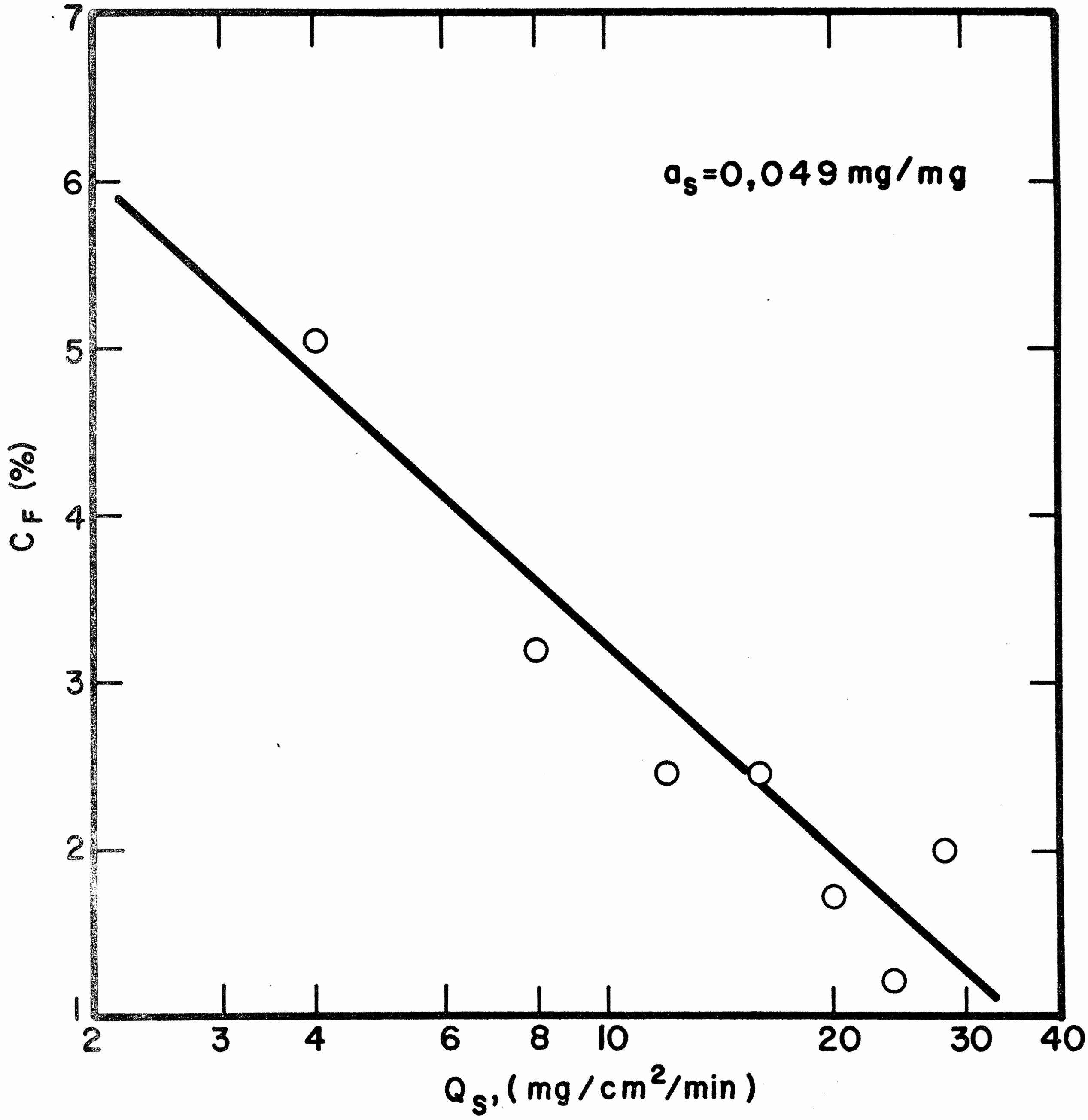


FIGURE 24(f): As for Figure 24(a) but a_s = 0,049 mg/mg (Table 17).

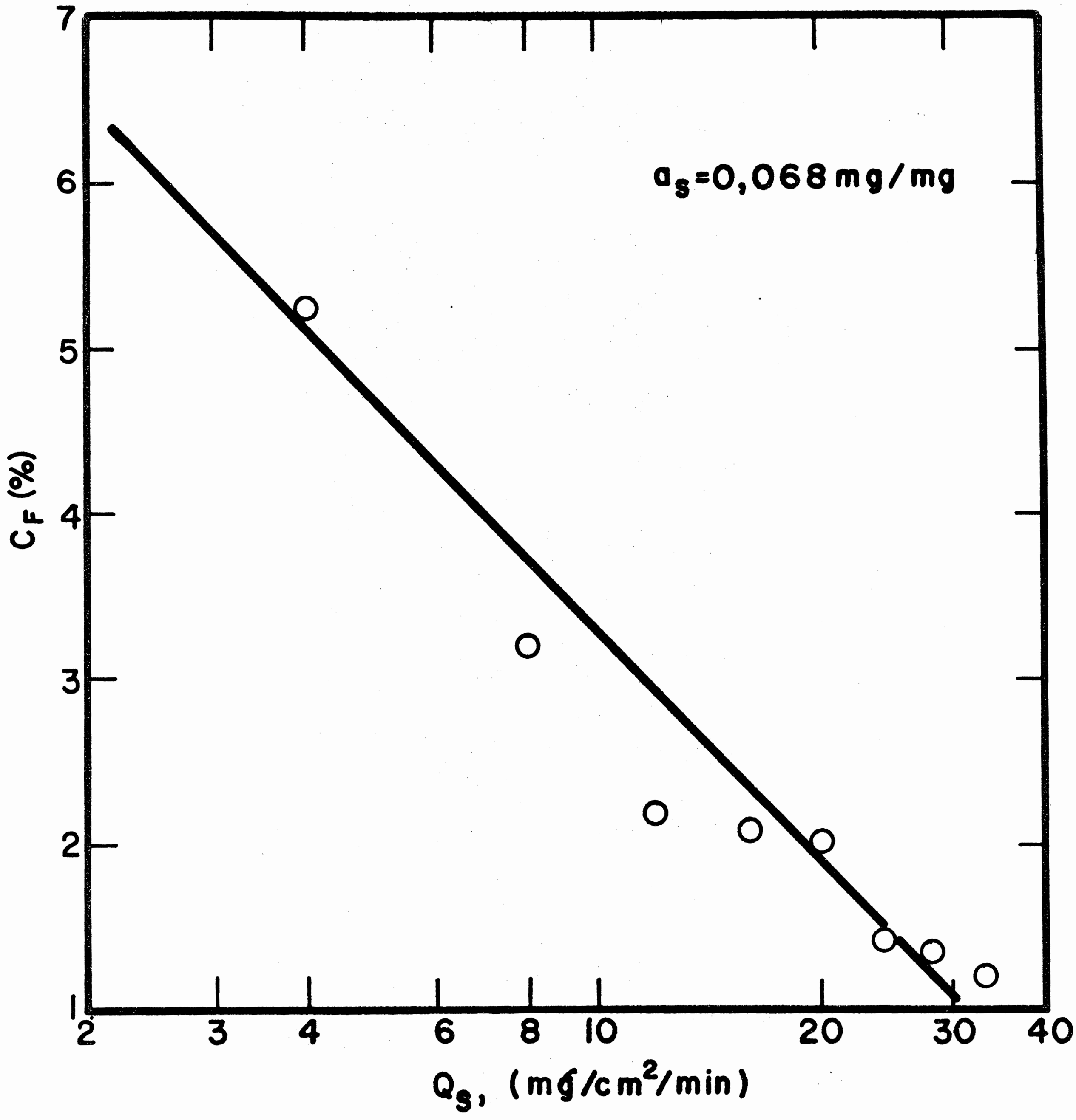


FIGURE 24(g): As for Figure 24(a) but a_s = 0,068 mg/mg (Table 18).

