The Significance of the Root Mean Square Velocity Gradient and Its Calculation in Devices for Water Treatment

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Abstract

This paper is aimed at reviewing calculations of the root mean square velocity gradient for devices most often used in water treatment practice for coagulation of water. In particular, it summarizes calculations of the loss of head produced by hydraulically operated devices, and the work per unit time dissipated by the agitator into mechanically operated devices in order to determine the magnitude of the mean velocity under which the system coagulates in the given device. Consideration is given to the following technological devices: hydraulic agitation including baffled, orifice, jet and ring water jump flash mixers, baffled flocculation chambers and sludge blanket; mechanical agitation including revolving and reciprocating agitators.

Introduction

The fundamental principle of the chemical treatment of water is the conversion of ion, molecular and colloidally dispersed impurities present, into rough dispersions which can be separated from the water by subsequent separation processes. The crux of the problem of water treatment is the removal of the colloidal dispersions which form the major portion of impurities, and which are kinetically and aggregationally stable. Destabilization of impurities comprises the removal of the energetic barrier of particles by the effect of coagulation reagents. This is the first phase of the water treatment process.

Immediately after the energetic barrier has been removed, the destabilized particles combine by mutual contact into multiple aggregates. The kinetics of the process of aggregation is derived from the intensity of particle movement. Aggregation is a stepwise process, and the partial phases of aggregate development are distinguished by the prevailing mechanism of movement. With increasing particle velocity the frequency of mutual contact between particles increases and consequently also the rate of aggregation.

From the kinetic point of view, the sequence of aggregate formation proceeds as follows: the destabilized particles combine *via* basic aggregates into primary particles which represent an intermediate stage to the formation of micro-particles, and micro-particles combine into macro-particles which should be the final product of particle aggregation.

The movement of particles in the water being treated is effected by the agitation of the water. As a rule the agitation stages are distinguished by three different agitation intensities, *viz.* rapid mixing aimed at achieving homogenisation of reagents with the water to be treated; rapid agitation aimed at the formation of micro-particles; and slow agitation aimed at the formation of macro-particles.

Velocity Gradient

The rate of aggregate formation and the size and structure of aggregates formed are primarily controlled by the intensity of agitation, since the collision of particles are effected by their different drag velocities. The drag velocity of particles depends on the velocity difference between the neighbouring layers of liquid. This difference in velocity across the liquid layer is called the instantaneous velocity gradient G in a flow, and its value expresses intensity of agitation at any point in the agitated system. The value of G, however, varies throughout the profile of an agitated system and is difficult to calculate. Therefore, Camp and Stein (1943) substituted the root mean square velocity gradient $\bar{G}$ for the instantaneous velocity gradient G, which they recommend should be calculated from the work per unit time put into a unit volume.

$$\bar{G} = \sqrt{\frac{W}{Vt\mu}} = \sqrt{\frac{P}{V\mu}} \qquad (1)$$

where

$\bar{G}$ = the root mean square velocity gradient (s^{-1})
W = work, dissipated energy ($m^2.kg\ s^{-2}$)
P = work per unit time, useful power input ($m^2.kg\ s^{-3}$)
V = volume (m^3)
t = effective retention time (s)
μ = dynamic viscosity ($m^{-1}\ kg\ s^{-1}$)

The velocity gradient $\bar{G}$ is, however, a guideline only for ascertaining the intensity of agitation. From the operational point of view it is important that the distribution of the velocity gradient throughout the total volume of the agitated vessel is uniform. Similarly, it is important that the period required for agitation corresponds to the retention time available in the vessel and is not decreased by inefficient hydraulic utilization of the agitated vessel.

The Significance of Agitation in Water Treatment Practice

The kinetics of the destabilization process is given by the speed of the partial phases occuring during destabilization. For destabilization to proceed quantitatively while ensuring maximum economy of chemical usage, the most important objective is the achievement of perfect homogenization, i.e. the instantaneous distribution of destabilizing reagent with the highest degree of pH uniformity and concentration uniformity of added reagent throughout the total volume of water to be treated. Under optimum conditions of homogenization the quantity of destabilizing reagent is in stochiometric ratio to the surface of particles. When homogenization occurs too slowly, the highly active hydrocomplexes mutually combine in the regions of higher concentrations of reagent. In this case, to achieve optimum destabilization the total volume of water to be treated will require higher reagent dosage than is predicted by the stochiometry since the reagent bound into hydrocomplexes does not actively affect the destabilization process.

The intensity of mixing, when applied for the purpose of homogenization should be as high as possible to achieve this in the shortest period. Constant and continuous addition of reagent is equally important for achieving effective homogenization while pulsating (intermittent) chemical dosing is detrimental.

It is recommended that the homogenization be performed in a unit which is separated from the subsequent technological stage in which micro-particles are formed under rapid agitation, because the character of the agitation in the two stages is different (macro- vs. micro-turbulence).

Rapid agitation in the technological phase of micro-particle formation is one of the most important stages in the process of water treatment. The conditions under which micro-particles are formed determine the final character of the suspension as well as the overall efficiency of the treatment process. The applied velocity gradient should be high ($\bar{G}$ = 100 — 500 s^{-1}) and the distribution of the velocity gradient should be as uniform as possible throughout the agitated volume of water. The duration of agitation depends on the quality of the water to be treated and of coagulation reagent used, and is usually from 1 — 5 min. Agitation at high $\bar{G}$ for too long a period, however, could lead to a decrease in and possible loss of the binding capability of the formed particles which is important for the subsequent formation of macro-particles.

Conditions of slow agitation are dependent on the type of suspension being formed because slow agitation determines the final character of the suspension to be separated. The result of slow agitation should be flocs with sufficiently high settling velocity. Uniformity of distribution of the velocity gradient throughout the volume of agitated water is equally important in the slow agitation phase, because breakdown of macro-particles occurs when hydrodynamic forces acting on the macro-particles, are greater than the adhesive forces acting between the micro-particles forming the macro-particles.

The efficiency of the slow agitation phase is derived from the magnitude of the velocity gradient and its duration. The value of the velocity gradient applicable is determined by the size (settleability) of floc required, because there is always a maximum size associated with each magnitude of velocity gradient. The value of the velocity gradient therefore must not exceed a critical limit beyond which floc breakdown occurs. The optimum $\bar{G}$ value for different conditions of flocculation is stated by various authors (Fair and Geyer, 1958; Hudson, 1957; Ives, 1968; Ritchie, 1955; Tesarik, 1968; and Walker, 1968).

In the current practice of aggregating micro- into macro-particles a value for $\bar{G}$ in the range of 15 — 30 s^{-1} is commonly considered optimal. The duration of slow agitation should be determined from the optimum value of $\bar{G}t$ for the given water and the coagulation reagent used. For Al coagulant the value of $\bar{G}t$ is in the order of 75 000, for Fe coagulant approximately 140 000, and for organic coagulant is in the range of 200 000 — 500 000. The optimum value for $\bar{G}t$ should however be experimentally determined for each particular installation.

