MIXING EQUIPMENT

Principle application of mixing equipment in the process industries are:

1. Blending of solid powders or pastes
2. Suspension of solids in liquids
3. Dispersion or emulsification of immisible liquids
4. Solutions of solids, liquids or gases.
5. Promotion of chemical reactions.

Standard equipment for liquid mixing is a vessel furnished with a rotating stirrer.

Assume physical mixing and it is desired to reproduce this intensity of mixing in the prototype.

SCALE-UP RULE (By Hixon and Rushton).

The conditions for equal degree of mixing can be obtained by keeping -

The power input same for both model and prototype and by basing the scale-up on RPM - (N). of the agitator.

These conclusions were reached by studying the heat and mass transfer coefficients and obtaining equal heat and mass coefficients

in model and prototype for the above two conditions.

The generalised dimensionless equations for fluid motion and heat & mass transfer are applicable to mixer also.

Consider the equation for fluid motion:-

$$\frac{\Delta P}{\rho U^2} = (N_{Re}, N_{Fr}) \propto \frac{V^2}{lg}.$$

write pressure coefficient in terms of power consumption and putting $Nd = v$

$N \rightarrow$ Angular velocity of agitator

$d \rightarrow$ diameter of agitator

$v \rightarrow$ fluid velocity.

$$N_{Re} = \frac{dv\rho}{\mu} = \frac{Nd^2\rho}{\mu}.$$

$$\frac{P}{\rho N^3 d^5} = \left(\frac{Nd^2\rho}{\mu}, \frac{N^2 d}{g} \right).$$

For heterogeneous ~~equation~~ liquid phase, weber group should be included.

$$\frac{Pgc}{\rho N^3 d^5} = \phi \left(\frac{Nd^2\rho}{\mu}, \frac{N^2 d}{g}, \frac{N^2 d^3 \rho}{\sigma} \right)$$

Basic eqn from which basic rules can be derived

Q4(a) ~~Power Number~~ is dimensionless parameter used for estimating the power consumed by the mixing impeller. $N_P = \frac{P}{n^3 D_a^5 \rho}$.

A power number is a characteristic # of an impeller geometry that can be used to calculate the impeller power. can be calculated for any size impeller with the same geometry, no. of blades, blade width, etc.

Pressure coefficient → $\frac{\Delta P}{\rho V^2}$

Ratio of Pressure force to inertial force.

Convert Pressure coefficient into power number

$$\frac{\Delta P}{\rho V^2} = \frac{\text{Power} \times \text{Time}}{\text{Distance} \times \text{area}} \cdot \left(\frac{1}{\rho V^2}\right)$$

$$\Delta P = \frac{\text{Force}}{\text{Area}} =$$

$$\text{Power} = \frac{\text{Work}}{\text{Time}} = \frac{\text{Force} \times \text{Distance}}{\text{time}}$$

$$\frac{\text{Power} \times \text{Time}}{\text{Distance}} = \text{Force}.$$

$$\Delta P = \frac{\text{Power} \times \text{Time}}{\text{Distance} \cdot \text{area}}$$

$$= \frac{\text{Power}}{\rho v^3 \cdot (d^2)}$$

$$= \frac{P}{\rho (Nd)^3 (d^2)} = \frac{P}{\rho N^3 d^5}$$

classify the mixers into baffled and unbaffled mixer.

$$\frac{P}{\rho N^3 d^5} = C\left(\frac{nd^2\rho}{\mu}\right)^m \left(\frac{N^2 d}{g}\right)^n$$

The value of n varies according to the flow conditions.

The value of m varies between 0 to −1.0 as the flow varies between turbulent to streamline.

When $N_{Re} < 10$, the flow is classified as streamline.

$N_{Re} > 100$ to 1000, it becomes turbulent.

Baffled $n \to 0$.

unbaffled $n = \dfrac{a - \log \frac{Nd^2}{\mu}}{b}$.

where a and b depends on geometry.

The power number is influenced by both N_{Re} and N_F.

The aim is to determine total flow pattern and power consumption in an unbaffled mixer.

This can be achieved by employing a liquid of lower kinematic viscosity in the model than in the prototype.

