Reprinted from AIChE SYMPOSIUM SERIES, Volume 68, Number 122, 1972

MIXING TECHNIQUES FOR THE CONTROL OF WASTE SLUDGES AND SLURRIES

J. Y. Oldshue

The anaerobic digestion of solids from clarifying, filtering and centrifuging steps is one method of waste solids disposal. Impeller mixers can be very effective in promoting the anaerobic reactions in the digestors. Various mixing principles that govern the agitation of thin and thick biologically active sludges are presented.

Mixing is involved in using slurries effectively to treat other types of waste, and also in the application of gaseous, liquid, and solid reagents to a variety of waste streams.

Solid waste materials come in a variety of forms. When the solids are particulate and suspended by either liquids or gases, impeller type mixers can be used to carry out various processing steps. A variety of these process possibilities will serve to survey the general field.

FLOWING LIQUID-SOLID WASTE STREAMS

The common way to treat a stream containing solid waste material is to pass the slurry through a filter or a settling device. The underflow is a relatively thick sludge stream. This sludge often contains a high percentage of organic material and is often treated by anaerobic digestion.

The use of impeller type mixers can have several beneficial effects. Top-entering units are used to break up the scum layer on the surface which has a tendency to inhibit the release of gases from the anaerobic decomposition. Dramatic effects are often produced by the use of only top-entering mixers. The top-entering mixers are installed similar to those shown in Fig. 1.

Side-entering mixers, Fig. 2, are often used to give motion to the entire slurry in the tank.

The recycling of digester gas for sludge mixing is relatively ineffective. On a general time scale, an anaerobic digester can operate at three times the rating of a completely unagitated digester. For example, let's say that it would take 30 days to produce 80% decomposition of a particular sludge in a completely unagitated unit. By using gas mixing this time might be reduced to 25 days. By using thorough mixing of the contents as well as removing the scum layer, this time could be reduced to 10 days.

Top-entering mixers can be used with one impeller to eliminate the scum, and a lower impeller to give circulation throughout the mass, Fig. 3. At a similar impeller diameter-to-tank diameter ratio, D/T, side-entering mixers require less horsepower than that required by top-entering mixers for typical blending applications. It is, therefore, quite common to use a top-entering mixer as shown in Fig. 1 to eliminate the scum layer and a side-entering mixer, Fig. 2, to mix the bulk of the material.

With a residence time of 10 days, there is no reason to run the mixers continuously. Usually they are operated one or two hours a day and also during the time the feed pump is operating if the tank is being intermittently fed.

If the top-entering mixer were used only for scum breakup, it would probably give a retention time on the order of 20 days in the example above. In one case, the installation of top-entering scum mixers in a given plant increased gas production from 40,000 cubic feet per day to 70,000 cubic feet per day which is almost in the ratio of 30 to 20 days in the above example.

The optimum tank size, in terms of mixer horsepower, can be predicted from Fig. 4. The comparison between top-entering and side-entering agitators for this kind of service is shown in Figs. 5 and 6.

Fig. 7 shows that at liquid-level to tank diamter ratios (Z/T) below .6, the surface is moved before the bottom is completely swept, while at Z/T's above .6, the surface will not move until extra power above that required for bottom motion is added.

Another interesting application is the use of mixers to smooth out sludge properties coming to an anaerobic digester. Fig. 8 illustrates one of the largest mixers ever manufactured which is producing uniformity in the sludge going to the rotary vacuum filters.

Fig. 9 shows how a sinusoidal feed stream can be damped in going through a properly designed mixing vessel. The damping factor is a single sinusoidal wave.

Fig. 10 illustrates the damping of a more complex feed stream where apparent sludge viscosity is used as a criterion to evaluate sludge properties.

For example, if the sludge viscosity varies from 10 to 1000 centipoises at 5 seconds^{-1} over a 24-hour cycle, a properly agitated tank with 12-15 hours retention time can reduce the apparent viscosity to viscosities of 100-200 centipoises.

AERATION OF THICK SLUDGES

Any required oxygen uptake rate can be maintained by properly designing submerged turbine impeller systems.

Fig. 11 illustrates a typical submerged turbine aerator and submerged air supply. These units can be of any required horsepower and in the fermentation industry have been used to provide uptake rates as high as 5000 mg per liter per hour. The limiting factor is the maximum practical concentration of sludge from the underflow from the filters or clarifiers. Slurries can have the consistency of applesauce and still be effectively agitated and aerated.

Fig. 12 shows typical uptake rates for various air rates in a 20-ft. deep tank. Fig. 13 shows the same sort of data in a 40-ft. deep tank.

Cross-plotting the data shows the optimum horsepower combination for these two depths, Figs. 14 and 15.

Table I illustrates that there is no marked effect of liquid depth on horsepower required and these aeration basins can be of any desired depth depending upon overall economics.

Table II shows that any mixture of gases can be used effectively all the way from air to oxygen-enriched air to pure oxygen.

COAGULATION AND FLOCCULATION

Low concentrations of solids can be removed if they are coagulated and flocculated. This usually involves two steps. The first step is flash mixing and involves making a good contact of the chemical with the colloidal solids to start the coagulation process. There may also be a requirement for some of the chemical additives to react with each other in producing the proper chemicals to effect coagulation.