Generally, for agitation aiming at the formation of well settleable flocs it is necessary to maintain the principle of a gradual decrease of velocity gradient. In the initial rapid agitation phase the water is agitated at high, usually constant $\bar{G}$, and in the subsequent slow agitation phase the water is agitated at a substantially lower $\bar{G}$. The optimum course of macro-particle formation requires a gradual decrease of the velocity gradient where the ideal is a smooth transfer from the velocity gradient applied in the rapid agitation phase to the magnitude of velocity gradient given by the hydraulics of the first separation stage (settling tank, clarifier).

Means of Agitation

Agitation is brought about by external action. The sources of power being dissipated into the individual stages for floc formation are most commonly gravitational or mechanical.

Gravitational systems (often called hydraulic or stream) are based on the creation of turbulent flow through a device. These can be designed as open gravity or in-line pressure arrangements. The intensity of agitation depends only on the total head loss produced by the device. It is directly proportional to the rate of flow, and cannot, therefore, be varied at will by the operator which makes the gravitational devices relatively inflexible. Therefore, gravitational flocculation chambers are seldom included in modern, sophisticated, large plants, even though they may possess some quite useful features such as simplicity of operation and the fact that they require no additional power.

In mechanically operated systems the intensity of agitation depends on the resistance of water acting against a moving body. The resistance of water is proportional to the speed of the impeller, and is independent of the rate of flow. The intensity of agitation is a function of the energy consumed to overcome the resistance of the water. It can be varied independently of flow conditions to meet the changes in raw water quality and temperature by changing the speed of the impeller.

When mechanical agitation is employed in a continuous system a hydraulic effect is also manifested. The final $\bar{G}$ then can be calculated from the simultaneous action of the work put into the system by the impeller and the work to overcome the hydraulic losses.

$$\bar{G} = \sqrt{\frac{P_M + P_H}{V\mu}} = \sqrt{\bar{G}_M^2 + \bar{G}_H^2} \qquad (2)$$

where

P_M, P_H = work per unit time put into the system by mechanical and hydraulic means;
$\bar{G}_M$, $\bar{G}_H$ = root mean square velocity gradient induced by mechanical and hydraulic means.

According to equation (2) only the velocity gradients which act simultaneously on the flowing liquid can be computed, but it cannot be used in cases where velocity gradients act sequentially. Where velocity gradients act sequentially the agitation effect can only be described in terms of the duration of action of the individual velocity gradients and their sequence, i.e. $\Sigma\,\bar{G}t$.

1. Agitation by Hydraulic Means

Agitation in gravitational devices is achieved at the expense of pressure head. The work per unit time depends on the total head loss Δh according to the relationship

$$P = Qg\,\varrho\Delta h \qquad (3)$$

where

Q = rate of flow ($m^3\,s^{-1}$)
Δh = total head loss (m)
g = gravity constant ($g = 9{,}81\ m\,s^{-2}$)
ϱ = water density (m^{-3} kg)

The total head loss Δh is produced by local losses Δh_l, and frictional resistance of flow Δh_f.

Combining equations (1) and (3) the general formula for the mean velocity gradient induced by gravitational devices becomes

$$\bar{G} = \sqrt{\frac{g\,\Delta h}{t\,\nu}} \qquad (4)$$

where ν = kinetic viscosity (m^2s^{-1}).

1.1 Flash Mixers

Hydraulic flash mixers are always preferred. The most suitable method of mixing is by means of inline pressure mixers with narrowing of the throughflow profile. In such arrangements a high intensity of turbulence at the point behind the narrowing of the profile is achieved. The reagent is added at a point immediately after the constriction. For large pipes it is recommended that multiple dosing points around the circumference be provided. The loss of head should be at least 0,5 M.W.C. In open troughs homogenization is accomplished by means of a Parshall flume.

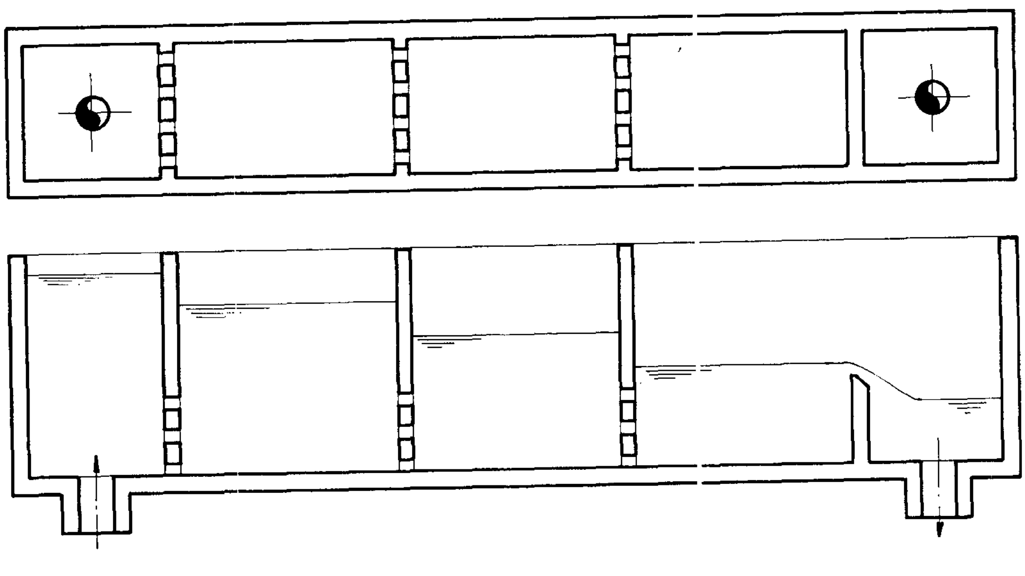

Figure 1
Perforated baffle flash mixer

There are a number of different flash mixers commonly used but only 4 basic arrangements will be analysed: a baffled, orifice, jet, and ring water jump mixer.

1.1.1 *Baffled Mixers*

Effective mixing of water with chemicals is achieved in mixers in which increase in turbulence occurs through impact of the water stream on the baffles inserted into a channel or pipeline. These can be arranged with baffles forming slots or built-in perforated baffles.

- Perforated Baffle Mixer

Two to three perforated baffles are usually used when only one chemical is dosed, and 3 to 4 perforated baffles are recommended when two chemicals are sequentially added. The mixer is usually sized to give a nominal velocity across the mixer v_m = 0,4–0,6 m/s with a velocity of $v_h \doteq$ 1,0 m/s in the holes. Effective mixing is achieved when the ratio L/d = 5–7, where L represents the distance between the two neighbouring baffles and d is the diameter of the holes. The recommended diameter of the holes is from 20–100 mm. The head loss produced by the flow through a perforated baffle is calculated in the same way as the loss where the flow is being discharged through a flooded opening in a vertical wall. For the n-number baffle arrangement the total head loss is

$$\triangle h = n \frac{v_h^2}{\mu^2 \, 2g} \doteq 0{,}10-0{,}13 \, n \, V_h^2 \qquad (5)$$

where μ = discharge coefficient; μ = 0,62–0,70.