$d = d_{Prototype}/d_{Model}$. $\nu = \nu_P/\nu_m$

Simultaneous equality of Reynolds and Froude number can be attained in geometrically similar unbaffled mixers if the agitator diameters and speeds are related to the ratio of kinematic viscosities as follows.

$$d = \nu^{2/3}$$

$$N = \nu^{-1/3}.$$

Under these conditions, the power numbers in model and prototype are equal.

$$\frac{P}{\rho N^3 d^5} = C$$

$$P = \rho N^3 d^5 = \rho (\nu^{-1/3})^3 (\nu^{2/3})^5$$

$$= \rho \frac{\nu^{10/3}}{\nu} = \rho \nu^{7/3}.$$

Baffled Mixer :-

In baffled mixers the effect of vortexing is completly eliminated. It may be possible to achieve this by offsetting the stirrer from the centre line, this is called an offset mixer.

In both the cases the influence of N_F is negligible and when we consider only the homogeneous fluid, the influence of N_{We} is also negligible and the generalized dimension - less equation reduces to :

$$\frac{P}{\rho N^3 d^5} = C\left(\frac{N d^2 \rho}{\mu}\right)^m$$

To bring the power equation into a form comparable with standard heat and mass transfer equations, Replace P/d^3 in the above eqn by the power input per unit volume $= \pi$

$$\left(\frac{P}{\rho N^3 d^5}\right)\left(\frac{N d^2 \rho}{\mu}\right) = C\left(\frac{N d^2 \rho}{\mu}\right)^{m+1}$$

$$\frac{\pi}{N^2 \mu} = C\left(\frac{N d^2 \rho}{\mu}\right)^{x_f}$$

$x_f = x_{n+1}$.

(Modified function)

Ratio of applied torque to to viscous forces.

Retaining the agitator diameter d as the characteristic length for the system, and assuming the Prandtl and Schmidt no. to be constant, the corresponding heat and mass-transfer equations for homologous mixing systems are

$$\frac{hd}{k} = C_a \left(\frac{N d^2 \rho}{\mu}\right)^{x_A}$$

$$\frac{Kd}{D} = C_s \left(\frac{N d^2 \rho}{\mu}\right)^{x_B}$$

where x_A and x_B are the Reynolds indices for heat and mass transfer respectively.

→ Above equations are applicable where transfer surface is fixed.

→ Mass transfer eqns is applicable only to the solution or crystallization of solid particles

→ Immiscible liquids and liquid-gas mixtures are subject to the influence of the Weber gr.

Liquid – liquid System:

→ For dispersion of immiscible liquids, Nwe cannot be neglected.

$$Nwe = \frac{\rho v^2 L}{\sigma} = \frac{\text{Inertial Force}}{\text{Surface Tension}}.$$

If S is the specific interface of the two-phase system, inversely proportional to the mean droplet or bubble diameter; then for geometrically similar mixers and equal volumetric proportions of two phases the empirical equations may be

$$Sd \propto \left(\frac{N^2 d^3 \rho}{\sigma}\right)^{0.6}$$

For Homologous System, the above relations reduces to

$$S \propto (N^3 d^2)^{0.4}$$

For gas-liquid system, it is found that

$$Sd \propto \left(\frac{N^2 d^3 \rho}{\sigma}\right) / \left(\frac{N d^2 \rho}{\mu}\right)^{0.5}$$

For Homogeneous system

$$S \propto (N^3 d^2)^{0.5}$$

Dispersion of solid in liquid:

For preparing dispersions of solids in liquids, the most suitable criteria is power input per unit volume.

In homologous mixing systems at equal values of π the degree of dispersion is same.

Since it is only fully baffled or correctly offset mixers that can be scaled up on Reynolds index alone, the value of x may be taken as 1.

$$\frac{P_{ge}}{\rho N^3 d^5} = C_1 \left(\frac{N d^2 \rho}{\mu}\right)$$

$$\frac{P}{d^3} = \pi$$

$$\frac{\pi}{\rho N^2 d^2} = C_1 \left(\frac{N d^2 \rho}{\mu}\right)^x$$

$$\boxed{\pi = \frac{C_1 N^3 d^2 \rho}{g}}$$

From this

$$N = d^{-2/3} \qquad P = d^3.$$

Liquid-Liquid & Gas-liquid

Froude Similarity requires, the Nwe should be equal in model and prototype.

$N^2 d^3$ = Constant.

Sd = Constant

$N = d^{-2/3}$ $\quad S = d^{-1}$

PROBLEM :-

A heavy-tar oil emulsion is to be prepared.

Steam jacketed pan — with stirrer.