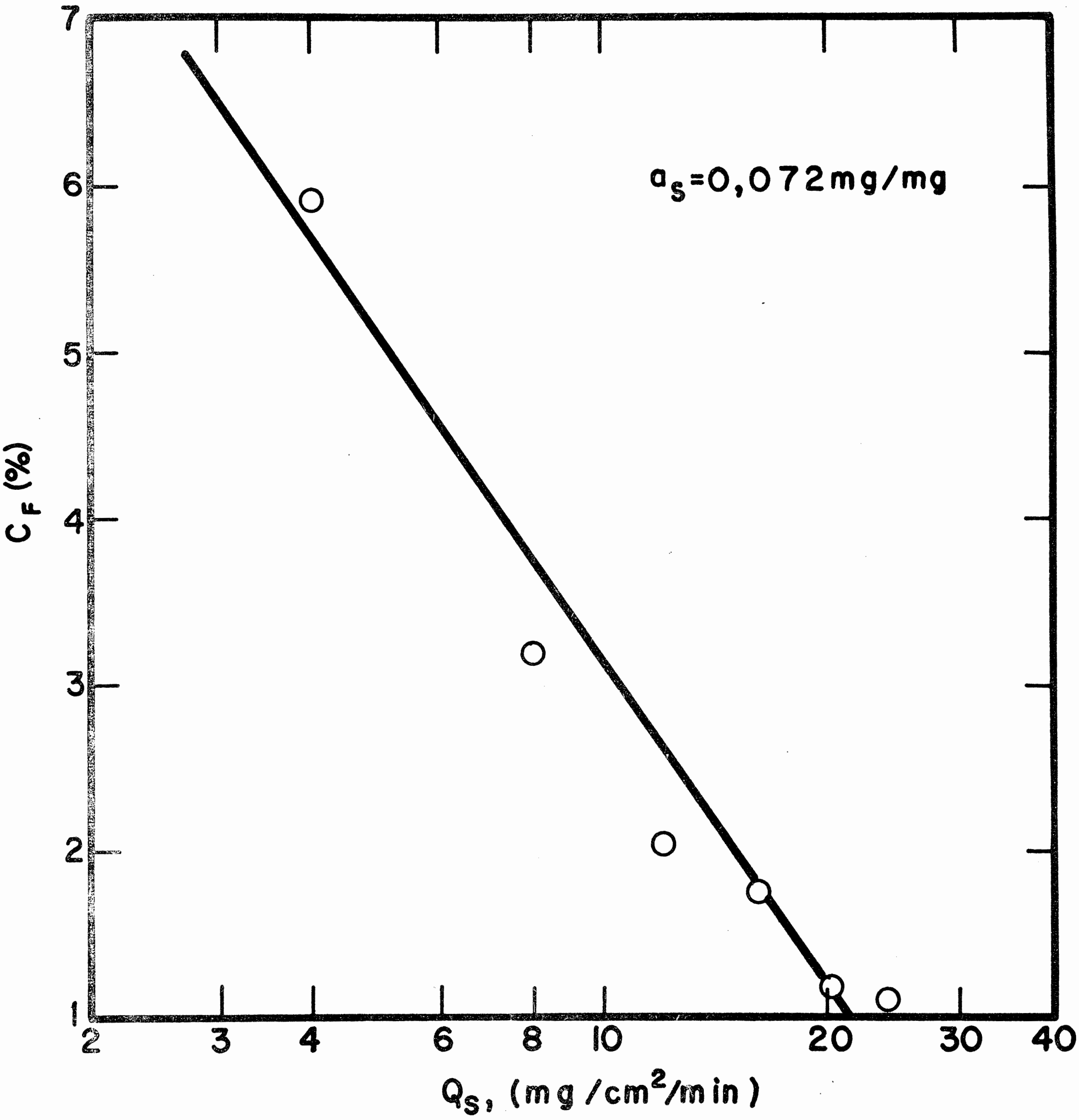


FIGURE 24(h): As for Figure 24(a) but a_s = 0,072 mg/mg (Table 19).

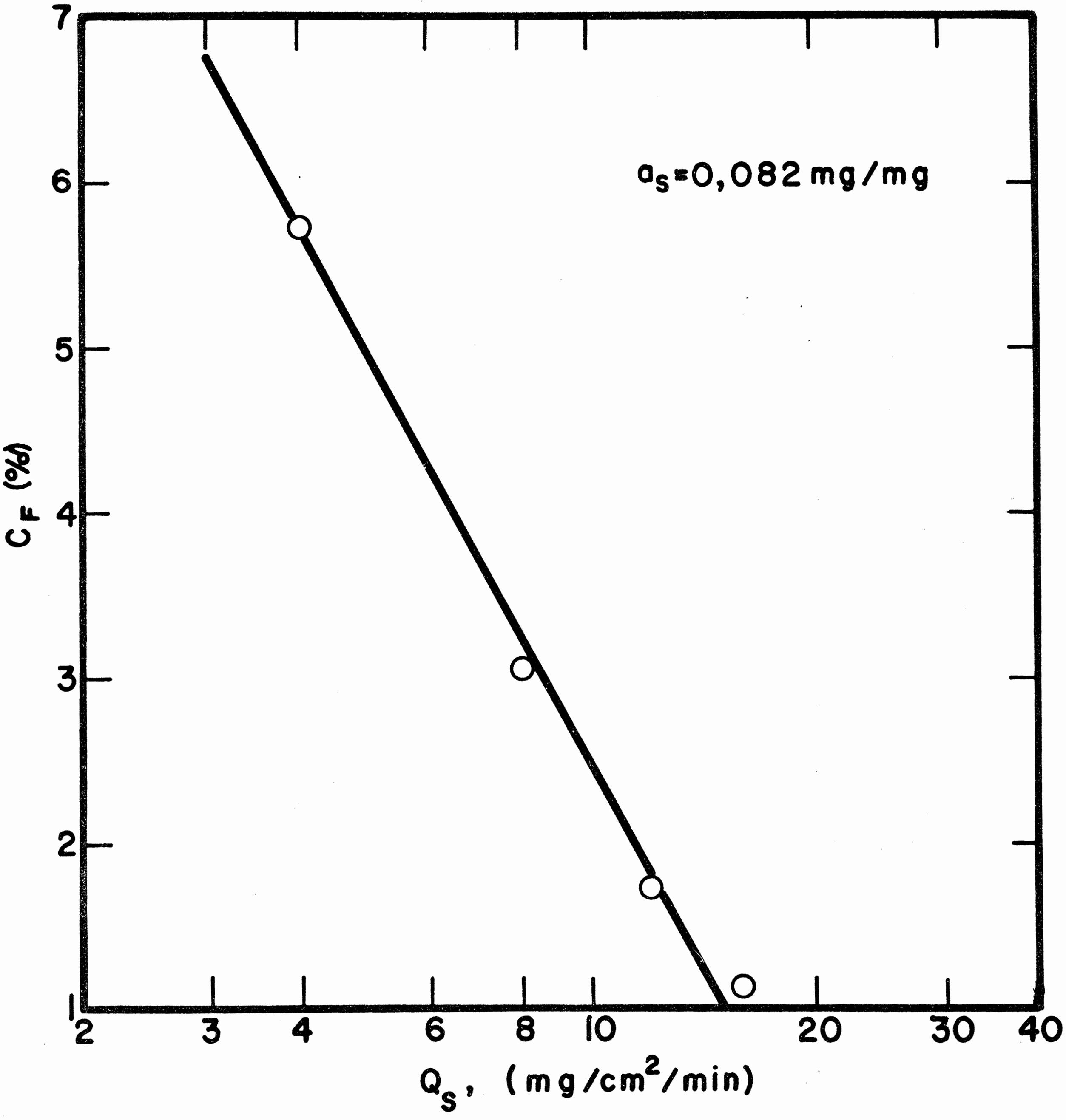


FIGURE 24(i): As for Figure 24(a) but a_s = 0,082 mg/mg (Table 20).

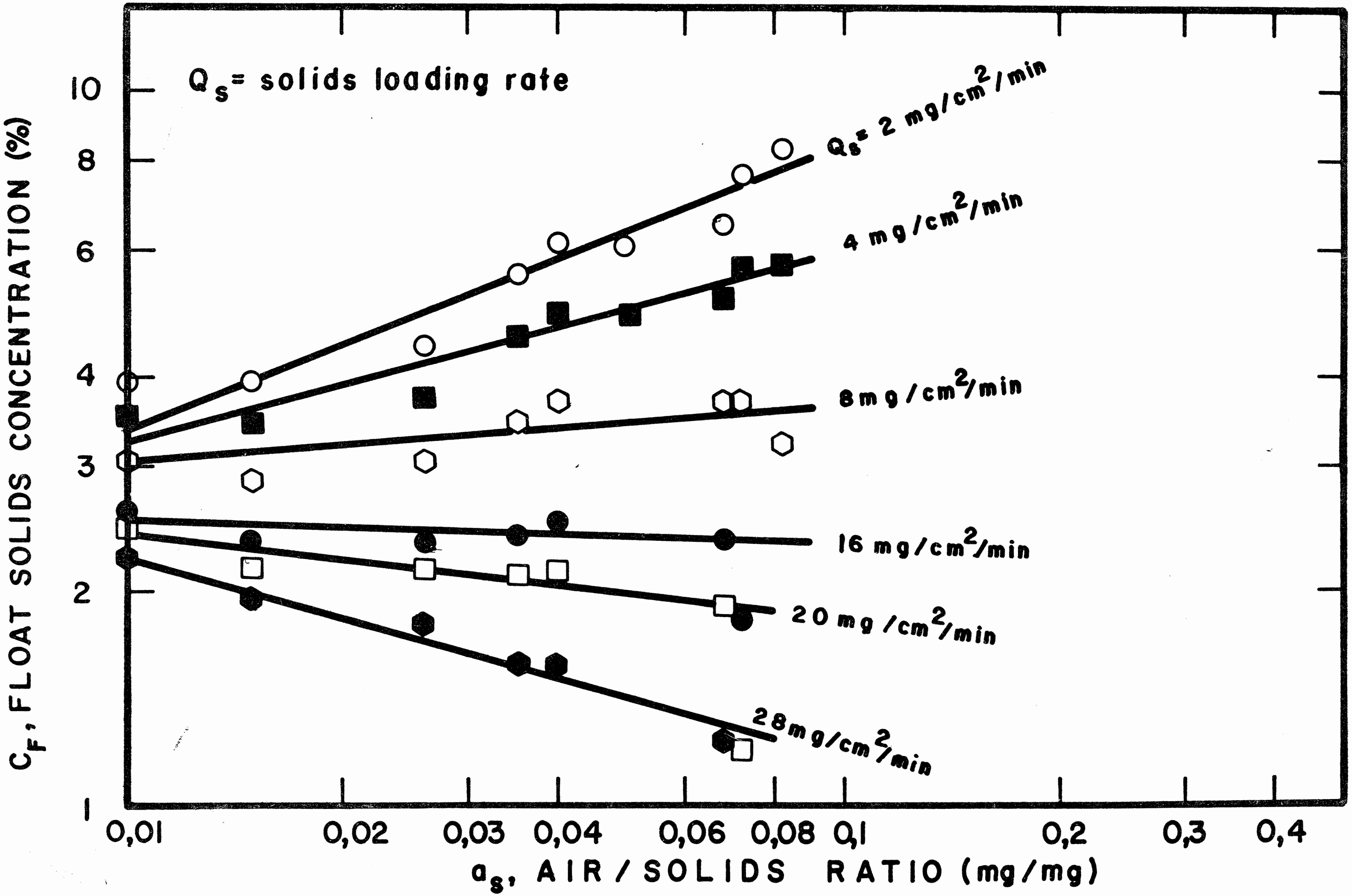


FIGURE 25: Relationship between Float Solids Concentration (C_F) and Air/Solids Ratio (a_s) for various solid loading rates, Q_s. (Values taken from Figures 24(a) to 24(i)).

each value of d_W since the trends would merely be repeated. Instead, Figure 26 shows the effect of d_W on C_F at different values of a_s for a select series of Q_s.