In open gravity arrangements the holes should be positioned at least 100 mm below water level to prevent entraining the air in the water (Fig. 1).

- Slot Mixer

Baffles forming the slots are usually inserted at 45° to the direction of flow. 3–5 Baffles are enough to achieve thorough mixing. The mixer is usually sized for a nominal velocity across the mixer, v_m = 0,4–0,6 m/s, with a velocity of v_s = 0,8–1,0 m/s in the slot. The head loss produced by the slot is calculated as a loss caused by the change in the direction of flow. For n-number slot arrangement the total head loss becomes

$$\triangle h = n \, \xi \, \frac{v_s^2}{2g} \qquad (6)$$

where ξ = drag coefficient; it was experimentally determined that ξ = 2,0 – 2,5 for a mixer with baffles inclined at 45° to the direction of flow, and ξ = 3,0 – 3,5 for reverse flow.

In an open gravity arrangement (Figure 2) the slots should be vertically positioned. The water level drops behind each slot for its head loss $\triangle h_2$. Therefore, to maintain the same velocity, v_s in each slot, the width of the slots has to widen in the direction of the flow to give the same effective area of the slots A_s. The width b of n-slot is given by the relationship

$$b_n = \frac{A_s}{H_e + n \, \triangle h_s} \qquad (7)$$

where

A_s = effective area of a slot (m^2)
b_n = width of the n-slot counted from the exit side of the mixer (m)
H_e = water depth behind the last baffle (m)
n = number of slots counted from the exit side of the mixer.

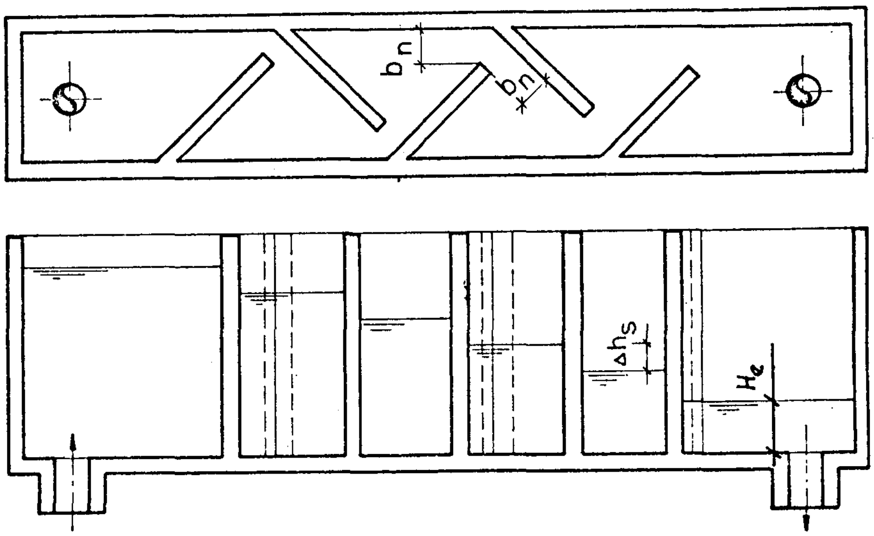

Figure 2
Slot (round-end) flash mixer

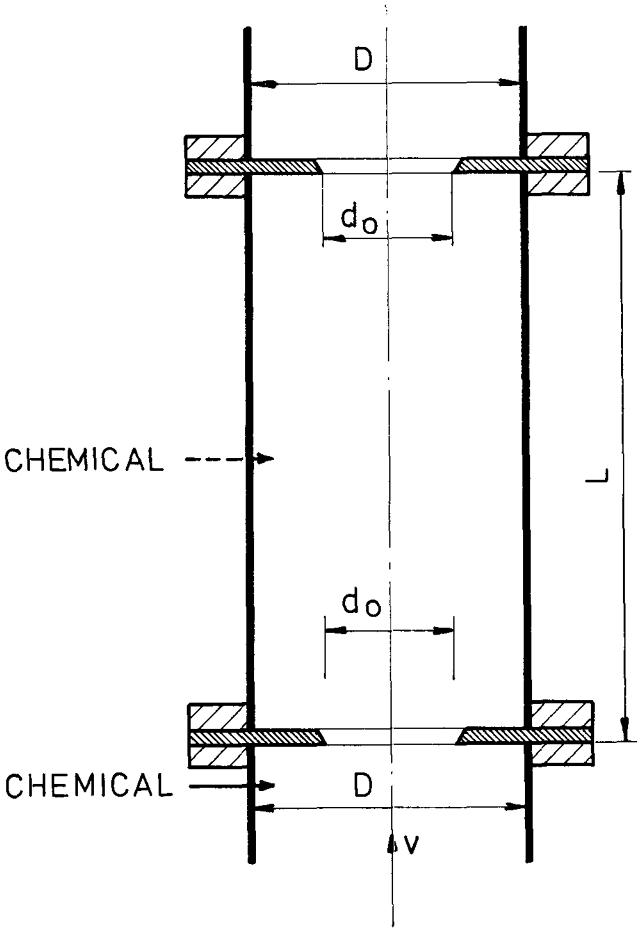

Figure 3
Orifice flash mixer

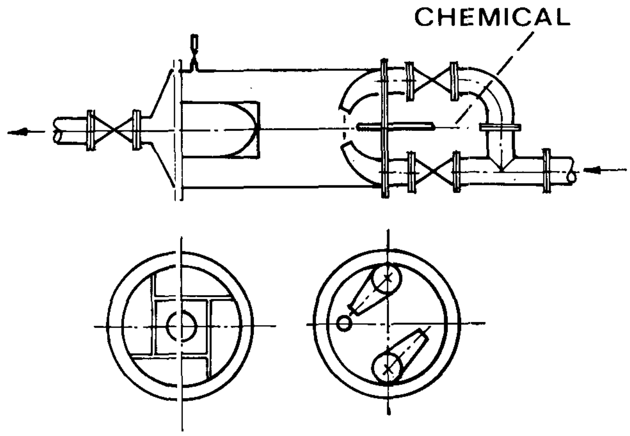

Figure 4
Jet stream flash mixer

1.1.2 *Orifice Mixer*

The orifice arrangement is the simplest and most common pressure (in-line) type of flash mixer, even though not the most effective. Mixing is achieved by high turbulence and eddies formed behind the orifice plate. The head loss of an orifice plate with sharpened edges for Re > 10^5 is

$$\triangle h_s = \frac{v^2}{2g} \left(1 + 0{,}707 \sqrt{1 - \frac{d_o^2}{D^2}} - \frac{d_o^2}{D^2}\right)^2 \qquad (8)$$

where

v = velocity of water in pipe (m s^{-1})
d_o = diameter of orifice hole (m)
D = pipe diameter (m)

