GATEFLIX

# MECHANICAL OPERATIONS

For
CHEMICAL ENGINEERING

# MECHANICAL OPERATIONS

## SYLLABUS

Particle size and shape, particle size distribution, size reduction and classification of solid particles; free and hindered settling; centrifuge and cyclones; thickening and classification, filtration, agitation and mixing; conveying of solids.

# CONTENTS

# 1 SCREENING OF PARTICLES

## 1.1 PARTICLE CHARACTERISTICS

The particle characteristics are defined by their size, shape and density.

The shape of an individual particle is expressed in terms of the sphericity $\phi_s$. The equivalent diameter of a non-spherical particle is defined as the diameter of a sphere having the same volume as the particle. The sphericity $\phi_s$ is the ratio of the surface area of the sphere, whose diameter is equal to the equivalent diameter of the particle, to the actual surface area of particle. The sphericity $\phi_s$ is given by the relation

$$\phi_s = \frac{6/D_p}{S_p/V_p} \quad \text{.....(1)}$$

Where $D_p$ = equivalent diameter of the particle

$V_p$ = Volume of a particle

$S_p$ = Surface area of a particle

Particle sizes are usually expressed in different units based on the size range involved. Coarse particles are measured in millimeters, fine particles in terms of screen size, very fine particles in micrometers or nanometers. Ultrafine particles are also described in terms of their surface area per unit mass, usually in $m^2$ / g.

If the total mass of uniform particles of diameter $D_p$, sphericity $\phi_s$ and density $\rho_p$ in a sample is m, the total surface area of particles is given by

$$A = \frac{6m}{\phi_s \rho_p D_p} \quad \text{.....(2)}$$

## 1.2 MEAN OR AVERAGE PARTICLE SIZES IN A MIXTURE OF PARTICLES

There are many definitions for the average particle size for a mixture of particles. These include:

- **VOLUME – SURFACE MEAN DIAMETER**

The volume – surface mean diameter defined as

$$\bar{D}_s = \frac{\sum_{i=1}^{n} N_i \bar{D}_{pi}^3}{\sum_{i=1}^{n} N_i \bar{D}_{pi}^2} \quad \text{.....(3)}$$

If the number of particles in each fraction $N_i$ is known, then it is given by

$$\bar{D}_s = \frac{1}{\sum_{i=1}^{n} \left( x_i / \bar{D}_{pi} \right)} \quad \text{.....(4)}$$

- **ARITHMETIC MEAN DIAMETER**

It can be defined as

$$\bar{D}_N = \frac{\sum_{i=1}^{n}\left(N_i \bar{D}_{pi}\right)}{\sum_{i=1}^{n} N_i} = \frac{\sum_{i=1}^{n}\left(N_i \bar{D}_{pi}\right)}{N_T} \quad \text{.....(5)}$$

Where $N_T$ = total number of particles in entire sample

- **MASS MEAN DIAMETER**

It can be defined as

$$\bar{D}_w = \sum_{i=1}^{n}\left(x_i \bar{D}_{pi}\right) \quad \text{.....(6)}$$

- **VOLUME MEAN DIAMETER**

It can be defined as

$$\bar{D}_V = \left[\frac{1}{\sum_{i=1}^{n}\left(x_i / \bar{D}_{pi}^3\right)}\right]^{1/3} \quad \text{.....(7)}$$

**1.3 SCREENING**

It is a method to separate the fine particles on the basis of their sizes. To perform screen analysis following steps are used:

- A set of standard screens are arranged serially in a stack with the smallest mesh at bottom and largest at top.
- The sample is placed on the top screen and then the stack is shaken mechanically for a definite time, usually 20 minutes.
- The particles retained on each screen are removed and weighted and the masses of the individual screen increments are converted to mass fractions or mass % of the total sample.
- Any particles that pass the fines screen are caught in a pan at the bottom of the stack.
- The results of a screen analysis are tabulated to show the mass fraction of each increment as a fraction of the mesh size range of the increment.

Here, the notation 14/20 mean "through 14 mesh and on 20 mesh."

| **Mesh** | **Screen Opening $D_{pi}$, mm** | **Mass Fraction Retained, $x_i$** | **Average Particle Diameter in Increment, $\bar{D}_{pi}$, mm** | **Cumulative Fraction Smaller Than $D_{pi}$** |
|---|---|---|---|---|
| 4 | 4.699 | 0.0000 | - | 1.0000 |
| 14 | 1.168 | 0.25701 | 1.409 | 0.2722 |
| 48 | 0.295 | 0.0102 | 0.356 | 0.0282 |

| | | | | |
|---|---|---|---|---|
| Pan | - | 0.0075 | 0.037 | 0.0000 |

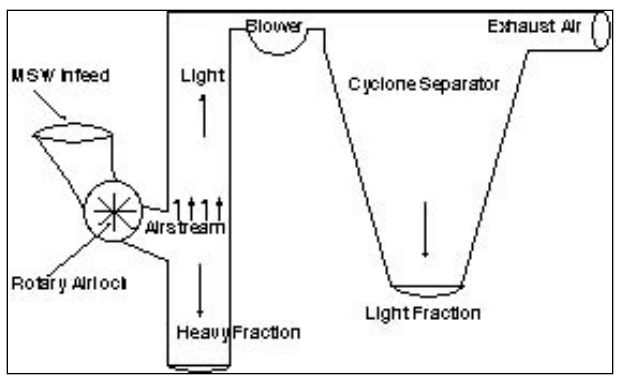
It is one method to classify the particles by their sizes.

## 1.4 ELUTRIATION

*Elutriation*, also known as *air classification*, is a process for separating lighter particles from heavier ones using a vertically-directed stream of gas or liquid (usually upwards). This method is predominately used for particles with size (>1μm). The smaller or lighter particles rise to the top (overflow) because their terminal velocities are lower than the velocity of the rising fluid. The terminal velocities of any particle in any media can be calculated using Stokes' Law if the particle Reynolds number is below 2.

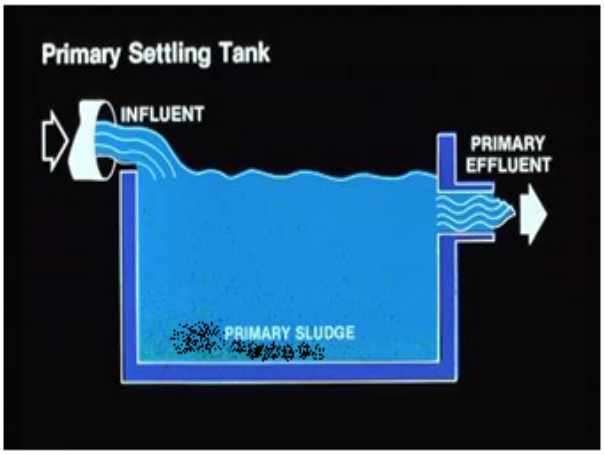
Elutriation is a materials separation method where smaller and larger materials are forced apart with the use of a column of liquid or gas. It can be used on the macro and micro level, from preparation of cell samples for analysis to separation of single stream recycling. Equipment for use in elutriation is available from a number of manufacturers and people can order custom equipment for unique applications. Such equipment tends to be more expensive than offerings on the conventional product lineup.

In elutriation, materials are fed into a rising column of separation medium. This can be something like a buffer solution or a jet of air. Small, light particles drift up in the column, while large, heavy particles sink down. This creates a series of layers of different materials of different sizes. In something like single stream recycling, large air jets are used for quick and basic separation, allowing plastic containers to go to one side of a processing facility, while glass lands on the other, for example, with metals being separated earlier in the process with the use of magnets.




## 1.5 SETTLING

*Settling* is the process by which particulates settle to the bottom of a liquid and form a sediment. Particles that experience a force, either due to gravity or due to centrifugal motion will tend to move in a uniform manner in the direction exerted by that force. For gravity settling, this means that the particles will tend to fall to the bottom of the vessel, forming a slurry at the vessel base.




## 1.6 CLASSIFICATION

*Classification* is defined as the separation of a mixture of solid particles into various fractions according to their size or density, which are allowed or caused to settle through a fluid either in motion or at rest. The fluid in question is generally water but it can also be air.