It seems that at any a_s, the value of C_F for particular values of Q_s and a_s, does depend on d_W, but not in a simple fashion. One can visualise that as d_W increases the retention time of the float solids above the water level (that is, the time for dewatering) is increased so that C_F increases. This effect is more pronounced at high Q_s for then the retention time of the float above the water level is relatively short and even a small increase in d_W increases the retention time to a much greater extent (in a relative sense), than when the retention time is long, that is, when Q_s is low. Consequently, at high Q_s values, increases in d_W have a significant effect on C_F while at low Q_S values, increases in d_W have a relatively minor effect.

Total Float Depth

The total depth occupied by the float is given by $(d_W + d_B)$, where d_B = depth of float solids below the water level, required to support the depth of float solids above, d_W. The ratio d_B/d_W appears to be principally a function of the air/solids ratio, a_s, as is evident in Figure 27, the experimental points representing a wide range of Q_s values.

The investigation into the float characteristics reported here must be considered to be of a very preliminary nature. The data obtained provide information in that the influence of parameters Q_s, a_s and d_W may be predicted on a qualitative and semi-quantitative basis. It was not, however, possible to present an integrated theory quantitatively incorporating the effects of Q_s, a_s and d_W on C_F. This is an aspect requiring further investigation.

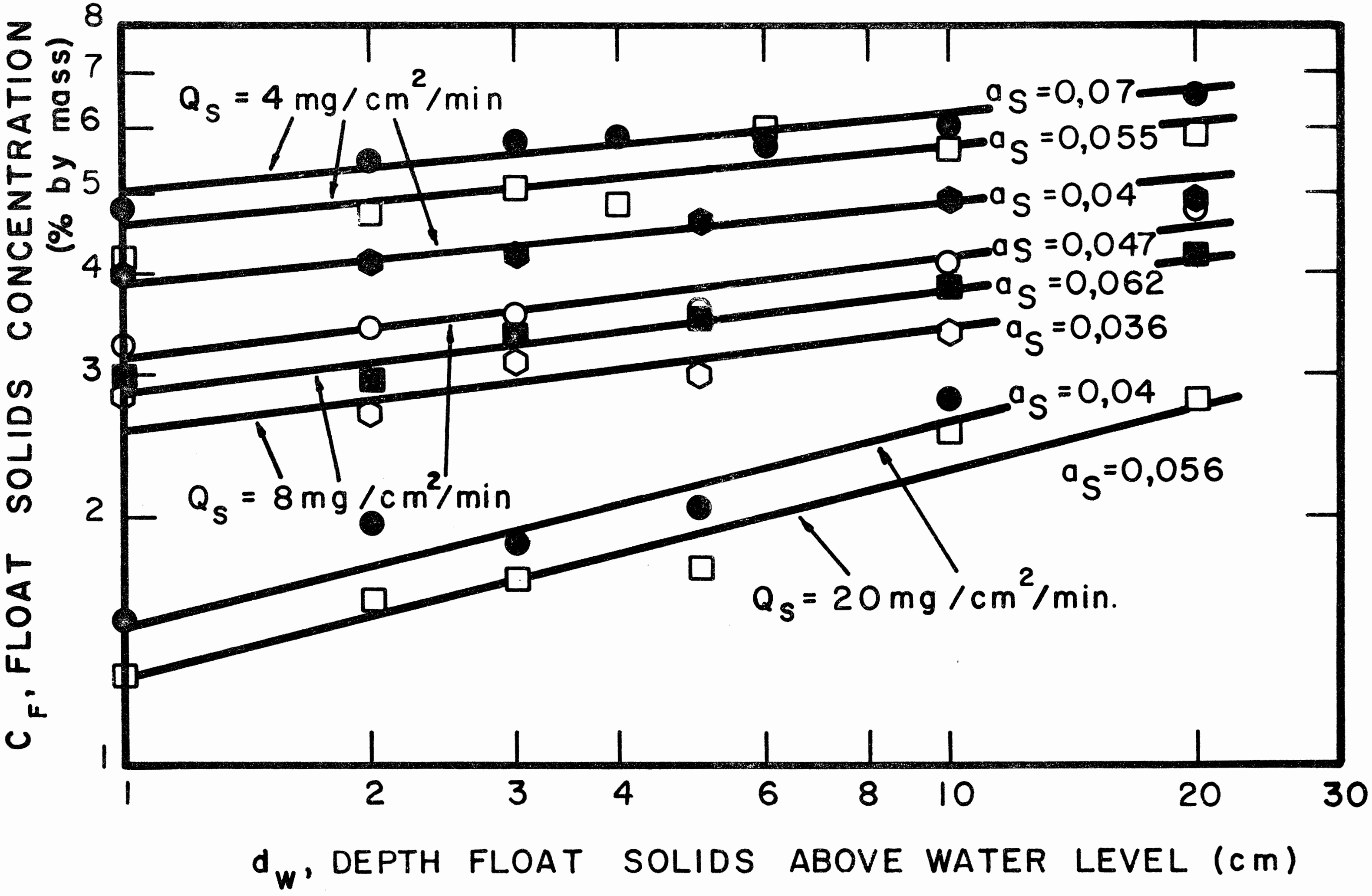


FIGURE 26: Effect of Depth Float Solids above Water Level on Float Solids Concentration for various Air Solid Ratios and Solid Loading Rates (Table 21).

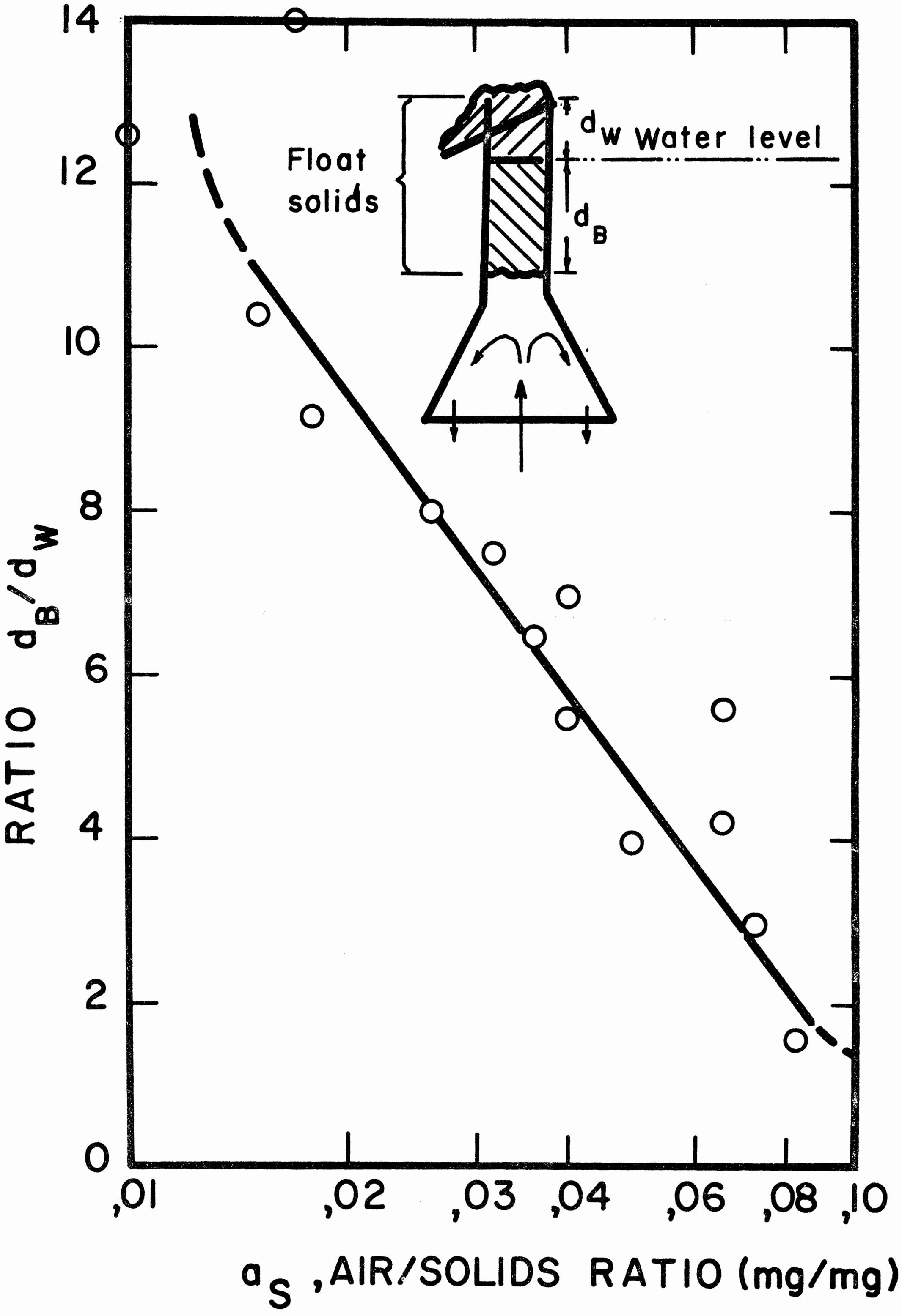


FIGURE 27: Relationship between Air/Solids Ratio and the Ratio (Depth Float Solids below Water Level, d_B)/(Depth Float Solids Above Water Level, d_W) (Table 22).