A well-mixed tank provides a uniform concentration and also provides a variety of residence times. A series of well-mixed tanks can provide a very sharp residence time distribution which can be effective in providing good additional contacting of the floc and chemical particles.

In the flocculation step, velocity gradients and shear rates are both desirable and undesirable. There is usually an optimum range of shear rates. It is well to pause here and discuss a few principles of shear rates with impeller type mixers.

An impeller produces a velocity profile similar to that shown in Fig. 16. The velocity gradient is the shear rate and varies all along the impeller discharge stream.

Shear stress, which actually does the work, is equal to viscosity times shear rate. If you multiply both sides of that equation by shear, you find that the instantaneous power per unit volume at a point is equal to viscosity times (shear rate)^2. By analogy, the average shear rate in a tank is equal to the average power per unit volume throughout the tank and has led to the concept of the symbol $G = (HP/V)^{1/2} (\mu)^{-1/2}$.

The shear work is the product of the length of time the shear rate acts on a particle and is given by the product Gt. However, each individual particle is in the tank for a different residence time and in a well-mixed flocculator, and also probably experiences a different shear history. The product Gt represents primarily an average concept of what is happening in the tank and is actually a composite of many different individual situations.

Impeller type flocculators are a very economical device compared to the slow speed, reel and rake type flocculators used in the past. Considerable work is going on in defining the shear rates at various points in the tank. Table III defines eight types of shear rates pertinent in mixing tanks, referring back to a published article defining these values[1].

Peripheral speed is an indication of shear rate only when comparing a given type of impeller versus known data on shear rates. If the pumping capacity at a given tip speed is changed, either by the number of blades. impeller position or blade design, then the shear rate is changed at that tip speed.

Tip speed can be used when comparing standardized impeller and tank geometries with themselves.

SOLID SUSPENSION

Many times it is desired to maintain solids in suspension in storage and holding tanks. In this case there are two different kinds of solid suspension processes; particulate solids must also be kept in suspension until they can reach the clarifier or filtering step. It is not usually necessary to maintain Complete Uniformity of all the solids in the tank. It is quite acceptable to have the larger particles at a higher concentration in the bottom of the tank just as long as they do not accumulate into progressive fillets.

Figs. 17 and 18 illustrate two different suspension conditions in a slurry suspension tank. The only requirement for steady state operation is that the solids at the discharge points must equal in composition the incoming feed stream.

Fig. 18 shows that if the draw-off point is moved upward in the tank, the entire composition in the tank is changed, assuming that the mixer is changed so that the new condition can be maintained in steady state.

Large holding and storage tanks which only require low horsepower level for blending must be analyzed carefully for the distribution of solids when a mixture of particle sizes flows into these tanks.

When the occasion arises, slurry holding tanks can be extremely large. Illustrative of what can be done, although not typical of current waste-treating practice, are large tanks of several million gallons, holding 45%-60% by weight coal slurries. The slurry is ground so the fine materials form a pseudoplastic viscosity to keep the heavier particles in suspension.

---

[1] "The Spectrum of Fluid Shear in a Mixing Vessel," J.Y.Oldshue, CHEMECA'70 Austral.Acad.Sci., Aust.Nat'l Comm. Inst. Chem.Eng.,Session 5, Melbourne-Sydney, Aug 19-26 (1970).

SOLID MATERIALS

Fluidized beds, where air or other gases are used to fluidize the particles, are a common way of treating powdered inorganic materials. Mixing can be used to reduce the amount of air required for fluidization or to improve the heat transfer for reaction control, Fig. 19. Mixers are not usually used for contacting air and organic wastes in combustion processes.

It is commonly required to repulp waste materials, and of course, the repulping of waste paper into liquid slurry for handling in paper processing is well known. Various types of side-entering and bottom-entering repulpers are used. This is primarily for putting solid materials back into chemical plant flow streams.

CHEMICAL REACTIONS

Limestone is often added to neutralize and cause precipitation of exit streams. In this case both solid-suspension and mass transfer rate between the limestone and the reagents must be considered. Pilot planting experiments can yield the required data on the effect of mixing and reaction rate.

The addition of CO_2 for various kinds of processes often causes calcium carbonate to precipitate. This can be a very troublesome phenomenon if it coats the blades of the impeller or tank structures. Proper agitation can allow the use of CO_2 in neutralization if the power level and mechanical shear stresses are properly controlled.

PILOT PLANTING

Pilot planting involves two essential considerations:

1. to determine the most prominent controlling parameter so the proper mixing scale-up techniques are used.

2. to be sure the parameter studies are related to the conditions to be expected in the full-scale plant.

Most mixing processes involve many steps. Table IV shows the categories under which mixing processes can be divided. Each of these areas has its own scale-up principles and design factors. By studying the effect of mixer power on a particular process, many of the controlling factors can be evaluated. For example, the effect of mixing power level on solid-liquid mass transfer is different from that on blending and from that on solid-suspension.

The other effect to be considered is the fact that fluid shear rates tend to increase on scale-up as shown in Fig. 20. Therefore, known geometric similarity must be used in the pilot plant if the effect of fluid shear is to be evaluated or controlled. Geometric similarity only provides similarity of geometry and does not give controlled ratios of any other parameters.