When the materials of the same density are separated according to their sizes, the operation is known as *sizing.*

When the materials of the same equivalent size are separated according to their densities, the operation is known as *sorting.*

### 1.6.1 PRINCIPLES OF CLASSIFICATION

When a solid particle falls through a vacuum under the influence of gravity alone, its velocity increases continuously due to acceleration. But when the same solid particle falls through a fluid such as air or water, its velocity increases at a lower rate due to friction caused by the movement of the particle in the fluid which cancels a part of the gravitational force. This frictional force increases with the increase in the velocity of the particle. And when the frictional force becomes equal to the gravitational force, the velocity of a particle reaches a constant value, known as the *terminal settling velocity.*

This velocity depends on the parameters such as the shape, size and density of the solid particle and the density and viscosity of the fluid.

### 1.6.2 LAWS OF CLASSIFICATION

Some of the general laws of classification are the following:

(i) The coarse particles have a relatively faster settling velocity than the fine particles of the same specific gravity and the same shape.

(ii) The heavy gravity particles have a relatively faster settling velocity than light gravity particles of the same size and the same shape.

(iii) The regular particles like the spherical ones have a relatively faster settling velocity than irregular particles of the same weight.

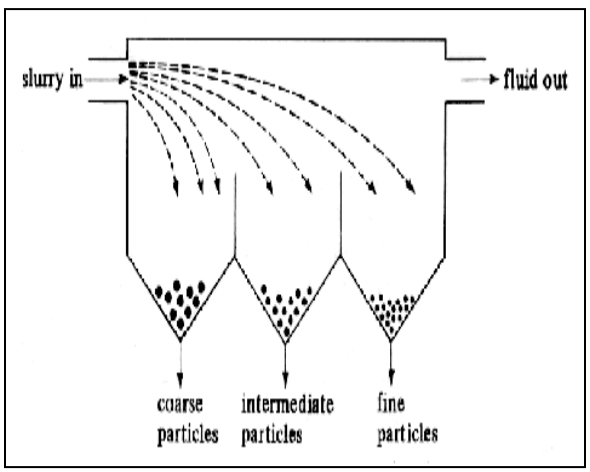
(iv) The settling velocity of solid particles decreases with the increase in fluid density and viscosity.

### 1.6.3 CLASSIFICATION EQUIPMENTS

There are basically two major type of classifying equipments:

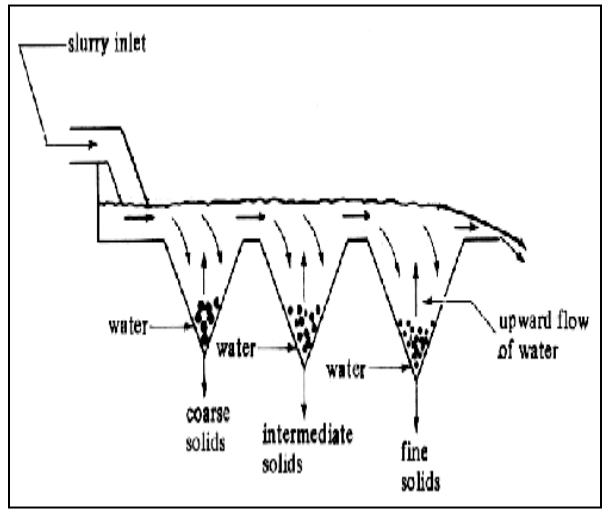
#### 1.6.3.1 SIMPLE CLASSIFIER

In the simple classifier, the design is similar to that of the straight gravity settling tank, except that the bottom half is divided into several equal partitions. What happens is that instead of just falling into a big mess on the bottom of the tank, the coarse particles get trapped in the first chamber, the intermediates get trapped in the middle partitions, and the fine particles, the dust, gets captured in the last section. Then, you can drain the sections from the bottom and have segregated sediment.




#### 1.6.3.2 SPITZKASTEN CHAMBER

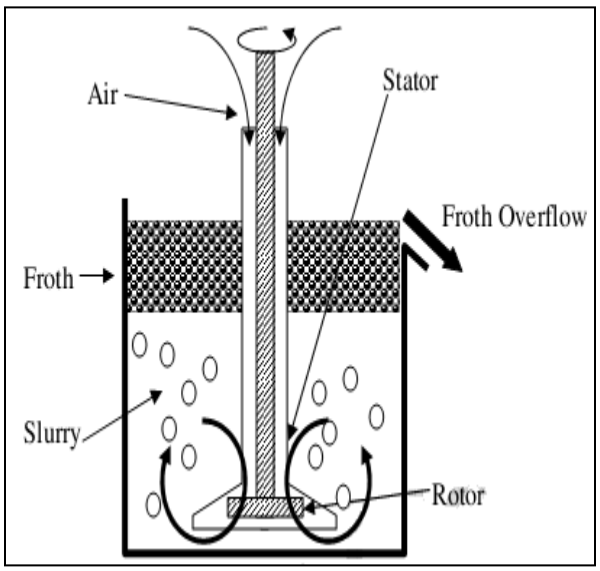
The Spitzkasten chamber runs like this. A series of conical vessels of increasing size is set up in the direction of flow. As the slurry enters the first vessel, the coarse particles get trapped, and the overflow continues on to the next, where more separation takes place. This particular settling chamber is unique because you can adjust the flow rates in between each vessel in order to provide the necessary degree of separation.




### 1.7 FROTH FLOATATION

Froth flotation is a highly versatile method for physically separating particles based on differences in the ability of air bubbles to selectively adhere to specific mineral surfaces in a mineral / water slurry. The particles with attached air bubbles are then carried to the surface and removed, while the particles that remain completely wetted stay in the liquid phase. Froth flotation can be adapted to a broad range of mineral separations, as it is possible to use chemical treatments to selectively alter mineral surfaces so that they have the necessary properties for the separation. It is currently in use for many diverse applications, with a few examples being: separating sulfide minerals from

silica gangue (and from other sulfide minerals); separating potassium chloride (sylvite) from sodium chloride (halite); separating coal from ash-forming minerals; removing silicate minerals from iron ores; separating phosphate minerals from silicates; and even non-mineral applications such as de-inking recycled newsprint. It is particularly useful for processing fine-grained ores that are not amenable to conventional gravity concentration




### 1.7.1 REAGENTS

The properties of raw mineral mixtures suspended in plain water are rarely suitable for froth flotation. Chemicals are needed both to control the relative hydrophobic ties of the particles, and to maintain the proper froth characteristics. There are therefore many different reagents involved in the froth flotation process, with the selection of reagents depending on the specific mineral mixtures being treated.

#### 1.7.1.1 COLLECTORS

Collectors are reagents that are used to selectively adsorb onto the surfaces of particles. They form a monolayer on the particle surface that essentially makes a thin film of non-polar hydrophobic hydrocarbons. The collectors greatly increase the contact angle so that bubbles will adhere to the surface. Selection of the correct collector is critical for an effective separation by froth flotation. Collectors can be generally classed depending on their ionic charge: they can be nonionic, anionic, or cationic. The nonionic collectors are simple hydrocarbon oils, while the anionic and cationic collectors consist of a polar part that selectively attaches to the mineral surfaces, and a non-polar part that projects out into the solution and makes the surface hydrophobic. Collectors can either chemically bond to the mineral surface (chemisorption), or be held on the surface by physical forces (physical adsorption).

#### 1.7.1.2 CHEMISORPTION

In chemisorption, ions or molecules from solution undergo a chemical reaction with the surface, becoming irreversibly bonded. This permanently changes the nature of the surface. Chemisorption of collectors is highly selective, as the chemical bonds are specific to particular atoms

#### 1.7.1.3 PHYSISORPTION

In physisorption, ions or molecules from solution become reversibly associated with the surface, attaching due to electrostatic attraction or van der Waals bonding. The physisorbed substances can be desorbed from the surface if conditions such as pH or composition of the solution changes. Physisorption is much less selective than chemisorption, as collectors will adsorb on any surface that has the correct electrical charge or degree of natural hydrophobicity.