The mixing effect can be significantly improved by an arrangement of two (Figure 3) or more identical orifices. Such an arrangement is also often used when more chemicals are sequentially added. The most effective mixing is achieved when ratio $L/d_0 = 5-7$, where L is the distance between the two neighbouring orifice plates. The total head loss of an n-number orifice arrangement is

$$\triangle h_n = n \triangle h_s \qquad (9)$$

1.1.3 *Jet Mixer*

The jet mixer (Figure 4) consists of a wide cylinder, put into the raw water feed main, to which water is fed by two tangentially connected jets, and the chemical is introduced in the centre of the mixer between the jets. Baffles inserted in the cylinder on its discharge side, counter the circular motion of water. The head loss is determined as the loss in the flow being discharged through a flooded opening according to equation (5) where v_h is the velocity in the jet and n = 1.

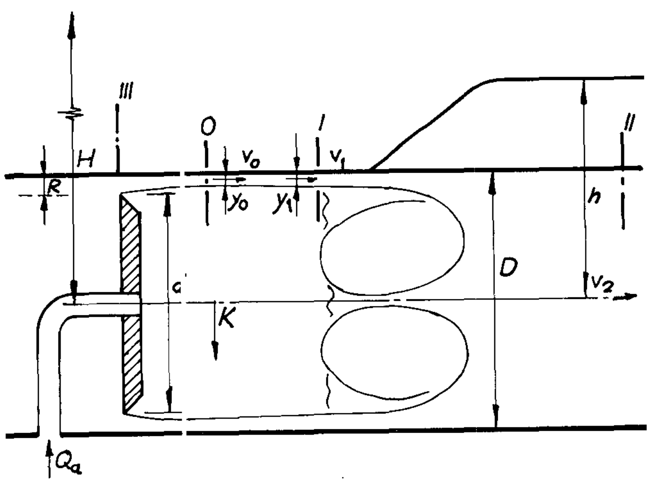

Figure 5
Ring water jump flash mixer

1.1.4 *Ring Water Jump Mixer*

Very effective mixing of chemicals and gases with water is achieved by the ring water jump mixer (Figure 5; Haindl, 1975).

The ring jump originates at the flow around shaped bodies inserted in the pipeline. Streams which originate behind the flow-around body along the pipe wall and which give the water a ring shape, pass into the flow through a full pipe profile by a jump characterized by high turbulence and formation of eddies. The temporary effect of the water jump forces gas or chemicals (magnitude Q_ρ in Figure 5) into the pressure flow behind the jump and by means of eddies all components are thoroughly mixed. It is a condition for the establishment of the ring water jump that the head h behind the temporary effect should equal the dynamic head of the jump Θ

$$\Theta = h = \frac{8}{D^2}\alpha_1\varphi(H+K)\mu R(D-R) - \frac{32}{D^4}\alpha_2\mu^2R^2(D-R)^2(H+K)$$
$$(1+\beta_1\epsilon_1) - K \tag{10}$$

where

H = pressure head at inlet pipe including velocity head (m)
h = pressure head behind temporary effect at the flow through a full profile (m)
D = diameter of pipe (m)
R = width of circular ring (m)
K = suction head (m)
α_1, α_2 = Boussinesq's coefficient of quantity of motion in inlet and outlet profile
$\beta = \frac{Q_\rho}{Q}$ = Q_ρ represents volumetric flow of added component at the pressure in a nucleus of ring stream; Q is the water flow rate
ϵ_1 = compression coefficient, $\epsilon = \frac{h_b - K}{h_b + h}$; h_b is the atmospheric pressure head (m)
μ = coefficient of contraction
φ = coefficient of change in velocity v_1 against v_o, $v_1 = \varphi v_o$; $\varphi < 1$

The head loss produced by the ring jump

$$\triangle h = H - (h + \frac{v_2^2}{2g}) \tag{11}$$

where

$$v_2 = \frac{4\,Q}{\pi\,D^2}\,(1 + \beta\epsilon) \tag{12}$$

Combining equations (10), (11) and (12) the total head loss produced by the ring jump becomes

$$\triangle h = H - 8\alpha_1\varphi\mu(H+K)\frac{R}{D}(1-\frac{R}{D}) + 16\mu^2(1+\beta_1\epsilon_1)(\frac{R}{D})^2(1-\frac{R}{D})^2$$
$$(H+K)[2\alpha_2-(1+\beta_1\epsilon_1)]+K \tag{13}$$

1.2 Agitation Arrangements

As in the case of the stream flash mixers, the hydraulically agitated flocculation tanks achieve the stirring action by the swirling motion of water at the expense of pressure head. Therefore, the work per unit time dissipated into the flocculating system should be expressed in terms of equation (3), and the value of $\bar{G}$ induced, under which aggregation takes place by equation (4).

The most commonly used agitation arrangement is a baffled channel arrangement. This is usually of a rectangular base-plan divided by baffles into channels in which water meanders either horizontally or vertically with a change in direction of flow of 180°. The channels with horizontal flow are the so called around-the-end channels, and with vertical flow the so called over-and-under channels.

The sludge blanket is also included among the agitation arrangements. The sludge blanket is considered to be an integral part of the flocculation process (Ives, 1968; Tesarik, 1968) in which aggregation proceeds at a rate as much as 100 times higher than in the flocculation channels due to a high concentration of floc particles (Hudson, 1965).

1.2.1 *Baffled Channels*

Baffled channels differ from unobstructed open channels in that turbulence is not merely a function of frictional resistance to flow. Turbulence is intensified by enforced changes in the direction of flow. Because of this the baffled channels are not very satisfactory for floc formation. The velocity gradient is intensified at points of enforced changes in the direction of flow and is not high enough in the straight channels.

The total head loss required to achieve good floc formation in the baffled channel arrangement commonly lies between $\triangle h$ = 0,3 – 0,8 m, depending on the quality of raw water and the kind of coagulant used.

To induce the required constant $\bar{G}$, the baffled channel of the same effective area A of each of the channels should be designed to produce the total head loss

$$\triangle h = \frac{\bar{G}\,L\,\nu}{g\,v} = \frac{\bar{G}^2\,t\,\nu}{g} \tag{14}$$

where

L = total length of all channels (m)
v = velocity of water in channel; the recommended optimum velocities for slow agitation are 0,25; 0,30; and 0,4–0,5 m/s for low, average and heavily turbid waters
t = retention time (varies from 10–60 min) (s)

It is, however, recommended that the velocity gradient gradually be decreased along the total length of the channels.