TABLE I

EFFECT OF DEPTH
CONSTANT TOTAL VOLUME

Z	TOTAL HP	$\%O_2$ ABSORBED
20	3200	29%
30	3200	39
40	3200	48

TABLE II

V = 12 million gallons,
Z = 20 ft
R = 100/mg/liter/hr

O_2 CONC.	IMPELLER HP	GAS HP	O_2 ABS.	O_2 IN EXIT
.21	1600	1600	.3	.14
.5	800	700	.3	.35
.5	1600	400	.5	.25
1.0	500	300	.3	.7
1.0	1600	100	.8	.2

TABLE III

There are four important shear zones in the tank. Shear rates may be defined based on average velocity at a point as well as the RMS velocity fluctuation.

AVERAGE POINT VELOCITY

MAX. IMPELLER ZONE SHEAR RATE
AVE. IMPELLER ZONE SHEAR RATE
MIN. TANK ZONE SHEAR RATE

RMS VELOCITY FLUCTUATIONS

$$\sqrt{\overline{(u')^2}}$$

TABLE IV

MIXING PROCESSES

PHYSICAL PROCESSING	APPLICATION CLASSES	CHEMICAL PROCESSING
Suspension	Liquid-Solid	Dissolving
Dispersions	Liquid-Gas	Absorption
Emulsions	Immisc.Liquids	Extractions
Blending	Misc. Liquids	Reactions
Pumping	Fluid Motion	Heat Transfer

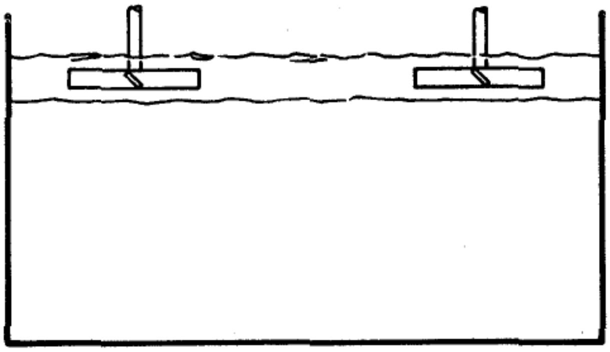
TABLE V

NOMENCLATURE

A,B,D	Locations of sampling points
C	Instantaneous average composition in either a flowing line or a blending tank
G	Average velocity gradient
Gt	Shear work, velocity gradient times residence time
HP	Horsepower
L	Length of tank
R	Uptake rate, mg/liter/hr
RMS	Root Mean Square velocity fluctuation
T	Tank diameter
t	Time
u'	Velocity fluctuation component
Z	Liquid level
$\frac{\Delta v}{\Delta y}$	Fluid velocity gradient

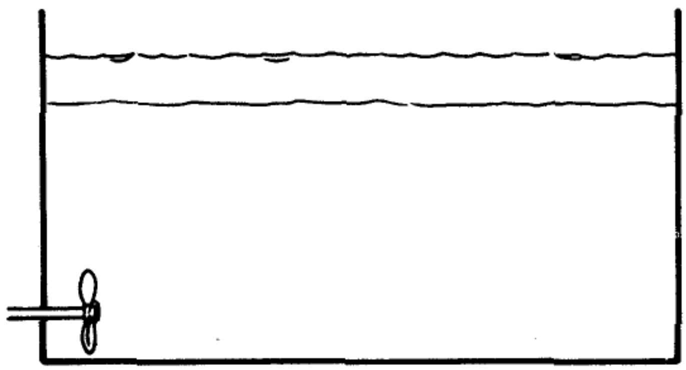
Nomenclature continued -

μ	Viscosity
Θ	Residence time in blend chest
λ	Period of concentration fluctuations



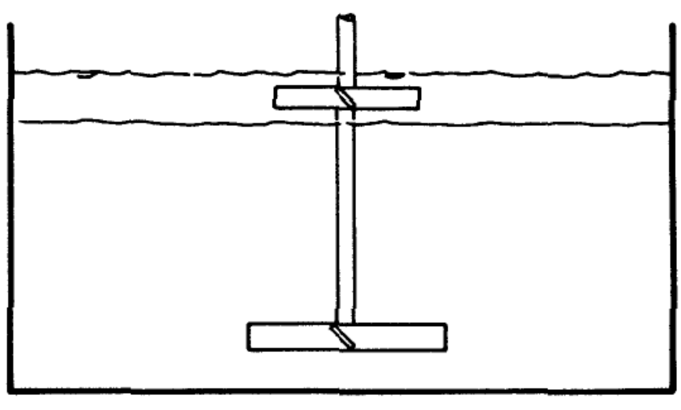
DIGESTERS

FIG. 1. TYPICAL ANAEROBIC DIGESTER WITH TOP-ENTERING MIXER IN SCUM LAYER



DIGESTERS

FIG. 2: TYPICAL ANAEROBIC DIGESTER, SIDE-ENTERING MIXER IN MAIN BLENDING ZONE.



DIGESTERS

FIG. 3: TYPICAL ANAEROBIC DIGESTER, TOP-ENTERING MIXER, IMPELLER IN SCUM LAYER AND IN MAIN BLENDING ZONE.