#### 1.7.1.4 NONIONIC COLLECTORS

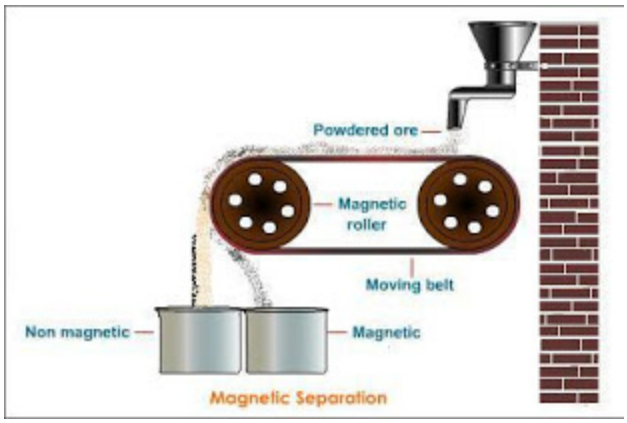
Hydrocarbon oils, and similar compounds, have an affinity for surfaces that are already partially hydrophobic. They selectively adsorb on these surfaces, and increase their hydrophobicity. The most commonly-floated naturally-hydrophobic material is coal. Addition of collectors such as #2 fuel oil and kerosene significantly enhances the hydrophobicity of the coal particles without affecting the surfaces of the associated ash-forming minerals. This improves the recovery of the coal, and increases the selectivity between coal particles and mineral matter. Fuel oil and kerosene have the following advantages over specialized collectors for froth flotation: 1) they have low enough viscosity to disperse in the slurry and spread over the coal particles easily, and 2) they are very low-cost compared to other compounds which can be used as coal collectors.

In addition to coal, it is also possible to float naturally-hydrophobic minerals such as molybdenite, elemental sulfur, and talc with nonionic collectors. Nonionic collectors can also be used as “extenders” for other collectors. If another, more-expensive collector makes a surface partially hydrophobic, adding non-polar oil will often increase the hydrophobicity further at low cost.

### 1.8 MAGNETIC SEPARATION

It is the method to separate the solid – solid mixture with the help of their magnetic properties. For this we use magnetic separator.

A magnetic separator is a device used to separate mixture of fine, dry materials based upon their magnetic properties. The principles governing this process are magnetism and the interaction between magnetic, gravitational and centripetal forces. Magnetic properties of a material are based upon atomic structure and magnetic field intensity. The principles involved in a separation apparatus include: feed rate, velocity of the particles and magnetic field strength. Magnetic separation has two general applications, purification of feeds, and the magnetic removal impurities or the collection of the magnetic components from the mixture.




## 1.9 ELECTROSTATIC SEPARATION

*Electrostatic separation* is a process that uses electrostatic charges to separate crushed particles of material. An industrial process used to separate large amounts of material particles, electrostatic separating is most often used in the process of sorting mineral ore. This process can help remove valuable material from ore, or it can help remove foreign material to purify a substance. In mining, the process of crushing mining ore into particles for the purpose of separating minerals is called *beneficiation.*

An electrostatic separator is a device for separating particles by mass in a low energy charged beam. It works on the principle of corona discharge, where two plates are placed close together and high voltage is applied. This high voltage is used to separate the ionized particles.

Usually these are used in power plants where the harmful gases coming out of the chimneys are first treated using electrostatic separator. here the two electrodes are oppositely charged, with a negative electrode the positive ions gets attracted and thus results in a reddish flame whereas the positive electrode is used to treat the negatively charged ions resulting in a bluish white flame that is visible at nights.

## 1.10 CENTRIFUGAL SEPARATION

A centrifugal separator is a machine that uses centrifugal, gravitational, and inertial forces to divide two or more substances. This device can be used to separate solutions, gas mixtures, or other matter that can be physically parted. Centrifugal separation occurs when a mixture in the machine's chamber is spun very quickly, and heavy materials typically settle differently than lighter ones. Centrifugal separators have a wide variety of applications, in many industries.

When a mixture enters the spinning chamber of a centrifugal separator, distinct substances within it are affected differently by the force created by the spinning. For example, gravitational force generally pulls heavier particles down more quickly than lighter ones, and the force of inertia affects the mixture as it spins. As the substances separate, they can be collected in various ways. Sometimes, they are collected mechanically, but, other times, they are physically separated. One method of this can be by screening

Heavier solid particles are often allowed to settle as they slide down the walls of the separator. They are then typically collected from the bottom. Generally, a gas can be purified by spinning any particulate matter and moisture out of it. The gas can

then be collected, as it rises to the top and through an exit opening in the centrifugal separator. Liquids of different weights and viscosities may be divided into differing chambers of a separator as it moves.

Some of the applications in which a centrifugal separator can be used include dividing cream from milk, sand from gravel, and oil from water. The food and beverage industries often use these machines in the making of syrups, sugars, and malt liquors. Manufacturers of paints and varnish also use these machines, as do pharmaceutical manufacturers, animal feed makers, and the ceramics and abrasives industries.

A particular kind of machine, known as a *centrifugal water separator*, is often used to remove water from compressed air. This is important because water in the air can cause rust to occur in the metal components of a compressor, and in any attached machines or tools that utilize the compressed air. Typically, any condensate is spun out of the air by the separator, and collected in bowls where it is then pumped out of the compressor.

As the technology of the centrifugal separator has progressed, new applications have been found for its use. There are devices known as *ultracentrifuges* that are being used to separate larger molecules into their components. This advancing technology is particularly useful in the pharmaceutical industry.

## 1.11 SIZE ENLARGEMENT

In a size – enlargement operation, small particles are brought together purposely to form larger ones, generally by some mechanical means. The size – enlargement operations are many, namely, agglomeration, granulation, compaction, encapsulation, pelletizing, sintering, etc. and the agglomeration method is discussed here in brief.

Size – enlargement operations are followed in the process industries with a wide variety of objectives, such as

- To improve storage and handling characteristics of materials
- To improve flowability and disability
- To minimize dusting or material losses
- To create a safe working environment
- To enhance appearance
- To control solubility and dispersibility

# GATE QUESTIONS

**Q.1** A sand mixture was screened through a standard 10 mesh 10 – screen. The mass fraction of the oversize material in feed, overflow and underflow were found to be 0.38, 0.79 and 0.22 respectively. The screen effectiveness based on the oversize is

(a) 0.50 (b) 0.58

(c) 0.68 (d) 0.62

**(GATE 2002)**

**Q.2** The cumulative mass fraction of particles smaller than size $d_j$ for a collection of $N_i$ particles of diameter $d_i$ and mass mi (I = 1, 2, 3…, ∞) is given by

(a) $\dfrac{\sum_{i=1}^{I} N_i d_i^3}{\sum_{i=1}^{\infty} N_i d_i^3}$ (b) $\dfrac{\sum_{i=1}^{I} N_i m_i d_i^3}{\sum_{i=1}^{\infty} N_i m_i d_i^3}$

(c) $\dfrac{\sum_{i=1}^{I} N_i m_i d_i^2}{\sum_{i=1}^{\infty} N_i m_i d_i^2}$ (d) $\dfrac{\sum_{i=1}^{I} N_i m_i d_i}{\sum_{i=1}^{\infty} N_i m_i d_i}$

**(GATE 2004)**

**Q.3** In Tyler series, the ratio of the aperture size of a screen to that of the next smaller screen is

(a) $1/\sqrt{2}$ (b) $\sqrt{2}$

(c) 1.5 (d) 2

**(GATE 2007)**

**Q.4** The particle size distributions of the feed and collected solids (sampled for same duration) for a gas cyclone are given below.

| Size range (μm) | Weight of feed in the size range (g) | Weight of collected solids in the size range (g) |
|---|---|---|
| 1 – 5 | 2 | 0.1 |
| 5 – 10 | 3 | 0.7 |
| 10 – 15 | 5 | 3.6 |
| 15 – 20 | 6 | 5.5 |
| 20 – 25 | 3 | 2.9 |
| 25 – 30 | 1 | 1 |

What is the collection efficiency (in PERCENTAGE) of the gas cyclone?

(a) 31 (b) 60

(c) 65 (d) 69

**(GATE 2011)**

**Q.5** In the Tyler standard screen scale series, when the mesh number increases from 3 mesh to 10 mesh, then

(a) the clear opening decreases,

(b) the clear opening increases,

(c) the clear opening is unchanged

(d) the wire diameter increases.