Design Procedure

The design of the thickening part of a flotation tank involves a comparative assessment between (a) the capital costs of providing deep tanks and tanks of large thickening area and (b) the running costs in providing large air/solids ratios.

Assuming that several plots such as Figure 25 are available, each at a particular value of d_W, a suitable design procedure is as follows:

1. Select a value of d_W.

2. From a plot of C_F versus a_s at different values of Q_s, for the particular value of d_W (for example, Figure 25), the combinations of Q_s and a_s to give a desired value of C_F are abstracted.

3. On the basis of economics, the optimum value of Q_s and a_s are determined : the parameter Q_s determines the area of the flotation tank for thickening (assuming that the wastewater flow rate and concentration are known) and parameters a_s and d_W determine the depth (see Figure 27).

4. The above procedure is repeated for several values of d_W. The optimum costs for each value of d_W are compared and the lowest will give the most economic combinations of d_W, a_s and Q_s.

ILLUSTRATIVE DESIGN EXAMPLE

It is required to design a flotation tank for clarification and for thickening. The wastewater is algal laden oxidation pond water with a solids concentration of 450 mg/l. The wastewater flow rate is 500 l/min. From the data presented in this thesis, (complete data are not presented), design a flotation tank to give 99,5% solids removal and a float solids concentration of 5%

Clarification

From Figure 23, v_L, the limiting downflow rate beyond which solids are carried down with the effluent may be determined at different values of a_s, the air/solids ratio.

e.g. At $a_s = 0{,}02$ mg/mg, $v_L = 25\ \mathrm{cm}^3/\mathrm{cm}^2/\mathrm{min}$.

Now, v_L = (Total hydraulic flow rate, Q_T) ÷ (cross sectional area at outlet of unit, A_2)

Assuming a wastewater feed : saturator feed ratio of 5:1, say, then $Q_T = \frac{6}{5} \times 500$ l/min = 600 l/min.

The area, A_2, at the outlet of the flotation unit is therefore given by

$$A_2 = \frac{Q_T}{v_L} = \frac{600 \times 10^3}{25}\ \mathrm{cm}^2$$

$$= \frac{600}{25} \times \frac{10^3}{10^4}\ \mathrm{m}^2$$

$$= \underline{\underline{2{,}4\ \mathrm{m}^2}}$$

For an air/solids ratio of 0,02 mg/mg, the requisite mass of air precipitated per litre of saturator feed, a_p, is given by

$$a_s = \frac{Q_{sat} \times a_p}{Q_w \times C_i}$$

where

a_s = air/solids ratio (mg/mg)

Q_{sat} = saturator feed rate (l/min)

a_p = mass air precipitated per litre saturator feed (mg/l).

Q_w = wastewater feed rate (l/min)

C_i = wastewater solids concentration (mg/l).

Hence

$$a_p = \frac{a_s \times Q_w \times C_i}{Q_{sat}} = \frac{0{,}02 \times 500 \times 450}{100}$$

$$= \underline{45} \text{ mg air/litre saturator feed.}$$

From Figure 13(a) (or 13(b)) the saturator pressure needs to be <u>380 kPa</u> (55 lb/in^2).

For a complete design, several value of a_s would be chosen and the above calculations repeated. Various wastewater feed : saturator feed ratios would also be tried. An economic comparison would then be made involving the capital (and running) costs of the saturation unit and the capital cost of the flotation tank.

Thickening

From Figure 25, for $\underline{d_W = 5\text{cm}}$, the combinations of Q_s and a_s to give a $\overline{\text{value of}}$ C_F = 5% are :

Q_s = 2 $mg/cm^2/min$, a_s = 0,027 mg/mg;
Q_s = 4 $mg/cm^2/min$, a_s = 0,051 mg/mg; etc.

(Other combinations are possible).

The area requirement of the flotation tank, for thickening, is given by

$$A_1 = (C_i - C_e)\, Q_w/Q_s$$

where A_1 = area of flotation tank for thickening, (cm^2);
C_i = influent wastewater solids concentration (mg/l);
C_e = effluent solids concentration (mg/l);
Q_w = wastewater flow rate (l/min);
Q_s = solids loading rate $(mg/cm^2/min)$.

For $Q_s = 2\ mg/cm^2/min$:
$A_1 = 448 \times 500/2\ cm^2$
$= 112\,000\ cm^2$
$= \underline{11{,}2\ m^2}$

The total depth occupied by the float is given by $(d_W + d_B)$. From Figure 27, at $a_s = 0{,}027$, $d_B/d_W = 8$. Therefore, for d_W = 5cm, the total depth = 5 +(5 x 8)= <u>45cm</u>.

For $Q_s = 4\ mg/cm^2/min$:
$A_1 = \underline{5{,}6\ m^2}$

From Figure 27, at $a_s = 0{,}051$, $d_B/d_W = 4{,}5$
Therefore, total depth = 5 +(5 x 4,5)= <u>27,5cm</u>.

Hence, for a float solids concentration of 5% the following specifications are required:

(1) Area = $11{,}2\ m^2$
Float depth = 45 cm
Air/Solids ratio = 0,027 mg/mg

(2) Area = $5{,}6\ m^2$
Float depth = 27,5 cm
Air/Solids ratio = 0,051 mg/mg

These calculations could be extended for further values of Q_s at the value for d_W of 5cm and an economic comparison made between the combinations of area, total float depth and air/solids ratio. (From the air/solids ratio, the parameter, a_p, is determined, as before).

This procedure is repeated for several values of d_W, say 10cm, 15cm and 20cm. A comparison is then made between the optimum cost at each value of d_W, which will give the optimum (economic) value of d_W together with the corresponding (optimum) values of area, float depth and air/solids ratio.

The above (illustrative) design example ingores any practical considerations pertaining to inlet and outlet arrangements.

FUTURE INVESTIGATIONS

The sucessful completion of the objectives set for this thesis has led to an increased awareness of the potential that the method of flotation holds in solids-liquid separation.

It is essential that the system developed be tested at pilot scale level to determine if any modifications are necessary when scaling up. Each of the units comprising the system requires further investigation at pilot scale level to ensure optimum performance.

Saturation Unit

It was shown that saturation of the pressurized water results in more efficient precipitation in the flotation unit at high pressures than at low pressures. At the present stage the results are only qualitative in nature and it would seem necessary that quantitative expressions be developed to determine the rate of precipitation at different saturator pressures.

It also seems possible to improve the saturator itself. Increased efficiency of dissolution and hence smaller saturator volumes may be attained by incorporating (say) slats in the path of the rising bubbles. This merits further investigation.

Flotation Unit

One of the favourable features of the flotation unit is that no collecting scraper mechanism is required to remove the float solids. In large scale plants, however, trouble may be experienced in ensuring a uniform rise of the float over the whole discharge area. A suggestion to ensure uniform discharge is to sub-divide the discharge area into smaller areal units by having deep dividing plates extending to below the bottom level of the float. A multiple weir system at the top is also a possibility.

A problem which was not investigated was the removal of particles which settle to the bottom of the flotation unit. (In the unit used in these investigations, all particles not carried upwards to the float were discharged with the effluent). Perhaps a wedged shaped bottom with a desludging point at the bottom apex of the wedge and clear water drawoff at a higher level may prove satisfactory.

Thickening

Further investigations into the mechanism of thickening by flotation is necessary. The information reported in this thesis is of a very preliminary nature. Extensive investigations using different wastewaters may lead to general and quantitative expressions linking the different parameters. A reasonable design procedure requires the minimum of prior experimentation.

Activated Sludge

Application of flotation for solids-liquid separation in activated sludge plants is another area meriting investigation. Recycling of thickened solids to the reactor is necessary to maintain a desired MLSS (mixed liquor suspended solids) concentration in the reactor. By using flotation, the high float solids concentrations attainable compared to the low concentrations attainable with sedimentation tanks, will greatly reduce the rate of recycle. Furthermore, it may be possible to operate the reactor at higher solid concentrations which will result in a reduction of the reactor volume requirements.(20)

A further application of flotation in the activated sludge system is the thickening of the wasted sludge prior to discharge to anaerobic digestors.