- Around-the-end Channel (Figure 6)

The total head loss $\triangle h$ is a sum of local losses (caused by a change in direction of flow) $\triangle h_t$, and frictional resistance of flow in the channels $\triangle h_f$

$$\triangle h = \triangle h_t + \triangle h_f \tag{15}$$

The head loss $\triangle h_t$ is estimated according to equation (6) in which the drag coefficient $\xi \doteq 3,0$ for 180° change in direction of flow, and n is the number of changes in the direction of flow.

The head loss $\triangle h_f$ for turbulent flow in a channel of total length L is derived from Chezy's equation

$$v = C\sqrt{R_h I} = C\sqrt{R_h\,\frac{\triangle h_f}{L}} \tag{16}$$

By solving the equation (16)

$$\triangle h_f = \frac{L\,v^2}{C^2 R_h} \tag{17}$$

where

R_h = hydraulic radius (m)
I = channel gradient (m m^{-1})
C = velocity coefficient ($m^{1/2}$ s^{-1})

According to Bazin, the velocity coefficient C for water equals

$$C = \frac{87\sqrt{R_h}}{\gamma + \sqrt{R_h}} \tag{18}$$

where

γ = factor dependent on the quality of the channel surface; for very smooth walls (smooth cement rendering) γ = 0,06, and for smooth walls (plain concrete) γ = 0,16.

Substituting

$$2\,g\,(\frac{\gamma + R_h}{87\,R_h})^2 = K \tag{19}$$

and combining equations (17), (18) and (19), the head loss

$$\Delta h_f = \frac{v^2}{2g} K\,L \tag{20}$$

Combining equations (6) and (20) the total head produced by the around-the-end baffled channel is

$$\Delta h = \frac{v}{2g}(n\,\xi + K\,L) \tag{21}$$

Introducing equation (21) into equation (3), the mean velocity gradient induced in the around-the-end baffled channel becomes

$$\bar{G} = 0{,}707\,v\sqrt{\frac{n\,\xi + K\,L}{t\,\nu}} \tag{22}$$

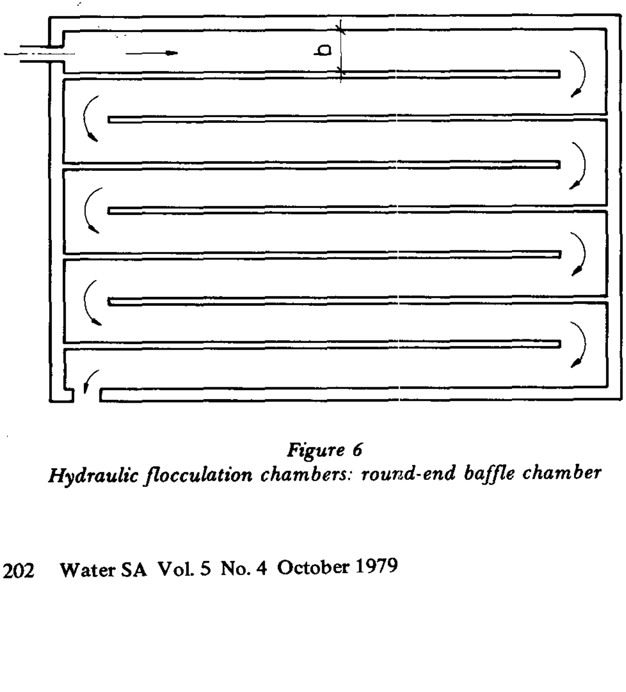

Figure 6
Hydraulic flocculation chambers: round-end baffle chamber

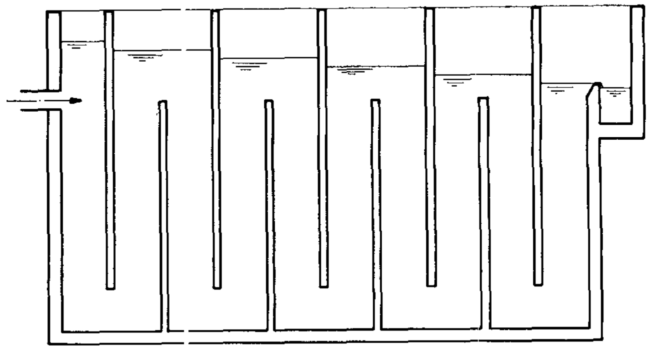

Figure 7
Hydraulic flocculation chambers: over-and-under baffle chamber

● Over-and-under Baffled Channel (Figure 7)

The total head loss produced by the over-and-under baffle arrangement is a sum of all losses calculated according to equation (15). The head loss caused by local losses Δh_l, is estimated as a sum of losses produced by the over and under flowing through the baffles. The head loss with under-flowing through the baffle is calculated as a loss by the flow being discharged through a flooded opening. The head loss with flowing over the baffle is calculated as a loss at the flow over the flooded weir. The loss of head by friction in a vertical channel is calculated in accordance with equation (17).

1.2.2 *Sludge Blanket*

The sludge blanket is a fluidised layer of floc particles. Fluidising is an effect caused by the upward flow of water through a layer of freely deposited particles. The flow through the sludge blanket is considered as flow through a porous, discontinuous bed of solid particles characterized by continuous inflowing of fresh suspension and continuous removal of excess suspension, with a state of equilibrium being maintained at all times.

Fluidising occurs as soon as the weight of the layer decreased by its buoyancy is approximately equal to the resistance offered by the layer to the water flow. Consequently, the fluidizing occurs when a certain minimum upward velocity is reached. This minimum velocity depends on the physical properties of particles and water. When the particles inside the layer are approximately evenly distributed, i.e. water flows around each single particle, it represents full fluidisation. Full fluidisation is achieved hydraulically, only by means of upward velocity and high turbulence at the inlet to the fluidised bed.

The fluidised layer tends to settle due to forces of gravity and because of higher density of the fluidised bed than that of water. Therefore the concentration of floc particles in the fluidised bed depends on the upward velocity. The fluidised bed expands when upward velocity increases, i.e. its porosity increases and resistance decreases, and *vice versa*.