**(GATE 2013)**

**Q.6** Size analysis was carried out on a sample of grade. The data for mass fraction ($x_i$) and average particle diameter $(D_{pi})$ of the fraction is given in the table below:

| $x_i$ | $D_{Pi}$ (mm) |
|---|---|
| 0.2 | 5 |
| 0.4 | 10 |
| 0.4 | 20 |

The mass mean diameter of the sample, to the nearest integer, is ____mm.

**(GATE 2017)**

**Q.7** Match the equipment in Column A with the corresponding process in Column B

| Column A | Column B |
|---|---|
| (P) Centrifugal sifter | (I) Mixing |
| (Q) Bowl Mill | (II) Sedimentation |
| (R) Gravity thickener | (III) Screening |
| (S) Two-arm kneader | (IV) Grinding |

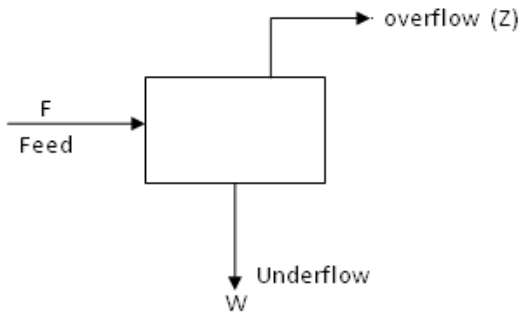
(a) P-I, Q-IV, R-II, S-III
(b) P-III, Q-IV, R-II, S-I
(c) P-IV, Q-I, R-II, S-III
(d) P-IV, Q-III, R-I, S-II

**(GATE 2018)**

| ANSWER KEY | | | | | | |
|---|---|---|---|---|---|---|
| **1** | **2** | **3** | **4** | **5** | **6** | **7** |
| (b) | (b) | (b) | (d) | (a) | 13 | (b) |

# EXPLANATIONS

**Q.1 (b)**




$$F = W + Z$$
$$0.38\ F = 0.22\ W + 0.79\ Z$$
$$0.38\ F = 0.22\ (F - Z) + 0.79Z$$
$$F(0.38 - 0.22) = Z\ (0.79 - 0.22)$$
$$F \times 0.16 = Z\ (0.57)$$
$$\frac{Z}{F} = \frac{16}{57}$$

Screen effectiveness

$$\frac{0.79 \times Z}{0.38 \times F} = \frac{0.79}{0.38} \times \frac{16}{57} = 0.583$$

**Q.2 (b)**

**Q.3 (b)**

Width of successive opening have a constant ratio of $\sqrt{2}$, areas of successive openings has a constant ratio of 2.

**Q.4 (d)**

$$x_i = \frac{\text{weight of the feed}}{\text{total weight of feed}}$$

$$\eta_i = \frac{\text{weight of collected solid}}{\text{weight of feed}}$$

$$\eta = \sum_{i=1}^{N} x_i \eta_i$$

$$x_1 = \frac{2}{20} = .01, x_2 = \frac{3}{20} = 0.15,\ x_3 = \frac{5}{20} = 0.25, x_4 = \frac{6}{20} = 0.3\, x_5 = \frac{3}{20} = 0.15, x_6 = \frac{1}{20} = 0.05$$

$$\eta_1 = \frac{0.1}{2} = 0.05, \eta_2 = \frac{0.7}{3} = 0.233\, \eta_3 = \frac{3.6}{5} = 07.2, \eta_4 = \frac{5.5}{6} = 0.917m, \eta_5 = \frac{2.9}{3} = 0.967, \eta_6 = \frac{1}{1} = 1$$

$$\eta = x_1 \eta_1 + x_2 \eta_2 + x_3 \eta_3 + x_4 \eta_4 + x_5 \eta_5 + x_6 \eta_6$$

$$\eta = 0.1 \times 0.05 + 0.15 \times 0.233 + 0.25 \times 0.72 + 0.3 \times 0.917 + 0.15 \times 0.967 + 0.05 \times +1 = 0.69$$

**Q.5 (a)**

In the Tyler standard screen scale series, when the mesh number increases from 3 mesh to 10 mesh, then the clear opening decreases

**Q.6 13**

$\sum x_i = 1$

Mass mean diameter of sample

$= \sum x_i D_{pi} / \sum xi = (0.2 \times 5) + (0.4 \times 10) + (0.4 \times 20) = 13 \text{ mm}$

**Q.7 (b)**

| Equipment | Process |
|---|---|
| Centrifugal shifter | Screening [Based on particle size separation] |
| Bowl mill | Grinding [ used for fine grinding to produce talcum powder] |
| Gravity Thickner | Sedimentation [used for waste treatment to settle and separate solid in fluid ] |
| Two arm kneader | Mixing [used mixing rubbery or polymer materials] |

# 2 SIZE REDUCTION

## 2.1 SIZE REDUCTION

Size Reduction may refer to the operation in which we reduce the size of coarse particles into fine or very fine particles. Reduction of particle size is an important operation in many chemical and other industries. The important reasons for size reduction are:

- Easy handling
- Increase in surface area per unit volume
- Separation of entrapped components

The operation is highly energy intensive; hence a variety of specialized equipment is available for specific applications. The equipment may utilize one or more of the following physical mechanisms for size reduction:

(i) Compression,
(ii) Impact,
(iii) Attrition,
(iv) Cutting.

Estimation of energy for the operation is important and is usually done by empirical equations. Enormous quantities of energy are consumed in size reduction operations. Size reduction is the most inefficient unit operations in terms of energy, as 99% of the energy supplied goes to operating the equipment and producing undesirable heat and noise, while less than 1% goes in creating new interfacial area. Reduction to very fine sizes is much more costly in terms of energy as compared to relatively coarse products.

## 2.2 EQUIPMENTS FOR SIZE REDUCTION

Crushers do the heavy work of breaking large pieces of solid material into small lumps. A primary crusher operates on run of mine material, accepting anything that comes from the mine face and breaking it into 6 – 10 in. lumps. A secondary crusher reduces these lumps to particles perhaps 6 mm in size. Grinders reduced crushed feed to powder. Size reduction equipments is divided into following categories as follows:

### 2.2.1 CRUSHERS

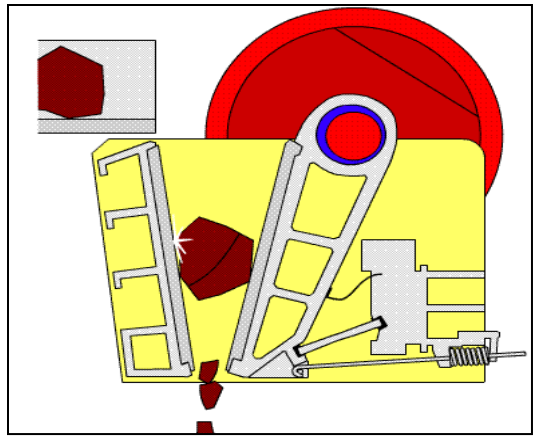
A crusher is a machine designed to reduce coarse particles into fine particles, gravel, or rock dust.

Crushers may be used to reduce the size, or change the form, of waste materials so they can be more easily disposed of or recycled, or to reduce the size of a solid mix of raw materials (as in rock ore), so that pieces of different composition can be differentiated. Crushing is the process of transferring a force amplified

by mechanical advantage through a material made of molecules that bond together more strongly, and resist deformation more, than those in the material being crushed do. Crushing devices hold material between two parallel or tangent solid surfaces, and apply sufficient force to bring the surfaces together to generate enough energy within the material being crushed so that its molecules separate from (fracturing), or change alignment in relation to (deformation), each other. The earliest crushers were hand-held stones, where the weight of the stone provided a boost to muscle power, used against a stone anvil. Querns and mortars are types of these crushing devices.

### 2.2.1.1 JAW CRUSHERS

A jaw or toggle crusher consists of a set of vertical jaws, one jaw being fixed and the other being moved back and forth relative to it by a cam or pitman mechanism, acting as a class II lever, like a nutcracker. The jaws are farther apart at the top than at the bottom, forming a tapered chute so that the material is crushed progressively smaller and smaller as it travels downward until it is small enough to escape from the bottom opening. The movement of the jaw can be quite small, since complete crushing is not performed in one stroke. The inertia required to crush the material is provided by a weighted flywheel that moves a shaft creating an eccentric motion that causes the closing of the gap.