APPENDIX I - HYDROPHOBIC AND HYDROPHILIC COLLOIDS

There are two general classes of colloids encountered in water and wastewater and these are given the names hydrophobic and hydrophilic. Hydro- refers to the liquid phase (water) in which the colloids exist, and -phobic and -philic refer to the degree of affinity of the colloids for the liquid phase (i.e. hydrophobic = "water hating"; hydrophilic = "water loving"). Although the term hydrophobic infers a definite solid-water boundary, there is in reality a layer perhaps only a molecule thick where water molecules are attracted to the solid surface strongly enough for them to be carried with the colloid as it moves around in the water - the plane of shear then is a water-water boundary. The water molecules are attached to the solid surface by means of electrostatic attraction since not only do water molecules exist as a dipole with one end negative and the other positive, but the colloid particles possess a surface charge which is almost always negative in the conditions dealt with in water and wastewater treatment.

The term "hydrophilic" describes colloids which have a high degree of hydration, the amount of water bound by these particles sometimes being three to ten times their own dry weight. Hydrophilic colloids may be thought of as either having water molecules co-ordinatively bound to their surface layers, or as consisting of a lattice of "water-armoured" molecules.

APPENDIX II - RAPID MIXING

Rapid mixing refers to that stage in the overall coagulation-flocculation process where complete mixing of the coagulant chemicals into the stream to be treated is achieved.

When carrying out coagulation tests in the laboratory, it will be found that optimum solids removal will be achieved by a particular combination of coagulant concentration and pH. The conditions of the laboratory experiments are such that complete and instantaneous mixing of the coagulant chemicals into the sample of 'water' will be achieved readily.

In practice however, by applying the optimum coagulant dosage and pH as found in the laboratory, one usually finds that the same degree of solids removal cannot be achieved. This is largely due to an inadequate appraisal of the mode of destabilization necessary for the particular type of 'water' in question - and consequently the installation of a rapid mixing tank which cannot provide the requisite mixing environment.

The rapid mixers commonly installed in practice are typified by Figure 1. Such mixers are given the general name 'back-mix reactors' and their characteristic (and in some situations - their shortcoming) is that the particles within them have a distribution of residence times. That is, some of the particles will be short-circuited out almost immediately whereas some will stay in the reactor for longer periods than the nominal time. These 'back-mix reactors' can be undesirable for the following reasons:

(a) Due to long residence times extensive hydrolysis of <u>some</u> of the applied metal coagulant to the relatively inefficient $Fe(OH)_3$ or $Al(OH)_3$ will take place

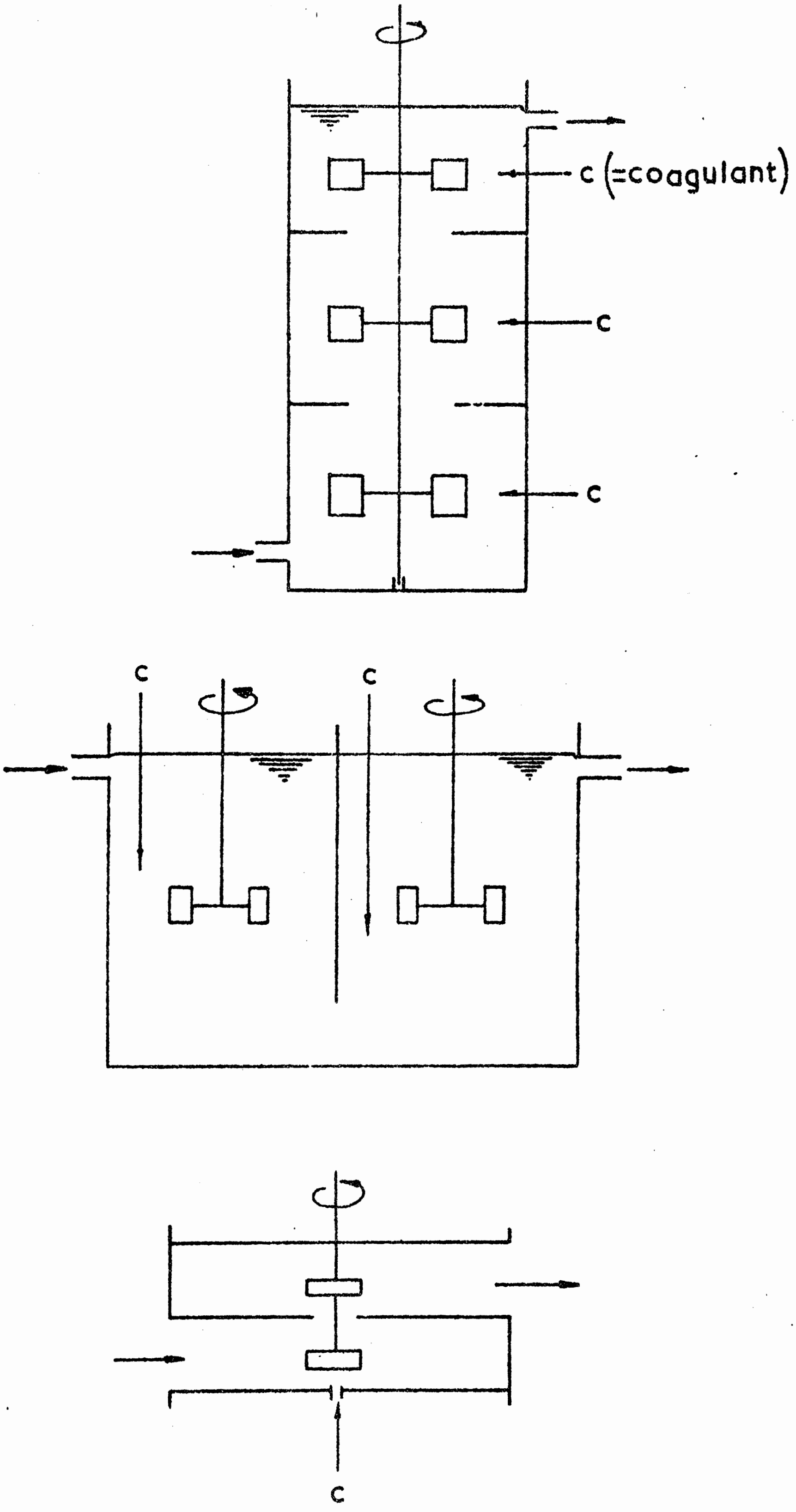


FIGURE 1: Representative Types of Mechanical Agitation Devices commonly used for Rapid Mixing.

(b) Poor overall destabilization of the particles will occur due to either incomplete adsorption or extensive adsorption of metal hydroxide species.

What is needed is for the same ('best') metal-hydroxide species to be available to every particle simultaneously. A plug flow type of rapid mixer will be seen to approach this 'ideal' situation the most satisfactorily since by virtue of its nature, all elements of liquid within it will theoretically have the same retention time.

Vrale and Jorden(19) investigated the above with particular emphasis on an adsorptive mechanism of destabilization by metal hydroxide species. In their experiments they devised the five units shown in Figure 2.

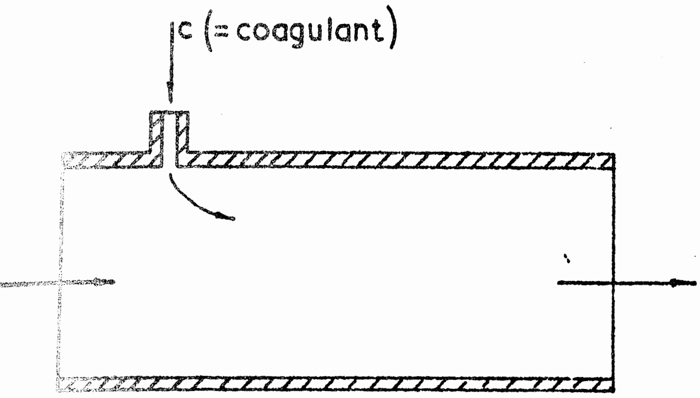
Unit 1 was designed to investigate the mixing effect within a pipe.

Unit 2 was designed to simulate a 'back-mix' reactor.

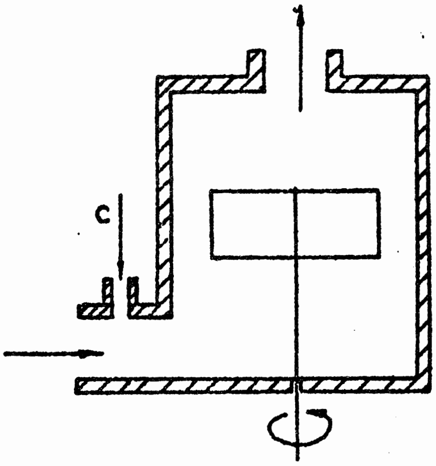
Unit 3 was selected because it had been proved elsewhere to afford a very rapid and high degree of mixing. It was intended that this unit should be the criterion for 'ultimate' rapid mixing.

Unit 4 was designed as a combination of Units 1 and 3. It was employed in order to determine the effect of the time lag before intense mixing occurred.