When a layer formed by solid particles is fluidised, the difference between the weight of the layer, decreased by its buoyancy and resistance offered by the layer to the water flow, is caused by turbulent fluctuations which assist in fluidising the layer. Therefore, the actual resistance of the fluidised layer will always be smaller than the resistance D of the weight of the layer, decreased by its buoyancy (Tesarik 1971). The resistance D of the sludge blanket is given by the equation

$$D = A\,g\,\varrho\,\triangle h_s = A\,g(\varrho_p - \varrho)C_v h \tag{23}$$

where

D = resistance of the fluidised layer (m.kg s^{-2})
A = area of the sludge blanket level (m^2)
C_v = volumetric concentration of the floc particles throughout a depth of the sludge blanket ($m^3\ m^{-3}$)
h = depth of sludge blanket (m)
$\triangle h_s$ = loss of head through the fluidised layer (m)
ϱ_p = density of floc particle (average density of hydroxide floc particle ϱ_p = 1,003 kg/m^3) (m^{-3} kg)
ϱ = density of water (m^{-3}.kg)

The head loss by the flow through the fluidised layer is derived by solving equation (23)

$$\triangle h_s = \frac{\varrho_p - \varrho}{\varrho} C_v h \tag{24}$$

It is often difficult to determine the floc density ϱ_p. Therefore, it is substituted by the density of suspension of the fluidised layer ϱ_s. According to Minc (1964) the density of suspension ϱ_s may be expressed in terms of floc density ϱ_p, of which the fluidised layer consists, and its volumetric concentration C_v by the relationship

$$\varrho_s = C_v\,\varrho_p + \varrho(1-C_v) \tag{25}$$

By combining equations (24) and (25) the head loss through the fluidised layer, in terms of density of suspension ϱ_s, becomes

$$\triangle h_s = \frac{\varrho_s - \varrho}{\varrho} h \tag{26}$$

The velocity gradient $\bar{G}$ induced by the sludge blanket is a function of the head loss through the fluidised bed $\triangle h_s$ and the head loss by local losses $\triangle h_t$, when the head loss by frictional resistance of flow $\triangle h_f$ is regarded as negligible in comparison with the above losses. The head loss by local losses $\triangle h_t$ is dependent on the shape of the vessel of the sludge blanket, inlet arrangement, inlet velocity and flow distribution across the sludge blanket. It is a sum of all local losses $\triangle h_i$ proportional to the respective velocity heads

$$\triangle h_t = \Sigma \triangle h_i = \Sigma\, \xi_i \frac{v_i^2}{2g} \tag{27}$$

The total head loss $\triangle h$ of the sludge blanket is relatively very small, of the order of a few millimeters (Tesarik, 1968). It was proved (Ives, 1968; Tesarik, 1968; and Tesarik, 1971) that the head loss through the resistance of the fluidised layer $\triangle h_s$ is far greater than the head loss through the local losses $\triangle h_t$ and that the head loss $\triangle h_t$ can in most cases be neglected. In such a case the total head loss $\triangle h$ corresponds to the head loss through resistance of the fluidised layer, i.e. $\triangle h = \triangle h_s$, and the mean velocity gradient induced by the sludge blanket becomes

$$\bar{G} = \sqrt{(\varrho_s - \varrho)\,h\,\frac{g}{t\mu}} \tag{28}$$

2. Agitation by Mechanical Means

Agitation by mechanical means is caused by the motion of a solid body, the impeller (paddle, blade, propeller) submerged in the water. In view of hydrodynamics the motion of the impeller is considered as a flow-around the solid body by the water stream. The water offers a resistance to the moving impeller characterized by the drag force. The total drag force F_p is produced by the force of frictional resistance of flow originating in the boundary layer, and the force of pressure connected with formation of eddies. In the eddy region there is a lower pressure than that acting on the front of the impeller. This difference in pressure $\triangle p$ has to be overcome by forces driving the impeller.

2.1 Flash Mixers

Mechanically operated flash mixers should be used only under exceptional circumstances such as when the head available is limited. The velocity gradient applied should be high ($\bar{G}$ = 500 – 1 000 s^{-1} or greater) and effective homogenisation should be completed within 1–5 s.

2.2 Agitation Arrangements

Mechanical agitators in flocculation tanks are used basically in 3 different arrangements:

- Revolving agitators with paddles parallel to the shaft (Figure 8), causing movement in a vertical plane (agitator with horizontal shaft) or in a horizontal plane (agitator with vertical shaft)
- Revolving agitators with horizontal paddles in the direction of the radial arm, causing movement in a horizontal plane (Figure 9)
- Reciprocating agitators, causing movement in a vertical plane (Figure 10), or in a horizontal plane (Figure 11)

2.2.1 *Revolving Agitators*

The useful power input is a function of the drag force of the impeller F_D and the distance s moved in a unit of time. The distance s moved per second is equal to the relative velocity of the impeller with respect to water v_r. The relative velocity v_r is estimated as a differential between the peripheral velocity of the impeller v_i and the velocity of water v_w surrounding the impeller

$$s = v_r - v_i - v_w \tag{29}$$

where

v_r = relative velocity of impeller with respect to water (m s^{-1})
v_i = peripheral velocity of impeller (m s^{-1})
v_w = velocity of water surrounding the impeller (m s^{-1})

At the turbulent flow-around the impeller the drag force F_D is estimated according to Newton's law

$$F_D = A\,\triangle p = A\,\xi\,\frac{\varrho\, v_r^2}{2} \tag{30}$$

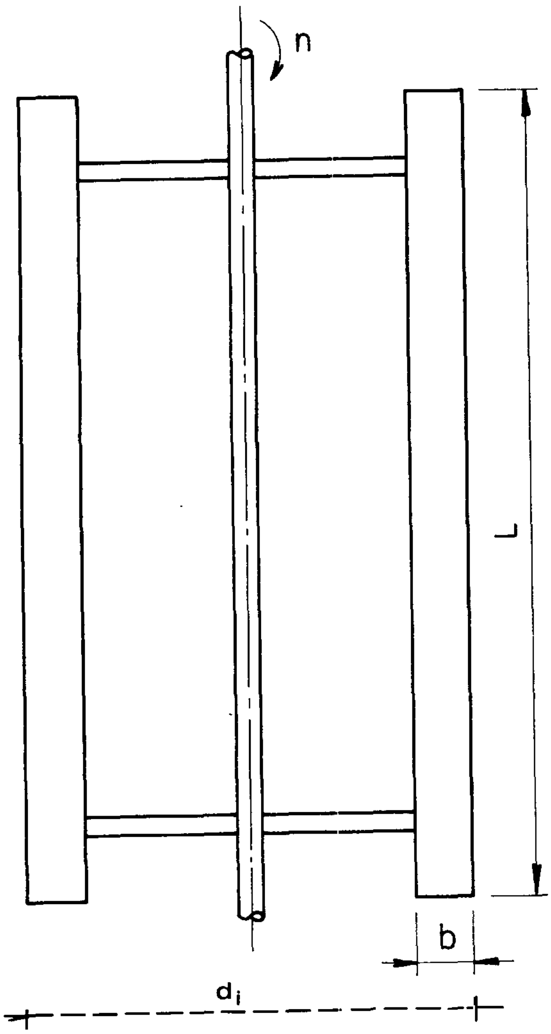

Figure 8
Revolving agitator with paddles parallel to shaft

and the useful power input equals

$$P = F_D s = F_D v_r = A\,\xi\,\frac{\varrho\, v_r^3}{2} \qquad (31)$$

where

F_D = drag force (m.kg s^{-2})
$\triangle p$ = difference in pressure (m^{-1} kg s^{-2})
A = submerged area of impeller perpendicular to direction of motion (m^2)
ξ = drag coefficient;
ϱ = density of water (m^{-3} kg)