Single and double toggle jaw crushers are constructed of heavy duty fabricated plate frames with reinforcing ribs throughout. Manganese steel is used for both fixed and movable jaw faces. Heavy flywheels allow crushing peaks on tough materials. Double Toggle jaw crushers may feature hydraulic toggle adjusting mechanisms.

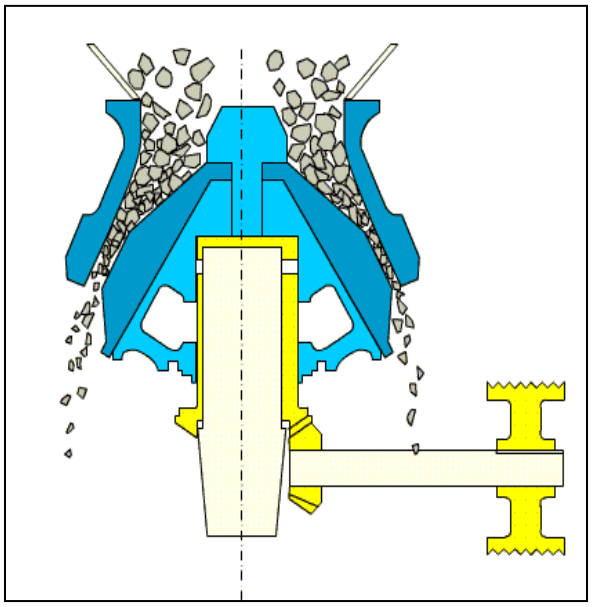
There are 3 types of jaw crushers according to the place the movable plate has been fixed around which position the rotates the movable jaw.

1. Blake crusher-fixed in the lower point
2. Dodge crusher-fixed in the upper point
3. Universal crusher-fixed in the midpoint

### 2.2.1.2 GYRATORY CRUSHERS

A gyratory crusher is similar in basic concept to a jaw crusher, consisting of a concave surface and a conical head; both surfaces are typically lined with manganese steel surfaces. The inner cone has a slight circular movement, but does not rotate; the movement is generated by

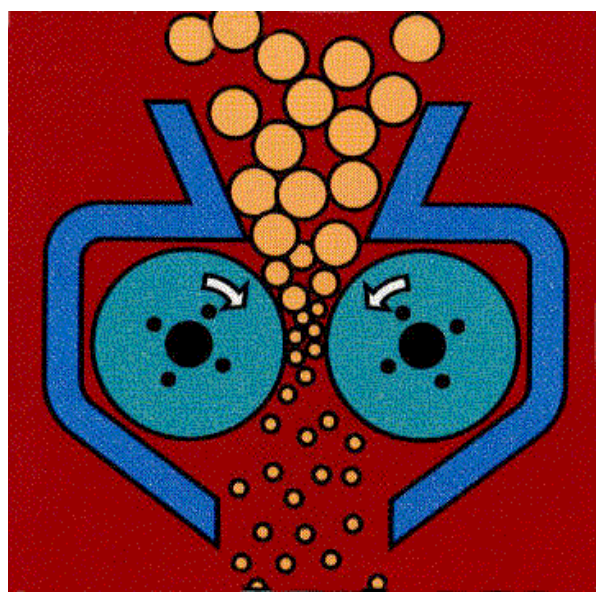
an eccentric arrangement. As with the jaw crusher, material travels downward between the two surfaces being progressively crushed until it is small enough to fall out through the gap between the two surfaces.



A gyratory crusher is one of the main types of primary crushers in a mine or ore processing plant. Gyratory crushers are designated in size either by the gape and mantle diameter or by the size of the receiving opening. Gyratory crushers can be used for primary or secondary crushing. The crushing action is caused by the closing of the gap between the mantle line (movable) mounted on the central vertical spindle and the concave liners (fixed) mounted on the main frame of the crusher. The gap is opened and closed by an eccentric on the bottom of the spindle that causes the central vertical spindle to gyrate. The vertical spindle is free to rotate around its own axis. The crusher illustrated is a short-shaft suspended spindle type, meaning that the main shaft is suspended at the top and that the eccentric is mounted above the gear. The short-shaft design has superseded the long-shaft design in which the eccentric is mounted below the gear.

### 2.2.1.3 ROLL CRUSHERS

The Roll Crushers are compression type crushers, and were once widely used in mining. They have, within the last 10 or so years, fallen into dis-favor among mining and processing companies. The probable reason is because the large mines require very large crushed product output with minimal cost, makes the roll crusher uncompetitive. The roll crushers are not nearly as productive as cone crushers, with respect to volume, and they do have a little higher maintenance associated with them. Roll crushers do, however, give a very close product size distribution, and if the ore is not too abrasive, they do not have high maintenance costs.



The particles are drawn into the gap between the rolls by their rotating motion and a friction angle formed between the rolls and the particle, called the nip angle. The two rolls force the particle between their rotating surface into the ever smaller gap area, and it fractures from the compressive forces presented by the rotating rolls. Some major advantages of roll crushers are they give a very fine product size distribution and they produce very little dust or fines. Rolls crushers are effectively used in minerals crushing where the ores are not too abrasive and they are also used in smaller scale production mining of more abrasive metal ores, such as gold. Coal is probably the largest user of roll crushers, currently, though. Coal plants will use roll crushers, either single roll or double roll, as primary crushers, reducing the ROM coal. Usually, these crushers will have teeth or raised forms on the face of the roll. (Roll crushers used for minerals and metal ores have smooth faced rolls.)

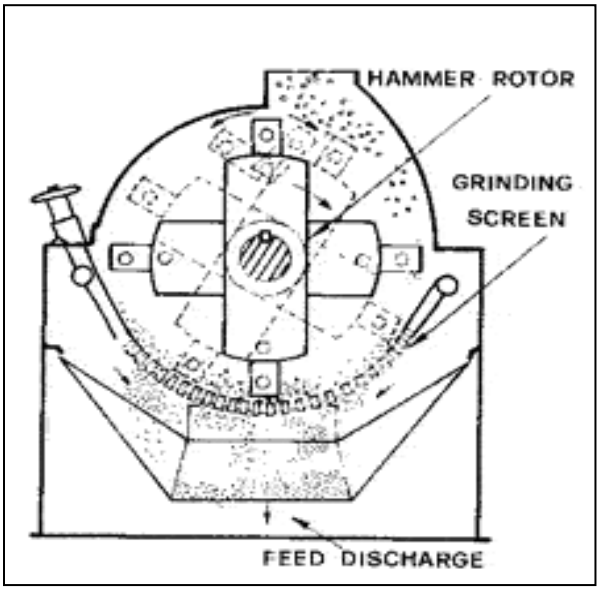
### 2.2.2 GRINDERS

Grinders are same as crushers but the only difference is that grinders reduce the fine particles size into very fine (powder) form, whereas the crushers reduced the coarse particle into fine particles.

There are following type of equipments used for grinding of fine particles:

#### 2.2.2.1 HAMMER MILLS

The hammer mill comprises of a heavy duty mild steel fabricated body. The grinding chamber of hammer mill is lined with serrated wear plates, which protects




the body from wear and tear. A rotor in hammer mill with a set of swing hammers accelerates the grinding process. The screen classifier forms the lower half of the grinding chamber of hammer mill. The blower in super type hammer mill is driven on a separate shaft with the help of a 'V' belt adjustable driven from the rotor shaft and the blower fan is mounted on same shaft of blower fan in economic type hammer mill. The hammer mill is especially designed for the course, and medium fine size reduction.

- **OPERATING PRINCIPLE OF HAMMER MILL**

The material to be crushed enters the hammer mill through gravity feed hopper having an adjustable slide to control the feed material. The material is crushed

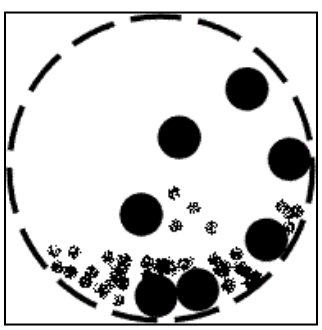
between the hammers and serrated liners. A powerful blower in hammer mill continuously sucks the ground material through a screen classifier and conveyed through the pipe into the cyclone for bagging. The blower maintains constant airflow in the hammer mill chamber in order to obtain a cool product and continuously cleans the screens, thus increasing the output. Particle size of the ground material can be varied over a large range by using sieves with the desired openings of hammer mill.