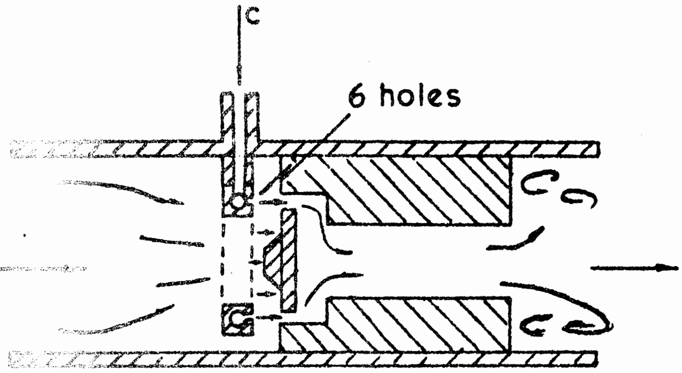
Unit 5 was designed to simulate Unit 3 without the latter's impractical head loss and consequental shearing intensity.



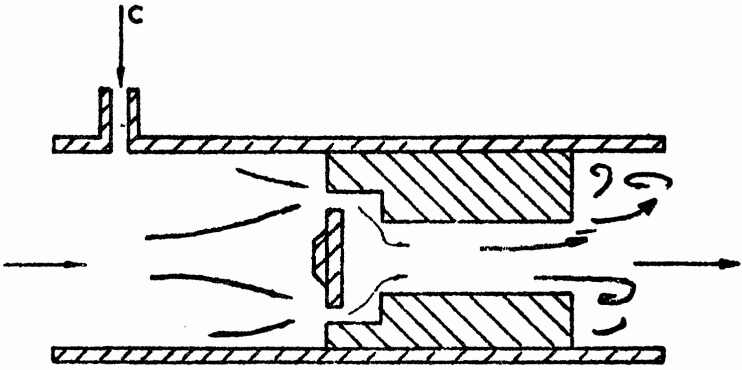
Unit No.1



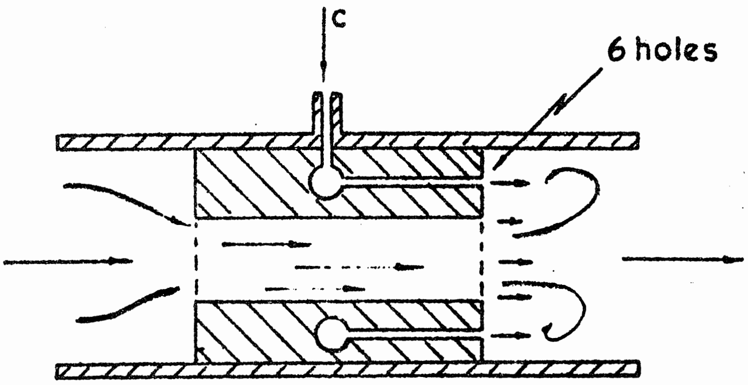
Unit No.2



Unit No.3



Unit No.4



Unit No.5

FIGURE 2: Rapid Mixing Units Devised by Vrale and Jorden(19). (All units have internal diameters of 2,54 cm).

The results obtained from their tests are illustrated in Figure 3 and the conclusions they came to with regard to the most efficient rapid mixing system are as follows: (Note: the criterion chosen for the efficiency of rapid mixing was the apparent aggregation rate K_{app} of the particles in a slow mix reactor - or flocculator.

where $$\frac{1}{T_t} = \frac{1}{T_o} + K_{app}.t$$

and T_t = supernatant turbidity after time t of flocculation

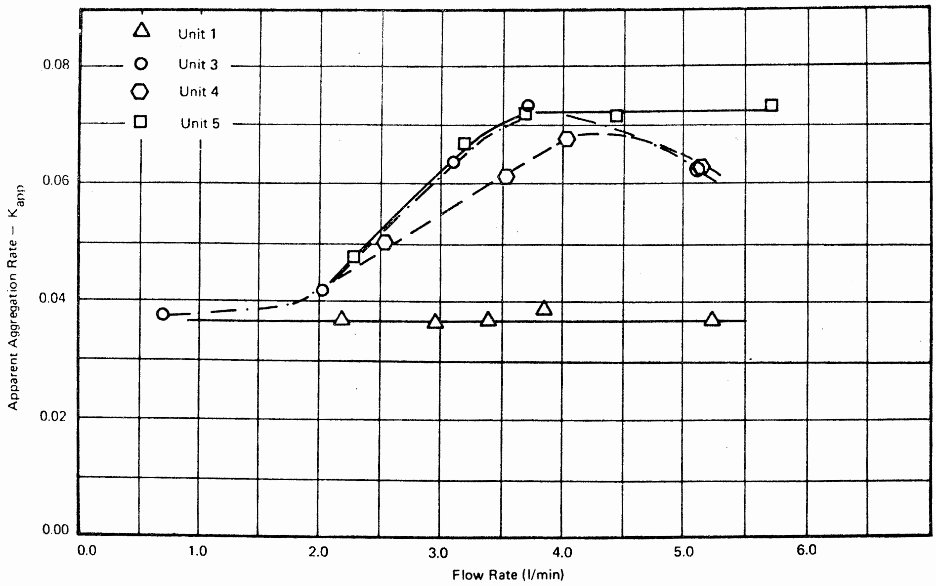
T_o = influent turbidity)

1. Unit 1 showed a constant rate of aggregation - <u>irrespective of the flow rate in the pipe</u>.

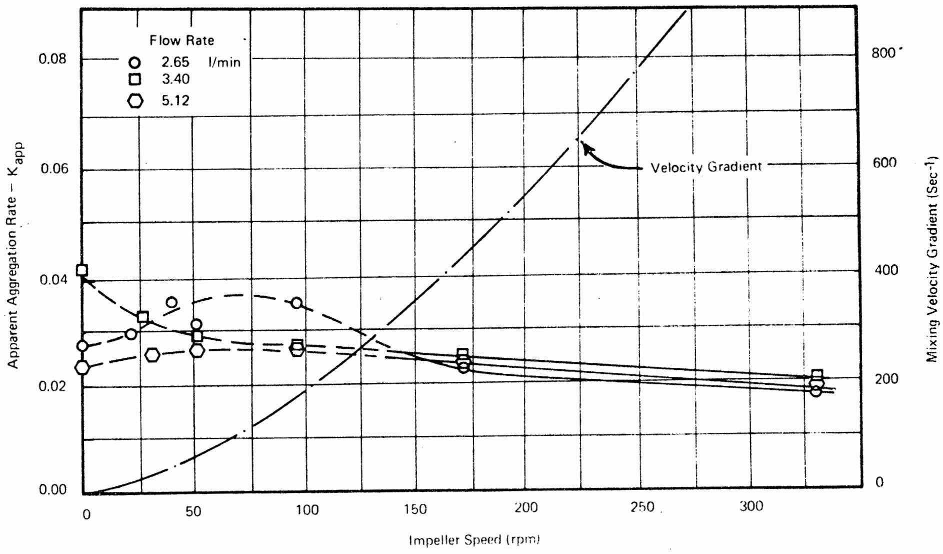
2. Unit 4 showed a marked increase in aggregation rate with increased flow rate, until at a certain flow rate, the rate of aggregation began to drop.

3. This drop in efficiency at higher flow rates was also inherent in Unit 3. It was reasoned that because of the very high head loss (and consequential shearing intensity) in these two units, then at the higher flow rates, break-up of the destabilized particles occured.

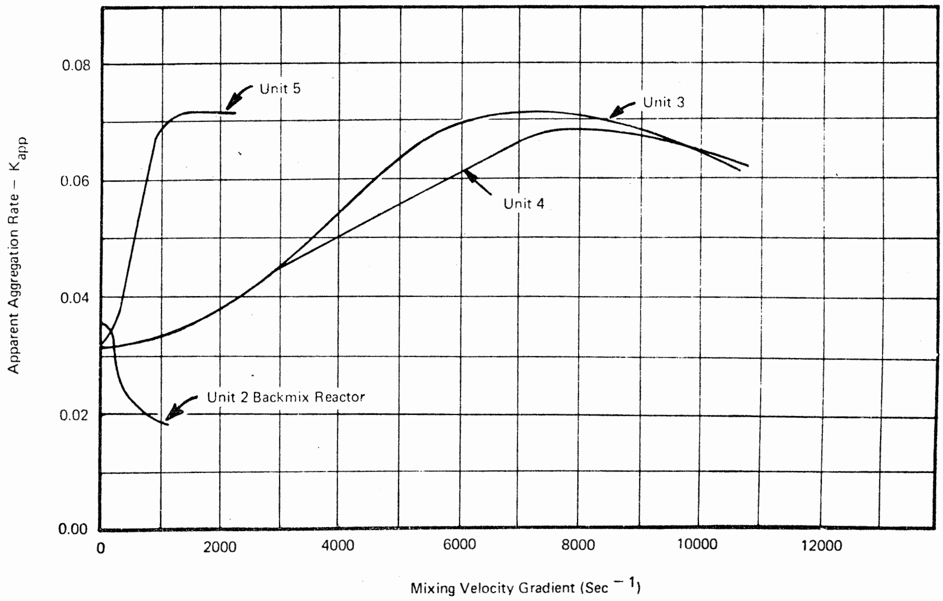
4. Unit 5 was found to give the best results since its efficiency reached a peak (similar to the peak in the 'next best' Unit 3) but did not decrease as did the others - it maintained a constant efficiency even at very high flow rates. The efficiency of Unit 5 reached a maximum when the hydraulic flow rate through the unit was increased to a value giving an average velocity gradient G of 1500 sec^{-1}, thereafter remaining constant.