Procedure → To charge the pan with cold water at 60°C.

Stirrer is used to accelerate heat transfer.

emulsifying agent is added from separate tank.

Diameter 10" - jacketed pan

(RPM)N = 1500 rpm.

heating up time for batch — 2.4 min.

P = 0.004 hp.

Full-sized unit — geometrically similar

dia. = 5ft.

Q: What should be the stirrer speed to be given a degree of dispersion equal to that obtained in the small unit, what will be the power consumption & how long will a charge of

heats to heat up?

The ratio of stirrer speed should be

$$N = d^{2/3}.$$

linear scale ratio $L = d = \frac{60\,in}{10\,in} = 6.0$

Stirrer speed should be $1/6^{2/3} = 1/3.3$

The correct speed for full sized stirrer is therefore

$$\frac{1,500}{3.3} = 455 \text{ rpm}.$$

Power consumption of the full sized unit will be

$$\frac{P}{d^3} \qquad 0.004 \times 6^3 = 0.86 \text{ hp}.$$

The jacked heating surfaces are geometrical similar and the model has no dummy surface. therefore, it has 6 times as much heating surface per unit volume as the large pan.

The ratio of peripheral stirrer speed

$$V = Nd = \frac{6}{0.33} = 1.82$$

Scale ratio $N = \frac{1}{0.33}$.

general Scale-up equation for heat transfer by forced convection -

$$h = \frac{v^x}{L^{1-x}}$$

Consider, $x = 0.6$

Reynold's Index

$$h = \frac{(1.82)^{0.6}}{(6)^{0.4}} = \frac{1}{1.43}$$

Hence, heating up time in full sized foam, heating time in model is 2.4 min.

$$h = \frac{h}{\text{per unit area per unit time}}$$

$2.4 \times 6 \times 1.43 = 20.6 \text{ min}$.

$\frac{t_p}{t_m}$

$$h = \frac{\text{heat transfer}}{\text{Area} \times \text{time}} \times \frac{(\text{Area} \times \text{time})_1}{\text{heat transfer}}$$

$$\frac{h_p}{h_m} = h_m \times (\text{Area})_m$$

$$\text{time} = 2.4 \times (6) \times 1.43 = 20.6 \text{ min}.$$

$$\frac{h_p}{h_m} = \frac{h_p (\text{area} \times \text{time})_m}{h_m \times (\text{area} \times \text{time})_p}$$

$$(\text{time})_p = \frac{h_p (\text{area})_m}{} \cdot (\text{time})_m .$$

Q4(b) Towers or columns are among the most commonly used devices for contacting two immiscible fluids.

Application :- In distillation, evaporating cool and liquid extraction -

The principle types of tower are wetted w spray, packed, baffle plate, Sieve plate perporated plate and bubble plate.

The scaling up of packed towers is a real problem beog the perpormance of a given tower can vary greatly with the conditions of operations.

In a countercurrent packed tower, the heavier fluid is introduced at the top end descends by gravity against a rising stream of the lighter fluid.

Either phase can be dispersed but in gas-liquid tower it is always a liquid that is dispersed because of the excessive power consumption when gas is bubbled through a tall column of liquid.

In liquid-liquid towers, the lighter fluid is often dispersed and caused to rise in droplets through the packing.

The relation of mass-transfer rates to the geometry and dimensions of the packing depends very much on whether or not the packing is preferentially wetted by the disperse phase.

→ To design a packed absorption tower one must know HETP, HTU, $K_G a$, $K_L a$, to make use of the theoretical equation. [HTU → height of transfer unit]

[HETP → Height Equivalent to Theoretical plate]

→ The two important conditions to be fulfilled in scale up of packed tower are:
the system should be geometrically similar and the flow pattern must be same.

→ To keep the flow pattern same we must have equal, N_{Re} and N_{Fr}.

→ To satisfy both, the following condition must be satisfied. $\frac{\mu'}{\mu} = \left(\frac{\rho'}{\rho}\right)^{3/2}$ END

HETP: Defined as the height of a section of packing such that the vapour leaving the top of the section has the same composition as the vapour in equilibrium with the liquid leaving the bottom of the section.

HTU: Defined as the height of a section of packing producing a change in vap. composition equal to mean [illegible] equilibrium vap. composition over the section

In an absorption tower we have
(1) liquid Flow (2) Gas Flow (3) Chemical R
if absorption with chemical reaction is
to be considered.