The relative velocity of impeller v_r is not constant and depends on the peripheral velocity of the impeller v_i. If k is the ratio of the water velocity v_w and the impeller velocity v_i, the relative velocity of the impeller with respect to water is v_i

$$v_r = v_i - kv_i = 2\pi r(1-k)n \qquad (32)$$

where

r = effective radius arm of the paddle (m)
n = number of revolutions per second (s^{-1})

In these terms the useful power input according to equation (31) for a single paddle becomes

$$P \doteq 124\,\xi\,\varrho\,(1-k)^3 n^3 r^3 A \qquad (33)$$

For an impeller with paddles parallel to the shaft (Figure 8) the peripheral velocity of the impeller v_i is usually designed for flocculation tanks in the range v_i = 0,3—1,0 m/s. For this range of v_i the average value of the ratio k $\doteq$ 0,25, and the relative velocity of the impeller becomes

$$v_r \doteq 0{,}75\, v_i = 1{,}5\,\pi\, r\, n \qquad (34)$$

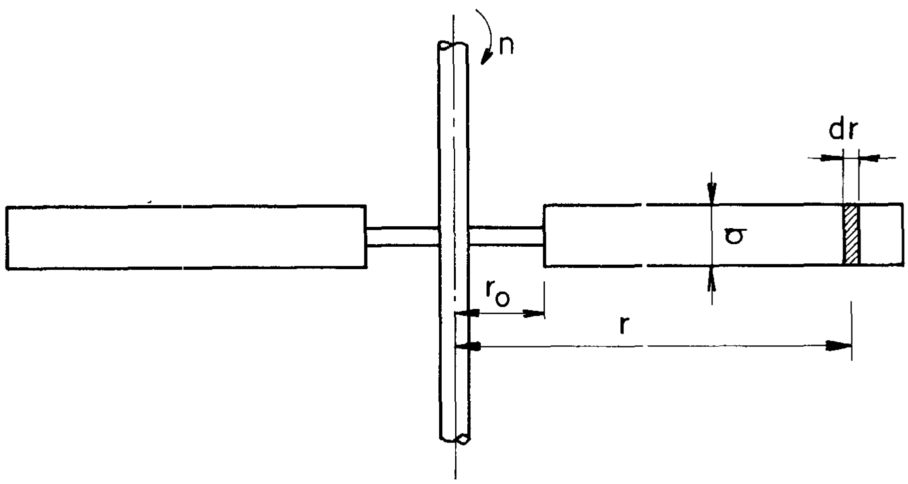

Figure 9
Revolving agitator with paddles in direction of radial arm (paddles perpendicular to shaft)

The useful power input P is estimated according to equation (33) in which drag coefficient ξ is replaced by the shape factor K. For Reynold's number Re $< 5.10^5$ Blasius determined the value of K as a function of Reynold's number

$$K = \frac{1,327}{\sqrt{Re}} \quad \text{where } Re = \frac{v_r\, b}{\nu} \tag{35}$$

where b = height of paddle perpendicular to direction of motion m.

The value of K however varies with the shape of the paddle. Experimentally determined values of K for different ratios of paddle length L to height b and Re $> 10^3$ are introduced in Table 1 (Konicek and Kroupa, 1956).

TABLE 1
K-VALUES FOR DIFFERENT RATIOS OF PADDLE LENGTH (L) TO HEIGHT (b)

L/b	1	1—2	2,5—4	4,5—10	10,5—18	18—
K	1,1	1,15	1,19	1,29	1,4	2,0

If the impeller includes a series of paddles then the magnitude of r^3A in equation (33) is replaced by Σr^3A, which is a sum of values r^3A for all paddles.

In the above terms the useful power input is

$$P = 52{,}3\, K\, \varrho\, n^3 \Sigma r^3 A \tag{36}$$

and the mean velocity gradient induced by an impeller becomes

$$\bar{G} \doteq 7{,}23 \sqrt{\frac{K\, n^3}{V\, \nu} \Sigma r^3 A} \tag{37}$$

The total area of paddles in a tank without stator blades should not be greater than 15—20% of the transverse cross-section area of the tank. If the area of paddles is 20% or even more, rolling of water results to the detriment of effective agitation.

For impellers (Figure 9) with a dimension of A which is substantial in the direction of the radius arm and with the constant height of paddle b in the plane perpendicular to the direction of motion the expression r^3A in equation (33) becomes

$$r^3A = \int_{r_o}^{r} r^3\, dA = b\int_{r_o}^{r} r^3\, dr = \frac{b(r^4 - r_o^4)}{4} \tag{38}$$

because dA = b dr

where r, r_o = demarcating radii of the paddle (m)

In these terms equation (33) becomes

$$P \doteq 31\, \xi\, \varrho (1-k)^3 n^3 b (r^4 - r_o^4) \tag{39}$$

Effective control of velocity gradient requires a definite relation between the power dissipation and the velocity of the paddle with respect to the tank which must be proportional to its distance from the shaft. In the case of agitators with paddles in the direction of the radial arm, the peripheral velocity of water v_w is indirectly proportional to the paddle length from the shaft, while the peripheral velocity of paddle v_i is directly proportional to this distance. The differential between the peripheral velocities of paddle and water increases with increasing length of paddle arm. The degree of turbulence, thereby, also increases with a resulting non-uniformity of the velocity gradient along the paddle. This effect is unfavourable in view of floc particle growth, and this type of impeller therefore cannot be recommended for agitation in flocculation tanks.

An estimate of the useful power input based on the relative velocity of an impeller is inaccurate since the velocity of water is not determined accurately. Accurate methods for calculating the useful power input are developed from the theory of hydrodynamic similarity. All these methods are based on direct or indirect measurement of torque T in pilot scale equipment, and the results are presented in the form of a dimensionless criterion.

Agitation in the flocculation tanks usually takes place without creating waves. For such conditions of agitation, resistance of the water is characterized with sufficient accuracy only by the Euler criterion of hydrodynamic similarity. It can be obtained by solving equation (30), in which relative velocity v_r is replaced by the peripheral velocity of the impeller v_i.

$$\xi = Eu = 2\, \frac{\Delta p}{\varrho\, v_i^2} \tag{40}$$

The useful power input P is incorporated in the Euler (power) number defined as

$$Eu = \frac{P}{d_i^5 n^3\, \varrho} \quad \text{where } P = 2\,\pi\, T\, n \tag{41}$$

where

d_i = diameter of impeller (m)
T = torque (m^2 kg s^{-2})
n = number of revolutions (s^{-1})

The Euler number is a function of Reynold's number modified for conditions of agitation

$$Eu = f(Re) = \frac{C}{Re^m} \quad \text{where } Re = \frac{n\, d_i^2}{\nu} \tag{42}$$

where C,m = factors experimentally determined for each individual type of impeller, and are valid as long as constant geometric ratios of both impeller and vessel dimensions are maintained.