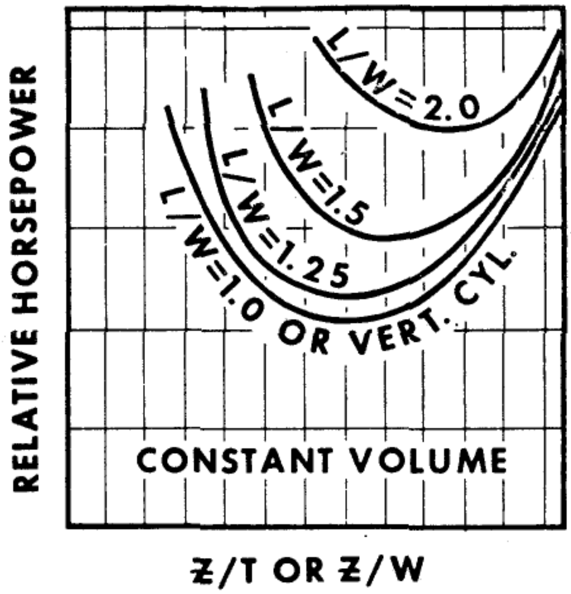
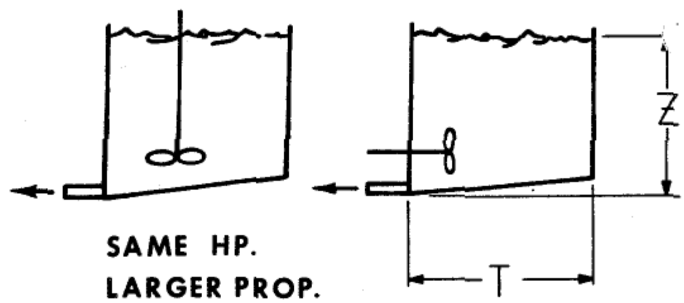


FIG. 4: HORSEPOWER REQUIREMENT FOR EQUAL PROCESS RESULT



SAME HP.
LARGER PROP.
SLOWER SPEED
LONGER SHAFT
LARGER DRIVE
STEADY BEARING
COST — 50% MORE

FIG. 5: COMPARISON OF TOP-ENTERING AND SIDE-ENTERING INITIAL COST AT Z/T GREATER THAN 0.4

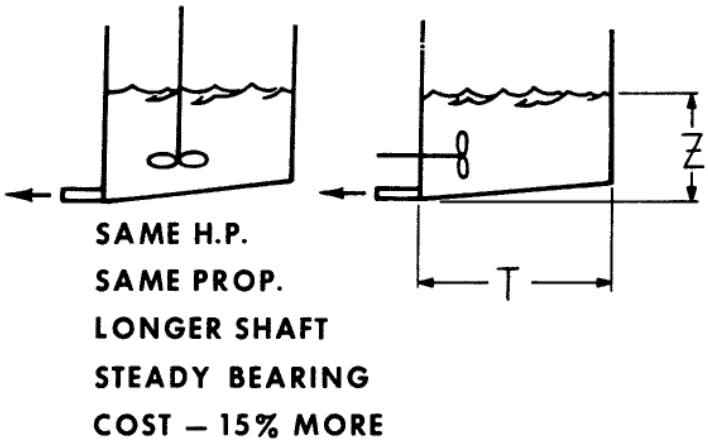


FIG. 6: COMPARISON OF TOP-ENTERING AND SIDE-ENTERING INITIAL COST AT Z/T BETWEEN 0.4 AND 0.25

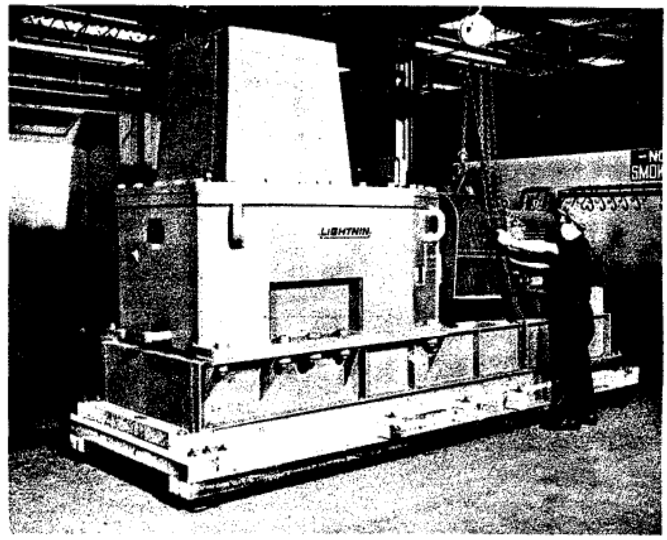


FIG. 8: ONE OF THE WORLD'S LARGEST MIXER DRIVES DESIGNED AND BUILT FOR A NEW WASTE TREATMENT SYSTEM.