Liquid Flow:- The criteria for liquid Flow
in a packed tower is given by -
(Russell's Equation)

$$\frac{W}{d} = \frac{\rho_l^2 g \delta^3}{3\mu_l} \quad \text{— (1)}$$

W = liquid rate
d = length of wetted perimeter
ρ_l = density of liquid
μ_l = viscosity of liquid.

For completely wetted tower packing

$$Ld = \frac{1}{c} \cdot \frac{\rho_l^2 g \delta^3}{\mu_l} \quad \text{— (2)}$$

L = Superficial mass velocity of the liquid
(only fluid present in a given cross-sec
area).
δ = thickness of streamline film

which gives the dimensionless equation

$$\frac{\delta^3}{d^3} = c \frac{L\mu_1}{\rho_1^2 d^2 g} \quad \text{— (3)}$$

8/b. For kinematic similarity, it is constant

$$\frac{L\mu_1}{\rho_l^2 d^2 g} = \text{const.} \quad \text{— (4)}$$

∴ This is similarity criteria for liquid phase.

For the gas phase, the similarity criteria is a constant Reynold's number.

$$\frac{dv\rho}{\mu} = \frac{Gd}{\mu_g} = \text{const.}$$

$G_1 \rightarrow$ Superficial mass velocity of gas, lb/ft^3-hr.

$\mu_g \rightarrow$ Viscosity of gas.

For chemical similarity, there is a further requirement of constant liquid/gas ratio.

$$\frac{L}{G_1} = \text{const.}$$

In homologous system, these three requirements reduce to

$$\frac{L}{d^2} = \text{const.}$$

$$Gd = \text{const.}$$

$$\frac{L}{G_1} = \text{const.}$$

which are incompatible when d is varied.

PTO.

LIQUID DISTRIBUTION:

→ In wetted packing tower there is a tendency for the descending liquid to drift toward the wall.

→ Kirschbaum and Weimann investigated the flow pattern in ring-packed towers and found that this tendency increases with increasing tower diameter for a given ring diameter. This suggests that for similar liquid-flow pattern there should be some relation between packing diameter and tower dia.

→ Weimann suggested that the ~~pa~~ tower dia. should not be less than 25 times the ring dia., to have proper liquid distribution.

→ Chilton, recommend that the tower dia. should not less than 15 times the packing dia. for rings or 8 times the other dia. for other shapes.

FLOODING VELOCITY -

→ For every liquid rate, there is a gas rate above which the pressure drop rises steeply and a layer of liquid appears on top of the packing.

→ The greater the liquid rate, the lower is the gas rate at which this occurs.

→ Flooding represents the limiting gas velocity at which it is possible to maintain counter-current flow. above the flooding velocity, most of the liquid is blown back out of the tower top

→ Sherwood, Shipley and Holloway obtained a correlation of flooding rate in the gas-liquid towers which for homologous systems at constant gas/liquid ratio reduces to -

$$\frac{v_g^2}{d} = \text{const.}$$

$$v_g = d^{1/2}.$$

PRESSURE DROP

Scale up Rule:

$\Delta P = G^{1.85}/d^{1.15}$ (turbulent range)

$= C/d^2$ (Streamline)

The above rules are valid for ling packings.

HEIGHT OF THE PACKING —

Scale up Rules are different for different mass transfer operation.

$(HTU) \propto d$ — Distillation.

$(HTU)_L \propto d^{0.5}$ — Absorption liquid film.

$(HTU)_g \propto d^{1.5}$ — Absorption gas film

$(HTU) \propto d^{0.75}$ — Liquid extraction

UNIT - IV

SLURRY PROCESSES

Slurry → liquid / Solid System

The various type of slurry system that we normally encounter are:

Under physical processes:- crystallization, Solution, melting, Solidification, adsorption, extraction and flocculation.

Under chemical processes:- polymerization with flocculation.

In both these type, the primary requirement is that the solid should be in good suspension. That is the degree of stirring should be good. Hence the scale up of a slurry reactor reduces to the scaling up of 'isotropic mixing' conditions. When we consider physical slurry processes mass transfer rates and heat transfer rates and heat transfer rates further influence the efficiency of mixing. In such a system we must consider the scale up rules for heat and mass transfer rates.

Consides an example in which the reactant A gives the product B and a side reaction product C

$$A \xrightarrow{k_1} B ; A \xrightarrow{k_2} C$$

The second reaction is found to be of higher order than the first.