Aggregation Rate Results Versus Flow Rate for All Plug-Flow Reactors



Aggregation Rate Results Versus Impeller Speed for Rapid-Mix Unit No. 2, a Backmix Reactor



Aggregation Rate Results Versus Mixing Velocity Gradient for Rapid-Mix Units 2 through 5

FIGURE 3

5. The results obtained for Unit 2, the back-mix reactor (the type most commonly employed in present practice) indicated that this is the worst possible method of rapid mixing with a hydrolizing metal, - increasing the impeller speed tended to decrease its efficiency, and conversely, at very low impeller speeds, incomplete mixing was obtained.

A very important point to emphasize at this stage is that the above results were based on a mechanism of adsorptive destabilization as described by Stern's model of the double layer or by the action of polyelectrolytes.

When a different 'type' of destabilization is employed then different considerations concerning rapid mixing will be necessary and one might find that a back-mix reactor would be preferable.

If it is found that one needs to operate at, say, pH 3,0 with Al(III) or Fe(III) then a 'Gouy-Chapman' type of destabilization will be effective and one will only need minimal rapid mixing since this type of destabilization is reversible.

If an entrappment or 'sweep-floc' type of destabilization is found to be necessary then a back-mix reactor might be desirable since it has been found favourable for the formation of metal-hydroxide precipitates.

If one desires nucleation effects to take place, such as in hydrophilic colloid destabilization (or in the water-softening process), then a back-mix reactor would also be desirable.

For sludge conditioning with say $FeCl_3$ then excessive turbulence would lead to floc break-up and hence an increase in particle surface area - this indicates that a laminar flow type reactor would be best.

For polyelectrolytes, where the method of destabilization is adsorptive in nature, then a unit such as Unit 5 would be best since it was found that it gave a high degree of efficiency with regard to adsorptive destabilization without an excessive degree of turbulence.

APPENDIX III - EXPERIMENTAL RESULTS

Table 1 - (Data for Figures 13(a) and 13(b)).

Mass of Air per litre of Pressure - Saturated Water Precipitated in Monitoring Unit, (a_p), at different Saturator Pressures, (P_g). Valid for Saturator feed rates from 0 to 10 l/min.

P_g (kPa)	a_p (mg air/litre water)
140	3,90 4,85
170	4,90 8,10 9,09
205	11,10 13,50 4,21
240	15,30 19,19 20,00 23,20
275	23,80 28,00 30,30 31,60
310	33,80 35,51 37,30
345	39,00
375	35,90 41,00 42,08
410	52,10 53,20 55,20 55,20 54,10 61,15

a_p calculated from time, T, taken to displace one litre of water in monitoring unit shown in Figure 12. (Hydrostatic head of 1 litre water in monitoring unit = 35,4cm)

$$\text{That is,} \quad a_p = \frac{(P_a + P_h)}{P_a} \times \frac{\rho_{air}}{V}$$

where P_a = atmospheric pressure (=1034 gm/cm^3)

P_h = hydrostatic pressure of displaced water (= 35,4 gm/cm^3)

ρ_{air} = density of air (= 1200 mg/l @ 20°C)

V = total volume of saturator feed required to precipitate sufficient air to displace 1 litre of water.

Since V = T (1 l/min saturator feed used) then

$$a_p = \frac{(P_a + P_h)}{P_a} \times \frac{\rho_{air}}{T}$$

Air/solids ratio calculated from

$$a_s = \frac{a_p \times Q_{sat}}{C_i \times Q_w}$$

where a_s = air/solids ratio (mg/mg)

a_p = mass air precipitated per litre saturator feed (mg)

Q_{sat} = saturator feed into flotation unit

C_i = wastewater solids concentration

Q_w = wastewater feed into flotation unit.

Table 2 - (Data for Figure 14(a)).

Supernatant turbidity in JAR TEST using Ferric Chloride ($Fe\,Cl_3$, $6H_2O$). Dosages expressed as Fe^{3+} (mg/l).

Dosage (mg/l)	pH	Turbidity (neph. rdng.)
10	4,0	56,0
20	"	31,2
30	"	18,8
40	"	11,2
50	"	8,0
60	"	6,9
10	5,0	43,0
20	"	21,2
30	"	9,7
40	"	4,5
50	"	3,1
60	"	3,2
70	"	2,8
10	6,0	30,2
20	"	14,5
30	"	6,8
40	"	3,6
50	"	2,2
60	"	2,8
70	"	3,0
10	7,0	34,7
20	"	17,7
30	"	8,6
40	"	5,0
50	"	4,0
60	"	3,9
70	"	3,7

Table 3 - (Data for Figure 14(b)).

Effluent turbidity in FLOTATION UNIT using Ferric Chloride ($FeCl_3, 6H_2O$). Dosage expressed as Fe^{3+}(mg/l).

Dosage (mg/l)	pH	Turbidity (neph. rdng.)
20	4,0	57,8
30	"	40,0
40	"	30,3
40	"	25,0
60	"	21,2
10	5,0	52,6
20	"	30,7
30	"	29,3
40	"	13,1
50	"	10,8
60	"	10,8
70	"	10,0
10	6,0	32,2
20	"	17,0
30	"	11,7
40	"	9,2
50	"	9,1
60	"	8,2
70	"	9,0
20	7,0	39,7
30	"	30,6
40	"	23,9
50	"	18,5
60	"	15,6

Table 4 - (Data for Figure 16(a)).

Supernatant turbidity in JAR TEST using Ferric Sulphate ($Fe_2(SO_4)_3.9H_2O$). Dosage expressed as Fe^{3+} (mg/l).

Dosage (mg/l)	pH	Turbidity (neph. rdng.)
10	4,0	72,8
20	"	31,1
30	"	14,5
40	"	17,9
50	"	3,0
60	"	3,1
70	"	3,8
10	5,0	26,2
20	"	10,5
30	"	3,9
40	"	0,8
50	"	0,1
60	"	0,2
70	"	1,2
10	6,0	32,4
20	"	13,5
30	"	5,6
40	"	1,8
50	"	0,8
60	"	1,0
70	"	0,2
10	7,0	59,9
20	"	30,0
30	"	17,9
40	"	9,9
50	"	4,3
60	"	3,9
70	"	5,6

Table 5 - (Data for Figure 16(b)).

Effluent turbidity in FLOTATION UNIT using Ferric Sulphate. ($Fe_2(SO_4)_3,9H_2O$). Dosages expressed as Fe^{3+} (mg/l).

Dosage (mg/l)	pH	Turbidity (neph. rdng.)
20	4,0	62,1
30	"	48,6
40	"	37,7
50	"	29,2
60	"	24,1
70	"	20,3
10	5,0	56,2
20	"	34,8
30	"	22,0
40	"	17,5
50	"	14,9
60	"	14,8
70	"	14,2
10	6,0	47,3
20	"	24,7
30	"	16,5
40	"	11,6
50	"	11,0
60	"	10,1
70	"	11,2
20	7,0	54,7
30	"	42,9
40	"	33,8
50	"	27,8
60	"	24,1
70	"	21,2

Table 6 - (Data for Figure 18(a)).

Supernatant turbidity in JAR TEST using Aluminium Sulphate. ($Al_2(SO_4)_3,12H_2O$). Dosages expressed as Al^{3+} (mg/l).

Dosage (mg/l)	pH	Turbidity (neph. rdng.)
10	4,0	68,1
15	"	54,0
20	"	47,0
25	"	42,6
30	"	37,8
5	5,0	30,0
10	"	17,4
15	"	8,4
20	"	5,7
25	"	5,5
30	"	5,0
35	"	5,9
40	"	5,5
5	6,0	38,5
10	"	24,5
15	"	13,4
20	"	8,0
25	"	6,0
30	"	6,1
35	"	6,5
40	"	5,5
10	7,0	55,2
15	"	39,7
20	"	34,9
25	"	28,5
30	"	26,8

Table 7 - (Data for Figure 18(b)).

Effluent turbidity in FLOTATION UNIT using Aluminium Sulphate. ($Al_2(SO_4)_3,12H_2O$). Dosages expressed as Al^{3+} (mg/l).

Dosage (mg/l)	pH	Turbidity (neph. rdng.)
10	4,0	95,0
15	"	91,8
20	"	91,3
25	"	88,0
30	"	87,9
5	5,0	54,7
10	"	34,1
15	"	27,7
20	"	22,8
25	"	23,0
30	"	22,0
35	"	23,0
40	"	23,8
5	6,0	62,9
10	"	38,0
15	"	26,7
20	"	22,1
25	"	20,5
30	"	20,9
35	"	21,3
40	"	20,2
10	7,0	59,9
15	"	48,0
20	"	42,8
25	"	36,9
30	"	34,4
35	"	33,0

Table 8 - (Data for Figure 20).

Effluent turbidity for different reaction times from point of coagulant addition to air precipitation.

Reaction Time (sec)	Effluent Turbidity (neph. rdng.)
1	9,4
5	9,1
10	7,6
15	6,1
20	5,9
25	6,2

Table 9 - (Data for Figure 21).

Effluent turbidity at different downflow rates and areas of upper chamber, A_1. Air/Solids Ratio = 0,0063 mg/mg.