### 2.2.2.2 BALL MILLS

A ball mill is a type of grinder used to grind materials into extremely fine powder for use in mineral dressing processes, paints, pyrotechnics, and ceramics.

- **OPERATING PRINCIPLE**

The principle of size reduction in ball mills is impact of balls, which fall from the top of the shell on t the feed particles near the bottom of the shell.



- **CONSTRUCTION OF DEVICE**

A ball mill is a horizontal cylinder partly filled with steel balls (or occasionally other shapes) that rotates on its axis, imparting a tumbling and cascading action to the balls. Material fed through the mill is crushed by impact and ground by attrition between the balls. The grinding media are usually made of high-chromium steel. The smaller grades are occasionally cylindrical ("pebs") rather than spherical. There exists a speed of rotation (the "critical speed") at which the contents of the mill would simply ride over the roof of the mill due to centrifugal action.

The mill is usually divided into at least two chambers,(Depends upon feed input size presently mill installed with Roller Press are mostly single chambered), allowing the use of different sizes of grinding media. Large balls are used at the inlet, to crush clinker nodules (which can be over 25 mm in diameter). Ball diameter here is in the range 60–80 mm. In a two-chamber mill, the media in the second chamber are typically in the range 15–40 mm, although media down to 5 mm are sometimes encountered. As a general rule, the size of media has to match the size of material being ground: large media can't produce the ultra-fine particles required in the finished cement, but small media can't break large clinker particles.

A current of air is passed through the mill. This helps keep the mill cool, and sweeps out evaporated moisture which would otherwise cause hydration and disrupt material flow. The dusty exhaust air is cleaned, usually with bag filters.

- **CRITICAL SPEED OF BALL MILL**

The critical speed (rpm) is given by:

$$N_C = \frac{1}{2\pi}\sqrt{\frac{g}{(R-r)}}$$

Where $R$ is the internal radius of shell in meters and r is the radius of balls used in mill in meters.

Ball mills are normally operated at around 75% of critical speed, so a mill with diameter 5 meters will turn at around ~14 rpm.

## 2.3 POWER REQUIREMENTS FOR CRUSHING

The particles of feed material in size reduction equipments are first distorted and strained. The work needed to strained them is stored temporarily in the solid as mechanical energy of stress. As additional force is applied to the stressed particles, they are distorted beyond their ultimate strength and suddenly rupture into fragments. New surface is created. The creation of new surface requires work, which is supplied by the release of energy of stress when the particle breaks. All energy of stress in excess of the new surface energy created must appear as heat.

### 2.3.1 CRUSHING EFFICIENCY

Crushing efficiency can be defined as the ratio of the surface energy created by crushing to the energy absorbed by the solid. It is denoted by $\eta_c$ .

$$\boxed{\eta_c = \frac{e_s\left(A_{wb} - A_{wa}\right)}{W_n}}$$

Where, $e_s$ =surface energy per unit area

$A_{wb}$, $A_{wa}$ =area per unit mass of product and feed

$W_n$=energy absorbed by a unit mass of the feed material

The surface energy created by the fracture is small in comparison with the total mechanical energy stored in the material at the time of rupture and most of latter is converted into heat. Hence, crushing efficiencies are low. (Typically ranges from 0.06 to 1 %)

### 2.3.2 LAWS OF CRUSHING

There are three laws have been proposed to correlate the power requirement for crushing and grinding with the feed and product sizes. All the three involve modeling of the energy required actually to break the solids. The total energy required is expressed as a product of this computed energy and an efficiency, which is assumed to be independent of the size of the solids. These laws are as follows:

#### 2.3.2.1 RITTINGER'S LAW

It states that the work required in crushing is proportional to the new surface created. It is written as:

$$\frac{P}{\dot{m}} = K_r\left(\frac{1}{\bar{D}_{sb}} - \frac{1}{\bar{D}_{sa}}\right)$$

Where, $K_r$= Rittinger's Constant

$\bar{D}_{sb}$, $\bar{D}_{sa}$ =Volume surface mean diameters of product and feed, respectively

$\frac{P}{\dot{m}}$ =power required per unit feed rate

This law assumes constant crushing efficiency which, for a given machine and feed material, is independent of the sizes of feed and product, this law also assumes equal sphericities of feed and product and constant mechanical efficiency.

### 2.3.2.2 KICK'S LAW

It states that work required for crushing a given mass of material is constant for the same reduction ratio, i.e., the ratio of the initial particle size to the final particle size. This is written as:

$$\frac{P}{\dot{m}} = K_k \ln \frac{\bar{D}_{sa}}{\bar{D}_{sb}}$$

Where, $K_k$ = Kick's law Constant

This law is based on stress analysis of plastic deformation within the elastic limit.

### 2.3.2.3 BOND'S LAW

It states that the work required to form particles of size $D_p$ from a very large feed is proportional to the square root of the surface – to – volume ratio of the product. This law is written as

$$\frac{P}{\dot{m}} = \frac{K_b}{\sqrt{D_p}}$$

Where $K_b$ = bond's constant and depends upon the type of machine and on the material to be crushed.

- **WORK INDEX ($W_i$)**

The work index, $W_i$, is defined as the gross energy requirement in kWh per ton of feed needed to reduce a very large feed to such a size that 80 % of the product passes a 100 $\mu m$ screen. This definition of $W_i$ has been used to relate $K_b$ and $W_i$, for $D_p$ in mm, P in kW and $\dot{m}$ in tons per hour, the relation between $K_b$ and $W_i$ is given by

$$K_b = \sqrt{100 \times 10^{-3}} W_i = 0.3162 W_i$$

If 80 % product of the feed passes a mesh size of $D_{pa}$ mm and 80 % of the product passes a mesh size of $D_{pb}$ mm, then the above equations can be reduced to

$$\frac{P}{\dot{m}} = 0.3162 W_i \left( \frac{1}{\sqrt{D_{pb}}} - \frac{1}{\sqrt{D_{pa}}} \right)$$

The power given by above equation is gross power because the work index includes the friction in the crusher. The work index, for dry crushing or wet grinding is available for many materials in the literature. For dry grinding, the power calculated from above equation is multiplied by 4/3.

# GATE QUESTIONS

**Q.1** The energy required per unit mass to grind limestone particles of very large size to 100 μm is 12.7 k Wh/ton. An estimate (using Bond's Law) of the energy to grind the particles form a very large size to 50 μm is

(a) 6.25 k Wh/ton (b) 9.0 k Wh/ton

(c) 18 k Wh/ton (d) 25.4 k Wh/ton

**(GATE 2001)**

**Q.2** What is he critical rotation speed in revolutions per second, for a ball mill of 1.2 m diameter charged with 70 mm diameter balls?

(a) 0.5 (b) 1.0

(c) 2.76 (d) 0.66

**(GATE 2002)**

**Q.3** Energy requirement (per unit mass of material crushed/ground) is highest for

(a) Jaw crusher (b) rod mill

(c) ball mill (d) fluid energy mill

**(GATE 2003)**

**Q.4** The critical speed of the ball mill of radius R, which contains balls of radius r, is proportional to

(a) $(R - r)^{-0.5}$ (b) $(R - r)^{-1}$

(c) $(R - r)$ (d) $(R - r)^2$

**(GATE 2005)**

**Linked Answer Questions 5 & 6 :**

A continuous grinder obeying the Bond's crushing law grinds a solid at the rate of $10^3$ kg/hr from the initial diameter of 10 mm to the final diameter of 1 mm.

**Q.5** If the market now demands particles of size 0.5 mm, the output rate of the grinder (in kg/hr) for the same power input would be reduced to

(a) 227 (b) 474

(c) 623 (d) 856

**(GATE 2006)**

**Q.6** In order to restore the output back to 1000 kg/hr, an additional grinder was installed. The two grinders can be operated either in series (configuration – 1) or parallel (configuration – 2). Compare the two configurations in terms of the additional power consumption over the case above.