The rates are $k_1 f_1(A)$ and $k_2 f_2(A)$ and $f_2(A)$ has higher order in A than $f_1(A)$.

The products formed are:

for B $$M_1 = \int_\theta \int_V k_1 f_1(A)\, dt\, dv$$

for C $$M_2 = \int_\theta \int_V k_2 f_2(A)\, dt\, dv$$

The integrations were carried out over reaction time θ and reactor volume V.

Condition of Immediate Mixing.

To keep the formation of side reactions under check, the conditions of immediate mixing and the isotropic mixing should be considered for scale up.

To keep M_2/M_1 to lie below a certain value, we must keep the conc.^{n} of A in the reactor at a lower value. In continous operations, the incoming stream which has a high concentration of A, must after stering a reactor, be diluted with the contents.

For any mixing situation following scale up rules are valid.

(a) Nre $\quad D\rho f/\mu$ = constant

(b) Nf $\quad \mu^2/g$ = constant

(c) F/V (Power/unit volume) = constant

(d) The peripheral stirrer speed ND must be kept constant.

Mainly two types of mixing is considered

(1) Isotropic mixing → under which all particles are uniformly suspended.

(2) Non Isotropic turbulence – In this case pocket are formed or non uniform mixing.

ABSORPTION

6/10/ In absorption we must know whether the system is liquid film controlled or gas film controlled. Without actual data it is difficult to asses this.

However, the following empirical rules guide one to select the regime.

$$\frac{f}{m} \cdot \frac{k_{ia}}{k_{ga}} \cdot \frac{C_{av}}{P} \qquad \text{—— (1)}$$

When this ratio is greater than 10 gas film is said to be controlling.

Without values of k_{ia} and k_{ga} one can make use of following relation.

$$f_s / HP = C$$

Where, H = Henry's law constant

f_s = density of ~~actual~~ soluble gas at actual temp. and pressure of gaseous mixtures

P = Total pressure.

When C is less than 0.0005 the system is said to be gas film controlled and when it is more than 0.2 liquid film is said to be controlling.

Liquid Film

Sherwood and Holloway : $HTU \propto d^{0.32} L^{0.2}$

Othmer : $HTU \propto L^{0.2}$ and independent of d.

Sherwood : $HTU \propto L^{n}$

Similarly many other workers have given different relationships.

Gas Film controlled :

Sherwood and Holloway $HTU \propto G^{0.5} / L^{0.4}$

and independent of d.

Sherwood and Othmer $K_G a \propto G^{0.8} / d$.

Sherwood, Pigford $HTU \propto G^{0.31} / L^{0.33}$.

Evaporation

$$HTU_g \propto G^{0.4} d^{1.41}$$

Liquid Extraction :

$$(HTU)_C \propto (L_c/L_d)^{0.75} \propto (L_c/L_d)^{n} \propto (L_c/L_d)$$

(HTU) increases with 'D' and d for dispersed to continuous phase and (HTU) increases with d for continuous to dispersed phase.

Gas film control

Similarly for gas film controlled regime we obtain —

$(HTU)_g \propto G/k_g a \propto G^{0.4} d^{1.4}$ Turbulent.

G^d Isothermal Streamline

$G^{0.5} d$ Transition.

Q:— Discuss the difference between distillation and absorption.

Q:— Scale-up conditions for mass transfer equipment ?

Q6(b)

(1) Packing diameter in model should not be less than $\frac{1}{2}$" and preferably not less than $\frac{3}{4}$".

(2) Tower diameter should be at least 8 times greater than packing diameter.

(3) Wetting rate in the small scale tower should be the minimum as given in eqn.

$L/a = 0.85$ lb/cu.ft.

larger tower should operate at 2½ times the minimum wetting rate.

$$V_g \propto \sqrt{d}$$

Both model and prototype should operate at some flooding velocities.

(4) L/G_1 ratio should be same and V_d/V should be same for liquid extraction tower.

(5) Scale ratio of packing diameter sho not be more than 3.

(6) $L \propto G \propto d'^{1/2} \propto (4\epsilon/a)$

(7) Scale up of HTU are as follows.

$(HTU) = d^n$

$n = 0.5$ liquid film absorption

$= 1.5$ gas film absorption

$= 1.0$ Distillation

$= 0.75$ liquid Extraction.