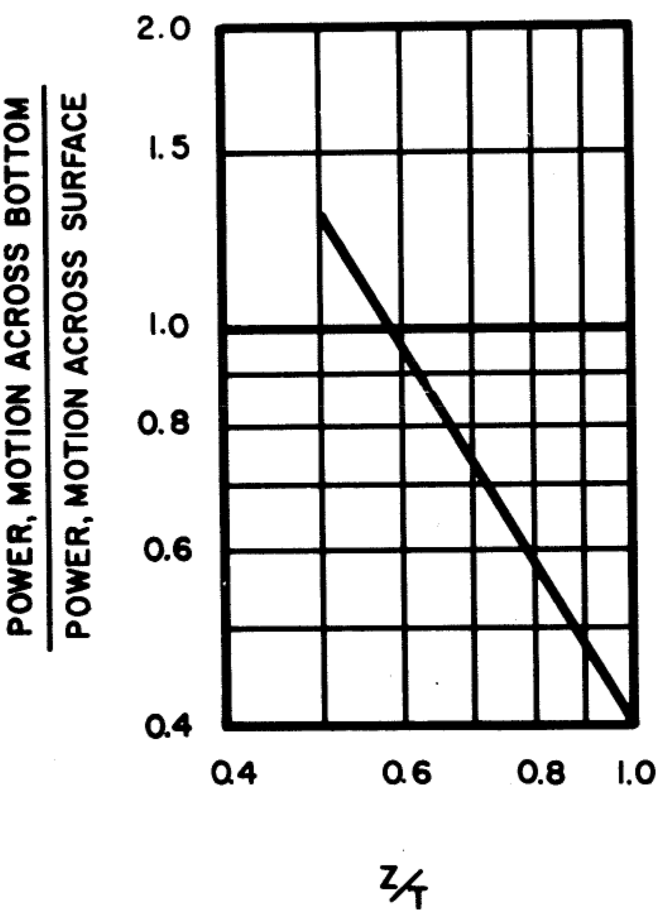


FIG. 7: THE RELATIONSHIP BETWEEN THE POWER REQUIRED TO PRODUCE MOTION ACROSS THE TANK BOTTOM RELATIVE TO THE POWER REQUIRED TO PRODUCE MOTION AT THE SURFACE OF A TANK WITH A SIDE-ENTERING MIXER, 4% PAPER PULP SLURRY.

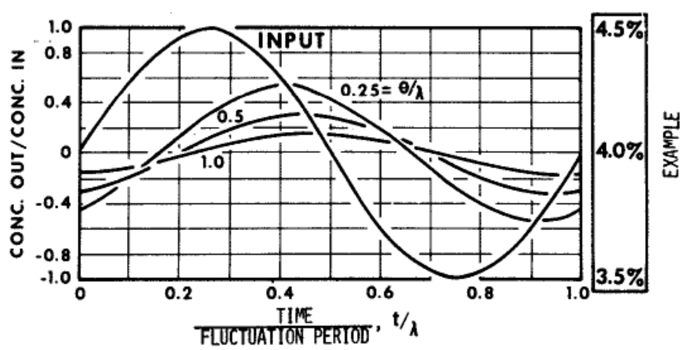


FIG. 9: ILLUSTRATION OF OUTPUT AMPLITUDE VERSUS INPUT AMPLITUDE FOR VARIOUS RATIOS OF RESIDENCE TIME, θ, TO THE PERIOD OF A SINGLE FREQUENCY SINUSOIDAL FLUCTUATION.

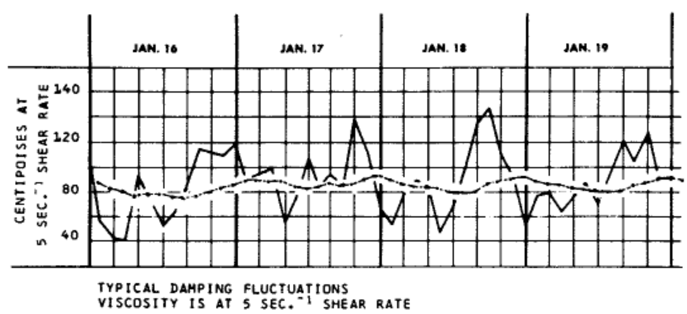


FIG. 10: TYPICAL REDUCTION IN VISCOSITY VARIATION BY USE OF BLENDING TANK

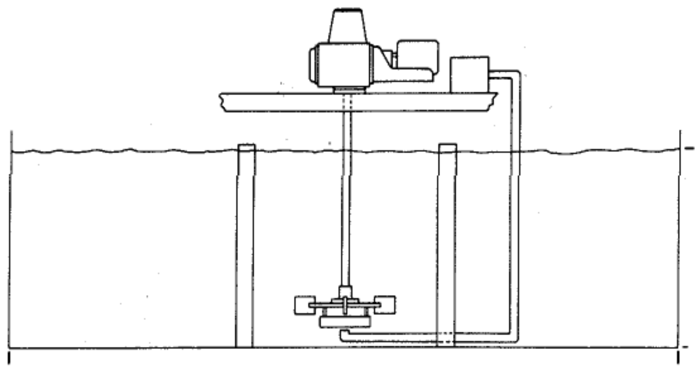


FIG. 11: SCHEMATIC OF PLANT SIZE INSTALLATION FOR SUBMERGED TURBINE REACTOR.