(a) Configuration-1 consumes less power than configuration-2

(b) Configuration-2 consumes less power than configuration-1

(c) Both configurations consume the same power,

(d) Configuration-2 consumes less or more power than configuration-1 depending on how the feed is distributed between the two grinders in configuration-2 (the parallel configuration).

**(GATE 2006)**

**Q.7** Size reduction of coarse hard solids using a crusher is accomplished by

(a) attrition (b) cutting
(c) compression (d) impact

**(GATE 2007)**

**Q.8** The power required for size reduction in crushing is

(a) $\propto \dfrac{1}{\text{surface energy of the matetrial}}$

(b) $\propto \dfrac{1}{\sqrt{\text{surface energy of the matetrial}}}$

(c) $\propto$ surface energy of the matetrial

(d) independent of the Surface energy of the material

**(GATE 2008)**

**Q.9** The critical speed (revolutions per unit time) of a ball mill of radius R, which uses balls of radius r, is

(a) $\frac{1}{2\pi}\sqrt{\frac{g}{\sqrt{Rr}}}$ (b) $\frac{1}{2\pi}\sqrt{\frac{g}{R}}$

(c) $\frac{1}{2\pi}\sqrt{\frac{g}{r}}$ (d) $\frac{1}{2\pi}\sqrt{\frac{g}{R-r}}$

**(GATE 2010)**

**Q.10** 100 ton/h of a rock feed, of which 80% passed through a mesh size of 2.54 mm, were reduced in size such that 80% of the crushed product passed through a mesh; size of 1.27 mm. The power consumption was 100 kW. If 100 ton/h of the same material is similarly crushed from a mesh size of 5.08 mm to a mesh size of 2.54 mm, the power consumption (in kW, to the nearest integer) using Bond's law, is _____

**(GATE 2013)**

**Q.11** In order to produce fine solid particles between 5 and 10 μm, the appropriate size reducing equipment is

(a) Fluid energy mill

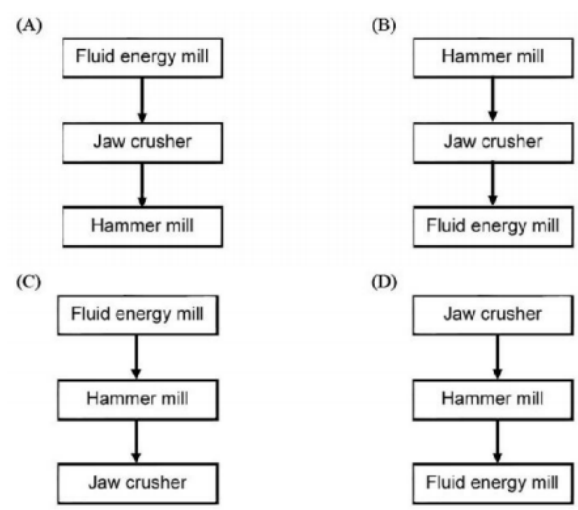
(b) Hammer mill

(c) Jaw crusher

(d) Smooth roll crusher

**(GATE 2014)**

**Q.12** Which of the following is the correct sequence for size reduction equipment




**(GATE 2017)**

**Q.13** Critical speed of a ball mill depends on

(a) the radius of the mill (shell) and the radius of the particles
(b) the radius of the mill (shell) and the density of the particles
(c) the radius of the balls and the radius of the particles
(d) the radius of the balls and the radius of the mill (shell)

**(GATE 2018)**

| ANSWER KEY | | | | | | | | | |
|---|---|---|---|---|---|---|---|---|---|
| **1** | **2** | **3** | **4** | **5** | **6** | **7** | **8** | **9** | **10** |
| (c) | (d) | (d) | (a) | (c) | (a) | (c) | (c) | (d) | 70.71 |
| **11** | **12** | **13** | | | | | | | |
| (a) | (d) | (d) | | | | | | | |

# EXPLANATIONS

**Q.1 (c)**

According to bond's law , $E = \frac{K}{\sqrt{D_p}}$

Where

E- energy required to reduce a unit weight feed from infinite size of particle. $D_p$

$$\frac{E_2}{E_1} = \sqrt{\frac{D_{p_1}}{D_{p_2}}} \qquad E_2 = 12.7 \times \sqrt{\frac{100}{50}}$$

$$E_2 = 18 \text{ kWh / ton}$$

**Q.2 (d)**

Critical speed for ball mill

$$N_C = \frac{1}{2\pi}\sqrt{\frac{g}{R-r}}$$

$$= \frac{1}{2\pi}\sqrt{\frac{9.81}{\frac{1.2}{2} - \frac{70}{2} \times 10^{-3}}}$$

$$= 0.663 \text{ rps}$$

**Q.3 (d)**

Fluid energy mill: fluid energy mill are used for size reduction, drying, blending coating and chemical reaction involving at least one solid material. There is no moving parts and no grinding media. Compressed air, gas or high pressure super heated steam is used to run the mill. Energy requirement is higher for this type of mill

**Q.4 (a)**

Let m is mass of a ball,

$n_c \rightarrow$ critical speed

At critical speed

Gravitational force = centrifugal force

$mg = m(R-r)(w_c)^2$

$$w_c = \sqrt{\frac{g}{(R-r)}}$$

$$n_c = \frac{1}{2\pi}\sqrt{\frac{g}{(R-r)}}$$

$$n_c \,\alpha\, (R-r)^{-0.5}$$

**Q.5 (c)**

According to Bond's Law, E= 100 E,

$$\left(\frac{1}{\sqrt{X_P}} - \frac{1}{\sqrt{X_F}}\right)$$

Where E= work done,

$E_i$ = Bond work index,

$X_P$ = Product size,

$X_F$ = Feed size

for initial conditions,

$$E_1 = 100E_i\left(\frac{1}{\sqrt{1}} - \frac{1}{\sqrt{10}}\right) \quad ....(i)$$

and $$E_2 = 100E_i\left(\frac{1}{\sqrt{0.5}} - \frac{1}{\sqrt{10}}\right) \quad ....(ii)$$

Dividing equations (i) by equation (ii), we get

$$\frac{E_1}{E_2} = \frac{1-0.316}{1.414-0.316} = 0.623$$

$\therefore$ New output rate of grinder =623 kg/hr

**Q.6 (a)**

**Q.7 (c)**

Impact is the primary action in hammer mill attrition in ball mill, cutting action in cutters.

**Q.8 (c)**

**Q.9 (d)**

**Q.10 70.71**

The Bond's crushing law is given as

$$W = \frac{P}{\dot{m}} = K_b\left(\frac{1}{\sqrt{\bar{D}_{sb}}} - \frac{1}{\sqrt{\bar{D}_{sa}}}\right)$$

Where in case 1:

$P = 100$ kW, $\dot{m} = 100$ ton / h,

$\bar{D}_{sb} = 1.27$ mm *and* $\bar{D}_{sa} = 2.54$ mm

Thus, $$\frac{100}{100} = K_b\left(\frac{1}{\sqrt{1.27}} - \frac{1}{\sqrt{2.54}}\right)$$

$$\Rightarrow \quad K_b = 3.847623279$$

And in case 2:

$P = ?$, $\dot{m} = 100$ ton / h, $\bar{D}_{sb} = 2.54$ mm

*and* $\bar{D}_{sa} = 5.08$ mm

Therefore

$$\frac{P}{100} = K_b\left(\frac{1}{\sqrt{2.54}} - \frac{1}{\sqrt{5.08}}\right)$$

$$\Rightarrow P = 100 \times 3.847623279 \times \left(\frac{1}{\sqrt{2.54}} - \frac{1}{\sqrt{5.08}}\right)$$

$$\Rightarrow \boxed{P = 70.71 \text{ kW}}$$

**Q.11 (a)**

Fluid energy mill is used for getting ultrafine particles while jaw crusher is used for producing bigger products.