The mean velocity gradient induced by the revolving agitator

$$\bar{G} = \sqrt{\frac{Eu\, d_i^5\, n^3}{V\, \gamma}} = \sqrt{\frac{C\, d_i^5\, n^3}{Re^m\, V\, \gamma}} \tag{43}$$

2.2.2 *Reciprocating Agitators*

The agitators with reciprocating motion are the so-called "walking beam" type, causing movement in a vertical plane (Figure 10), and the "pendulum" type, causing movement in a horizontal plane (Figure 11). The reciprocating motion may be obtained as in the case of a simple harmonic motion caused by the wheel on a shaft. It is characteristic for any type of reciprocating agitator that its relative velocity of blade/paddle changes continuously throughout the stroke. Due to forces of inertia in the liquid, the power input increases when the agitator is reversing its stroke. In order to calculate the mean velocity gradient it is, therefore, necessary to estimate the mean value of useful power input throughout the cycle. It can be calculated from equation (31), in which the relative velocity v_r is expressed in terms of peripheral velocity of harmonic motion per cycle $v_h = 2\ rn$, and distance moved per second as the mean feed rate per cycle $v_f = 4\ rn$

$$P = 8\,\pi^2\,\xi\,\varrho\,n^3k^3\Sigma\,r^3A \qquad (44)$$

where

v_h = peripheral velocity of harmonic motion (m s^{-1})
v_f = mean feed rate throughout the cycle (m s^{-1})
r = eccenter radius of the driving mechanism (m)
k = ratio of arm length of a double-armed lever mechanism

For the "walking beam" type with flat blades, Camp (1953) states the value of drag coefficient $\xi = 3{,}0$. For "pendulum" type the value of ξ may be determined as the value of shape factor K according to equation (35) or Table 1.

From the viewpoint of floc formation it is desirable to prevent an increase in the intensity of stirring in the direction of flow in order to prevent the disruption of flocs being formed. Based on analysis of $\bar{G}$ values for individual parts of the flocculation vessel in a diffuser shape, Tesarik and Vostrcil, (1976) have proposed the distribution of the paddles along the height of the "pendulum" agitator according to the power relationship

$$\frac{X}{h} = 1 - \log \frac{10\ r}{h} \qquad (45)$$

where

X = distance from the centre of gravity of paddle to the beginning of diffuser (m)
h = vertical length of agitator measured from the beginning of diffuser (m)
r = distance from centre of eddy (m)

The proposed logarithmic distribution of paddles along the height of the agitator induces an approximately uniform $\bar{G}$ value throughout the profile of agitated vessel, if paddles of the same height b are used.

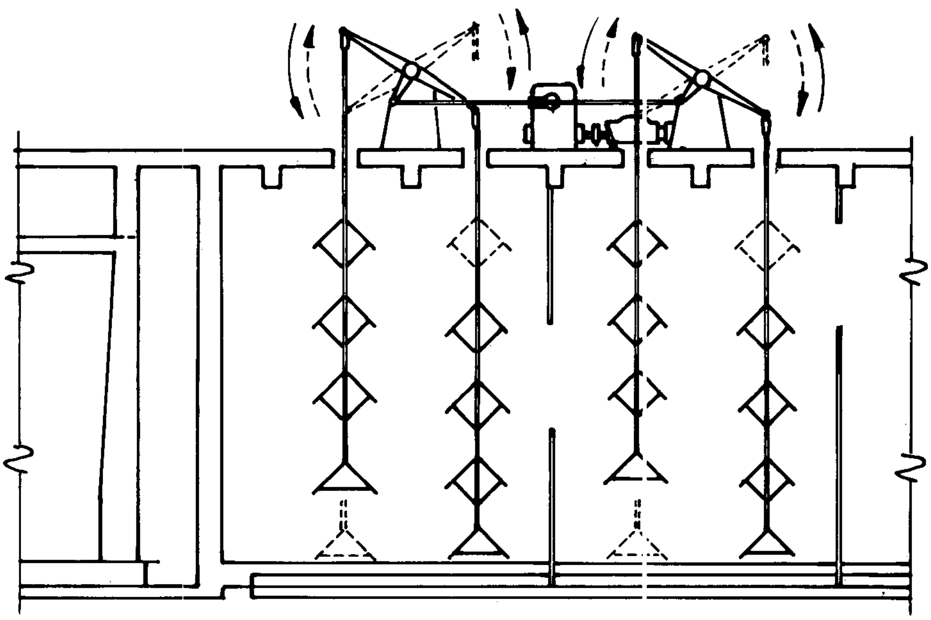

Figure 10
Reciprocating agitator with vertical motion — "walking beam"

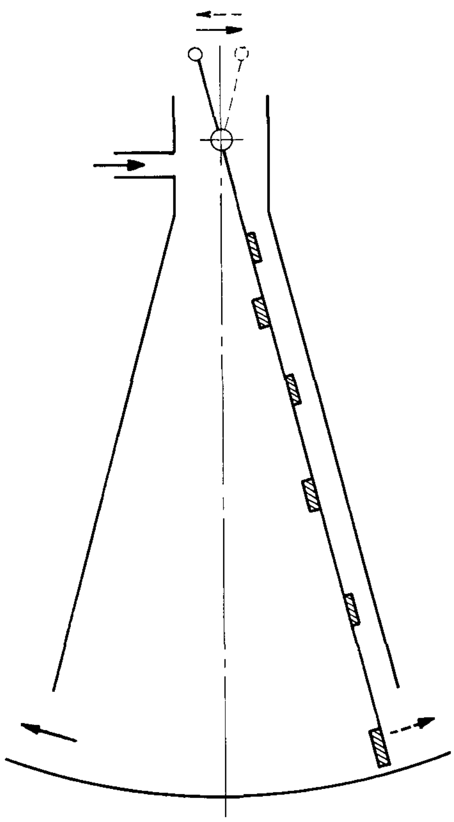

Figure 11
Reciprocating agitator with horizontal motion — "pendulum"

Conclusions

Intensity of agitation in gravitational devices is directly proportional to the rate of flow. Therefore, to achieve satisfactory aggregation of water, these devices can be recommended only when the nominal design flow can be maintained at all times and if there is no great variation in raw water quality and temperature. Furthermore, from the analysis of losses throughout the baffled channels, it follows that this arrangement is not very satisfactory for floc formation.

Calculation of the useful power input of mechanical agitators is more accurate when estimated from the Euler criterion of hydrodynamic similarity, rather than from the relative velocity of impeller with respect to water.

An impeller with paddles in the direction of the radial arm is not suitable as a flocculation agitator because the differential between the peripheral velocities of the paddles and the water is not constant but increases with increasing length of paddle arm.

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