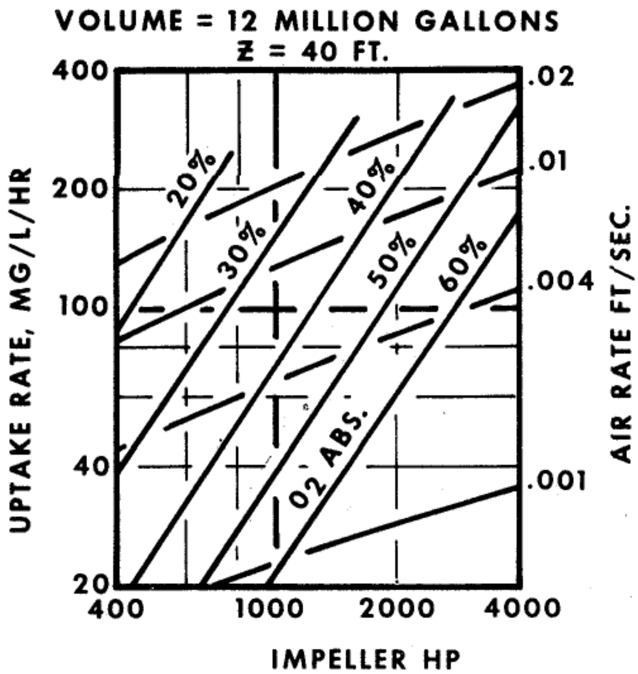


FIG. 13: IMPELLER HORSEPOWER AND AIR RATE FOR VARIOUS OXYGEN UPTAKE RATES WITH 40-FT. LIQUID DEPTH

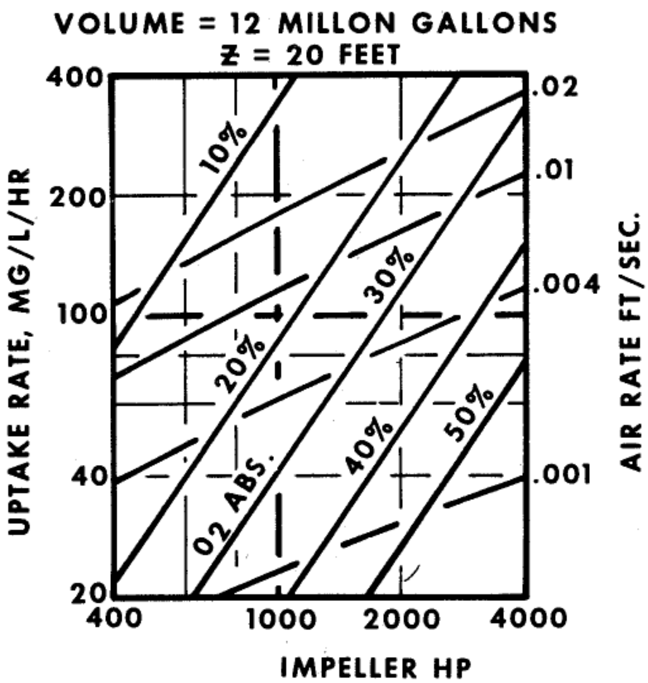


FIG. 12: IMPELLER HORSEPOWER AND AIR RATE FOR VARIOUS OXYGEN UPTAKE RATES WITH 20-FT. LIQUID DEPTH

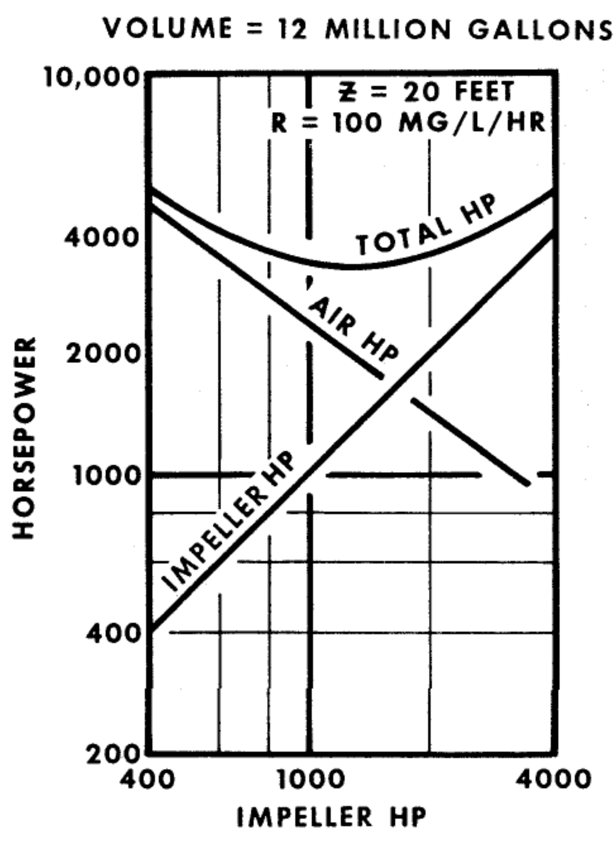


FIG.14: TOTAL HORSEPOWER FOR VARIOUS COMBINATIONS OF IMPELLER HORSEPOWER AND AIR BRAKE HORSEPOWER FOR 20 FT LIQUID DEPTH