Downflow Rate ($cm^3/cm^2/min$)	A_1 (cm^2)	Effluent Turbidity (neph. rdng.)
21,8	225	57,7
19,1	"	34,6
16,4	"	25,1
13,6	"	10,2
10,9	"	6,7
6,8	"	7,2
4,1	"	6,3
21,8	100	49,2
19,1	"	32,9
16,4	"	13,1
13,6	"	7,0
10,9	"	5,6
6,8	"	5,8
4,1	"	6,0
21,8	25	34,4
19,1	"	16,1
16,4	"	9,6
13,6	"	6,2
10,9	"	6,8
4,1	"	7,0

Table 10 - (Data for Figure 22(a)).

Effluent turbidity at different downflow rates for various air/solid ratios, a_s.

Downflow Rate ($cm^3/cm^2/min$)	a_s (mg/mg)	Effluent Turbidity (neph. rdng.)
5,4	0,0043	2,1
7,2	"	1,8
9,1	"	2,7
10,8	"	6,1
14,6	"	19,2
7,2	0,0054	2,1
9,1	"	1,9
10,8	"	3,3
14,6	"	11,2
16,4	"	16,1
9,1	0,0073	2,0
12,1	"	2,9
15,2	"	9,3
16,8	"	17,9
18,2	"	21,5
9,1	0,011	2,1
12,1	"	2,1
15,2	"	4,4
18,2	"	17,1
12,1	0,012	2,1
15,2	"	1,8
16,8	"	2,3
18,2	"	5,0
21,2	"	13,2
15,2	0,0125	2,0
18,2	"	2,1
21,2	"	10,4
24,3	"	20,7
18,2	0,0139	2,0
21,2	"	2,8
24,3	"	10,1
27,3	"	21,3
21,2	0,0167	1,9
24,3	"	3,4
27,3	"	11,5
24,3	0,0226	2,0
27,3	"	1,8
30,3	"	8,3

Table 11 - (Data for Figure 22(b)).

Effluent solids at different downflow rates for various air/solid ratios, a_s.

Downflow Rate ($cm^3/cm^2/min$)	a_s (mg/mg)	Effluent Solids (mg/l)
3,0	0,0027	2,9
6,1	"	5,2
9,1	"	17,1
12,2	"	63,1
15,2	"	132,5
3,0	0,0045	2,7
6,1	"	2,1
9,1	"	6,8
12,2	"	29,0
15,2	"	55,8
6,1	0,0080	2,1
9,1	"	2,0
12,2	"	4,4
15,2	"	27,3
18,2	"	105,2
12,2	0,0107	2,2
15,2	"	2,5
18,2	"	3,4
21,2	"	28,6
24,3	"	81,1
27,3	"	124,0
15,2	0,0169	3,5
18,2	"	3,0
21,2	"	3,1
24,3	"	12,1
27,3	"	50,2
30,3	"	102,3
21,2	0,0198	1,9
24,3	"	2,0
27,3	"	6,1
30,3	"	62,2
21,2	0,0217	1,9
24,3	"	1,2
27,3	"	1,0
30,3	"	6,7

Table 12 - (Data for Figure 24(a)).

Float Solids Concentration, C_F, at different Solids Loading Rates, Q_s. Air/Solids Ratio = 0,01 mg/mg.

Q_s (mg/cm^2/min)	C_f (%)
4	3,40
8	3,26
12	3,20
16	2,34
20	2,03

Table 13 - (Data for Figure 24(b)).

Float Solids Concentration, C_F, at different Solids Loading Rates, Q_s. Air/Solids Ratio = 0,015 mg/mg.

Q_s (mg/cm^2/min)	C_f (%)
4	3,49
8	3,20
12	2,64
16	1,92
20	1,86
24	2,01

Table 14 - (Data for Figure 24(c)).

Float Solids Concentration, C_F, at different Solids Loading Rates, Q_s. Air/Solids Ratio = 0,026 mg/mg.

Q_s (mg/cm^2/min)	C_F (%)
4	4,01
8	2,87
12	2,73
16	2,10
20	2,33
24	1,77
28	1,56

Table 15 - (Data for Figure 24(d)).

Float Solids Concentration, C_F, at different Solids Loading Rates, Q_s. Air/Solids Ratio = 0,035 mg/mg.

Q_s (mg/cm^2/min)	C_F (%)
4	4,80
8	3,10
12	2,79
16	1,96
20	2,18
24	2,20
28	1,43

Table 16 - (Data for Figure 24(e)).

Float Solids Concentration, C_F, at different Solids Loading Rates, Q_s. Air/Solids Ratio = 0,04 mg/mg.

Q_s (mg/cm^2/min)	C_F (%)
4	5,14
8	5,37
12	2,60
16	2,48
20	2,27
24	1,57
28	1,46

Table 17 - (Data for Figure 24(f)).

Float Solids Concentration, C_F, at different Solids Loading Rates, Q_s. Air/Solids Ratio = 0,049 mg/mg.

Q_s (mg/cm^2/min)	C_F (%)
4	5,07
8	3,20
12	2,48
16	2,46
20	1,73
24	1,22
28	2,00

Table 18 - (Data for Figure 24(g)).

Float Solids Concentration, C_F, at different Solids Loading Rates, Q_s. Air/Solids Ratio = 0,068 mg/mg.

Q_s (mg/cm^2/min)	C_F (%)
4	5,24
8	3,20
12	2,21
16	2,10
20	2,02
24	1,44
28	1,38
33	1,22

Table 19 - (Data for Figure 24(h)).

Float Solids Concentration, C_F, at different Solids Loading Rates, Q_s. Air/Solids Ratio = 0,072 mg/mg.

Q_s (mg/cm^2/min)	C_F (%)
4	5,89
8	3,20
12	2,07
16	1,78
20	1,21
24	1,12

Table 20 - (Data for Figure 24(i)).

Float Solids Concentration, C_F, at different Solids Loading Rates, Q_s. Air/Solids Ratio = 0,082 mg/mg.

Q_s (mg/cm^2/min)	C_F (%)
4	5,72
8	3,06
12	1,72
16	1,15

Table 21 - (Data for Figure 26).

Float Solids Concentration, C_F, at different Depth Float Solids above Water Level, d_W; for various Air/Solids Ratios, a_s, and Solids Loading Rates, Q_s.

d_W (cm)	a_s (mg/mg)	Q_s ($mg/cm^2/min$)	C_f (%)
1	0,070	4	4,32
2	"	"	5,45
3	"	"	5,76
4	"	"	5,81
6	"	"	5,70
10	"	"	6,00
20	"	"	6,59
1	0,055	4	4,14
2	"	"	5,29
3	"	"	5,02
4	"	"	4,81
6	"	"	5,99
10	"	"	5,64
20	"	"	5,90
1	0,04	4	4,01
2	"	"	4,11
3	"	"	4,18
6	"	"	4,57
10	"	"	4,83
20	"	"	4,89
1	0,047	8	3,23
2	"	"	3,40
3	"	"	3,52
5	"	"	3,61
10	"	"	4,10
20	"	"	4,72
1	0,062	8	2,99
2	"	"	2,94
3	"	"	3,33
5	"	"	3,44
10	"	"	3,82
20	"	"	4,20
1	0,036	8	2,81
2	"	"	2,66
3	"	"	3,08
5	"	"	2,98
10	"	"	3,35
1	0,04	20	1,51
2	"	"	1,96
3	"	"	1,84
5	"	"	2,03
10	"	"	2,78

Table 21 - (Continued).

d_W (cm)	a_s (mg/mg)	Q_s (mg/cm^2/min)	C_F (%)
1	0,056	20	1,28
2	"	"	1,59
3	"	"	1,67
5	"	"	1,73
10	"	"	2,53
20	"	"	2,79

Table 22 - (Data for Figure 27).

Ratio d_B/d_W (Depth Float Solids below Water Level/ Depth Float Solids above Water Level) for different Air/Solids Ratios, a_s.

a_s (mg/mg)	d_B/d_W (cm/cm)
0,010	10,61
0,0152	10,42
0,0172	14,01
0,0179	9,12
0,026	7,98
0,0318	7,47
0,036	6,45
0,040	5,45
0,041	6,94
0,0492	3,98
0,0655	4,21
0,066	5,57
0,072	2,93
0,082	1,59

LIST OF SYMBOLS

a_p = Mass of Air Precipitated per Litre of Pressure-Saturated Water (mg air/litre water).

P_g = (Gauge) Pressure in Saturation Unit (kPa).

v_L = Limiting Downflow Rate ($cm^3/cm^2/min$).

a_s = Air/Solids Ratio (mg/mg).

Q_s = Solids Loading Rate ($mg/cm^2/min$).

C_F = Float Solids Concentration (% by mass).

d_W = Depth of Float Solids Above the Water Level (cm).

d_B = Depth of Float Solids Below the Water Level (cm).

A_1 = Cross-Sectional Area of Upper Chamber (cm^2).

A_2 = Cross-Sectional Area at (Effluent) Discharge Point of Lower Chamber (cm^2).

Q_T = Total Hydraulic Flow into Flotation Unit (l/min).

Q_{sat} = Saturator Feed Rate (l/min).

Q_w = Wastewater Feed Rate (l/min).

C_i = Influent Wastewater Solids Concentration (mg/l).

C_e = Effluent Solids Concentration (mg/l).

P_a = Atmospheric Pressure (gm/cm^3).

P_h = Hydrostatic Pressure of Displaced Water (gm/cm^3).

ρ_{air} = Density of Air (mg/l).

V = Total Volume of Saturator Feed Required to Precipitate Sufficient Air to Displace 1 Litre of Water in Monitoring Unit (l).

T = Time taken to Displace One Litre of Water in Monitoring Unit (min).

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