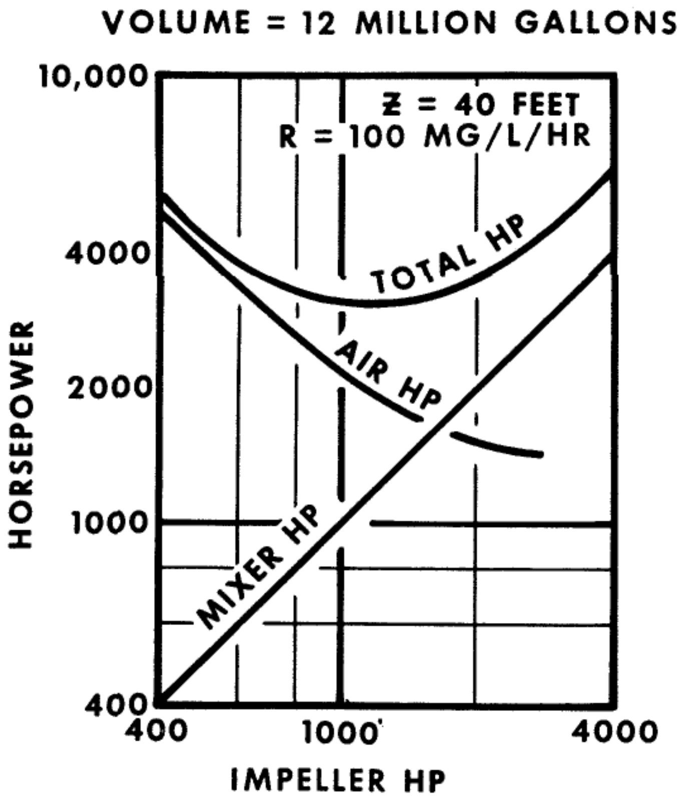


FIG. 15: TOTAL HORSEPOWER OF VARIOUS COMBINATIONS OF IMPELLER HORSEPOWER AND AIR BRAKE HORSEPOWER FOR 40 FT DEPTH.

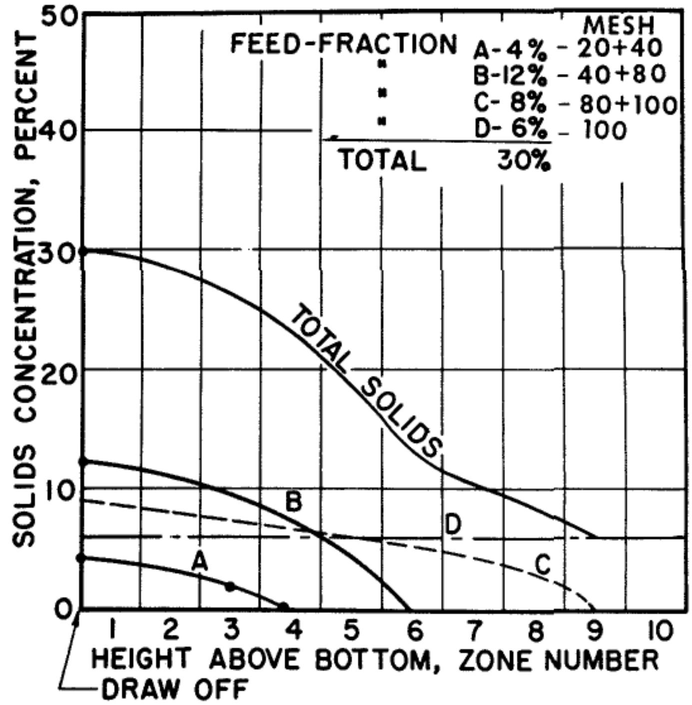


FIG. 17: TYPICAL DISTRIBUTION OF SOLIDS FOR COMPLETE OFF-BOTTOM SUSPENSION. DRAW-OFF AT BOTTOM

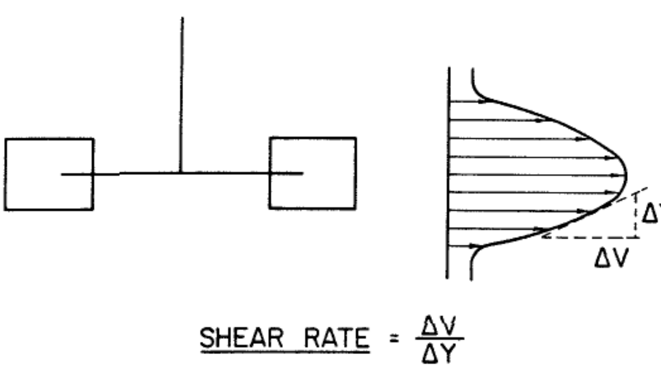


FIG.16: SCHEMATIC DRAWING OF VELOCITY PROFILE FROM TURBINE IMPELLER.

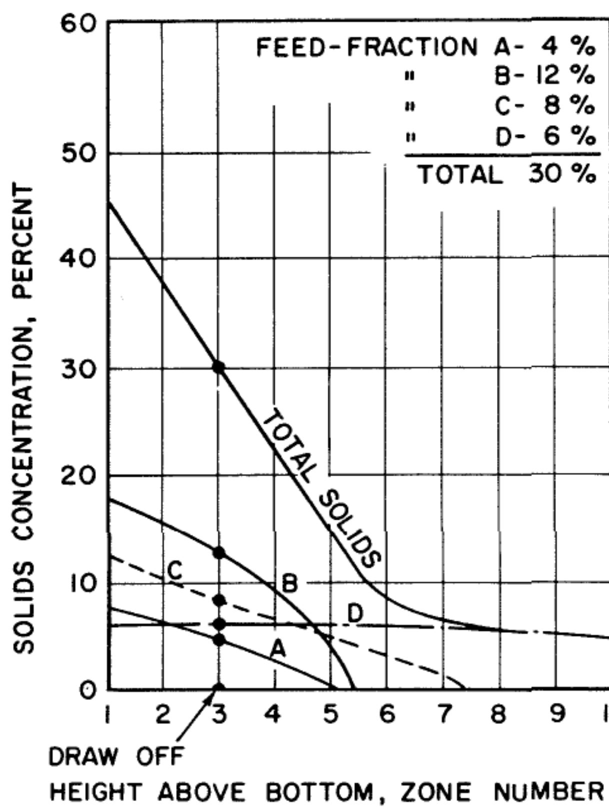


FIG. 18: TYPICAL DISTRIBUTION OF SOLIDS FOR COMPLETE OFF-BOTTOM SUSPENSION. DRAW-OFF ELEVATED FROM BOTTOM

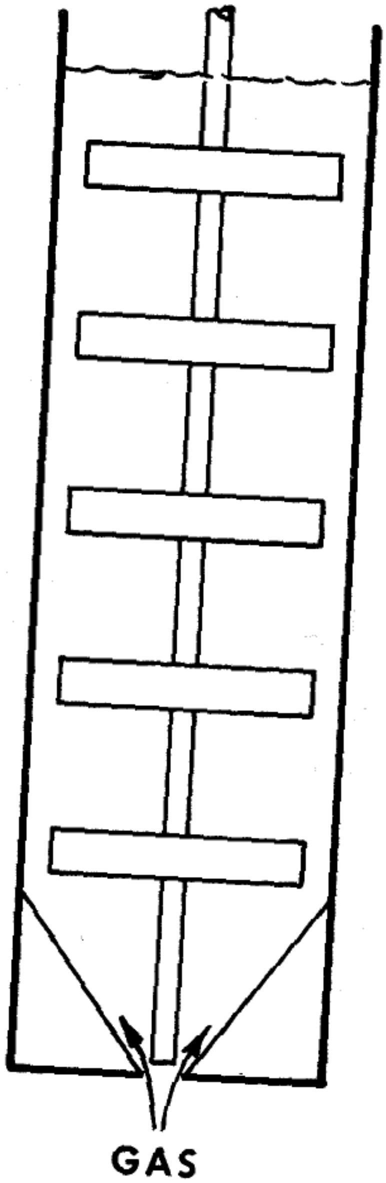


FIG. 19: MIXERS USED IN FLUIDIZED SOLIDS APPLICATION

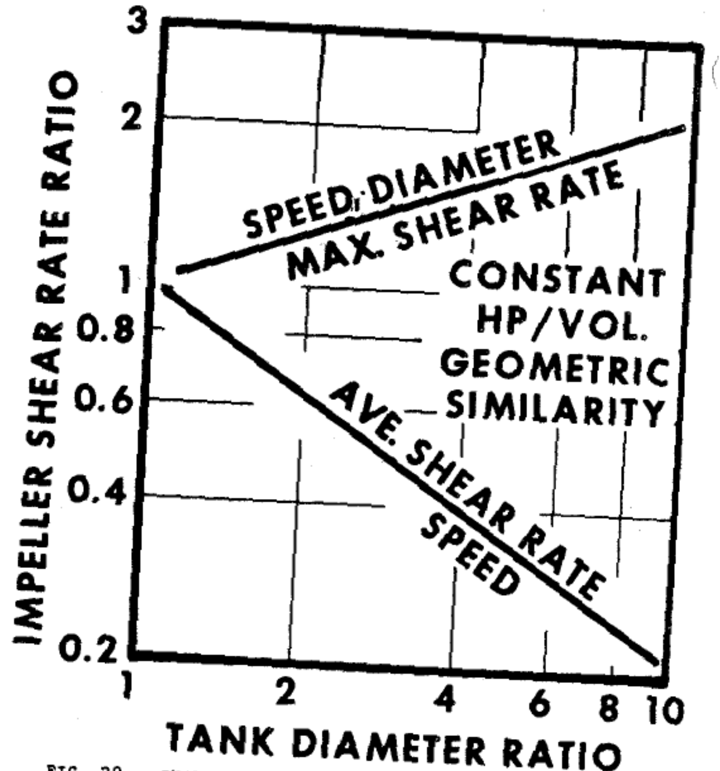


FIG. 20: CHANGE IN MAXIMUM IMPELLER-ZONE SHEAR RATE AND AVERAGE IMPELLER-ZONE SHEAR RATE WITH SCALE-UP