**Q.12 (d)**

**Q.13 (d)**

Critical speed of the ball mill is given by

$$n_c \,\alpha\, \sqrt{\frac{g}{R-r}}$$

Where R is radius of ball mill

r is radius of impact balls

# 3 AGITATION AND MIXING

## 3.1 INTRODUCTION

Dispersion of one component through the other is known as mixing, which can be of different solid components, pastes and liquids. Mixing is an unit operation in which a random distribution takes place between two or more initially separate phases (miscible or immiscible) or the elements of a single phase material with temperature or concentration gradients. It is one of the important unit operations practised in solid – as well as liquid – processing industries. Thus, the various mixing operations in chemical industries can be classified into two categories: *liquid mixing* and *solid mixing.*

The main objective of solid operation include

(i) Reduction in in–homogeneity in the properties of bulk materials
(ii) To promote heat or mass transfer
(iii) To promote chemical reaction

## 3.2 LIQUID MIXING

The objective of liquid mixing is to obtain a relatively uniform mixture from two or more components, both miscible and immiscible. The degree of uniformity obtained depends on the liquid characteristics. While it is possible to obtain an almost completer homogeneity in case of the miscible liquids, in case of the immiscible ones, the minor component is generally present as the dispersed phase in a continuous phase of the major component.

For high mixing, it is necessary to have an effective agitation of the components in the containing vessel. *Agitation* refers to the induced motion of a material in a specified manner, usually in the circulatory pattern inside some form of container. Agitation is generally accomplished by using mechanical impeller creates a flow pattern in the system, causing the liquid to circulate through the vessel and return eventually to the impeller.

The mechanical impellers broadly belong to two categories:

(i) Axial Flow Impeller

(ii) Radial Flow Impeller

## 3.3 AXIAL FLOW IMPELLERS

The blades of an axial flow impeller make an angle equal to or less than $90^0$ to the driving shaft to produce currents in the liquid primarily parallel to the impeller shaft. Propellers and fan turbines belong to this class. Propellers are widely used for agitation liquids of low viscosity with speeds varying from 400 to 1750 rpm.

### 3.3.1 RADIAL FLOW IMPELLERS

Radial flow impellers with blades parallel to the axis of the drive shaft produce currents in radial or tangential direction in the liquid. Paddles and turbines belong to this type of impeller. Normally, turbines have a number of short blades and operate at high speed while paddles are large slower – speed impellers with two or four blades.

## 3.4 MIXING EQUIPMENTS

Mixtures of solids and liquids are blended in different types of equipment depending on the physical characteristics of the mixture. The mixing equipment, hereafter mixers, are generally of three types: liquid mixers, solid mixers and paste or viscous mixers. Pumpable suspensions with thin consistency are normally handled in tanks agitated with an impeller or fluid jet while non-flowing pastes are handled in slow speed non-circulating mixers.

### 3.4.1 LIQUID MIXERS

Mixing of miscible liquids, dispersing immiscible liquids, heat transfer in agitated liquid, suspension of solids in liquids, etc. are generally carried out in agitated vessels by using mechanical impellers, which are broadly classified into two types: axial and radial.

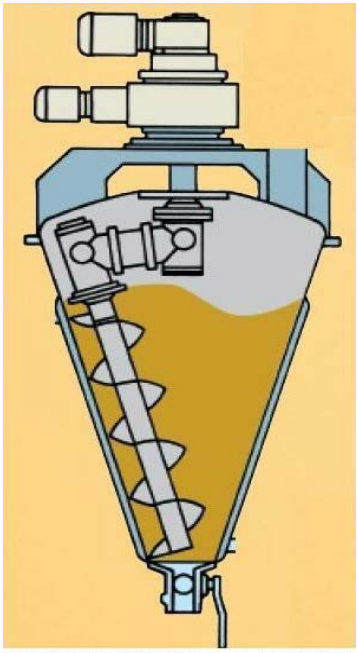
In *axial flow impellers*, the impeller blade makes an angle equal to or less than $90^0$ with the plane of impeller rotation. As a result, the locus of flow occurs along the axis of the impeller (parallel to the impeller shaft), e.g., marine propellers and pitched blade turbine.

While in *radial flow impellers,* the impeller blade is parallel to the axis of the impeller and as a result the radial flow impeller discharges flow along the radius in distinct patterns, e.g., flat blade turbine, paddle and anchor.

### 3.4.2 SOLID MIXERS

The mechanism of solid mixing, known as blending, is generally based on diffusion and convection. *Diffusion blending* is characterized by small scale random motion of solid particles, whereas, *convection blending* is characterized by large scale motion of solid particles.

Diffusion blending occurs where the particles are distributed over a freshly developed interface. Tumbler blenders like the V-blenders and double cone blenders function by diffusion mixing. In convection blending, groups of particles are rapidly moved from one position to another due to the action of a



VERTICAL CONE SCREW BLENDER

rotating agitator or cascading of material within a tumbler blender. The blending of solids in ribbon blenders and vertical cone screw blenders is mainly due to convection mechanism.

**MIXERS**

Mixing of heavy pastes, dough, plastic masses and rubbery products require heavy-duty machines, which involve stretching, folding and compression of the masses many times before the final mixing is affected. The popular viscous mixers are the kneading machines, known as *kneaders,* which are slow – speed machines requiring high energy.

## 3.4 POWER CONSUMPTION

The power consumption in mixing of low viscosity, Newtonian fluids can be given as:

$$\boxed{\frac{P}{\rho N^3 D_i^5} = f\left(\frac{\rho N D_i^2}{\mu}, \frac{N^2}{D_i}, \frac{D_V}{D_i}, \frac{W_i}{D_V}, \ldots\right)}$$

Where,

$P$ = Power consumed, $W$

$\rho$ = Density of mixture, $kg / m^3$

$N$ = Speed of impeller, $rps$

$D_i$ = diameter of impeller, $m$

$\mu$ = viscosity of fluid mixture, $Pa \cdot s$

$D_V$ = vessel diameter, $m$

$W_i$ = vessel height, $m$

And

$$\frac{P}{\rho N^3 D_i^5} = N_{Po} = \text{Power number}$$

$$\frac{\rho N D_i^2}{\mu} = N_{\text{Re}} = \text{Reynolds number for mixing}$$

$$\frac{N^2}{D_i} = N_{Fr} = \text{Forude number}$$

# GATE QUESTIONS

**Q.1** To keep the power input constant for a stirred vessel operating under fully developed turbulent flow conditions (constant power number), if the impeller diameter is increased by 20%, the impeller speed should be decreased by a factor of

(a) $(1.2)^{3/2}$ (b) $(1.2)^{3/5}$

(c) $(1.2)^{2/3}$ (d) $(1.2)^{5/3}$

**(GATE 2004)**

**Q.2** If the frequency of the stirrer in a mixing tank is increased by a factor of 2 while all other parameters are kept constant, by what factor is the power requirement increased at high Reynolds number?

(a) 4 (b) 8

(c) 16 (d) 32

**(GATE 2005)**

**Q.3** The mixing of rubber latex solution was studied in an un-baffled mixer in the laboratory. The mixer was equipped with a six blade turbine impeller. A tyre company scales this process up using a baffled tank. The baffled tank has 3 times the diameter of the lab scale mixer. It uses the same type of impeller operated at the same speed. The relevant shape factors are also the same. Assuming that laminar conditions prevail in both cases, the power requirement in the industrial scale mixer is

(a) is 3 times that of the lab scale mixer

(b) is 9 times that of the lab scale mixer

(c) is 27 times that of the lab scale mixer

(d) cannot be estimated reliably due to the presence of baffles.

**(GATE 2006)**

**Q.4** Consider the scale-up of a cylindrical baffled vessel configured to have the standard geometry (i.e. Height = Diameter). In order to maintain an equal rate of mass transfer under turbulent conditions for a Newtonian fluid, the ratio of the agitator speeds should be (Given $N_1$, $D_1$ are agitator speed and vessel diameter before scale-up; $N_2$, $D_2$ agitator speed and vessel diameter after scale-up)

(a) $\frac{N_1}{N_2} = \frac{D_1}{D_2}$ (b) $\frac{N_1}{N_2} = \frac{D_2}{D_1}$

(c) $\frac{N_1}{N_2} = \left(\frac{D_1}{D_2}\right)^{2/3}$ (d) $\frac{N_1}{N_2} = \left(\frac{D_2}{D_1}\right)^{2/3}$

**(GATE 2008)**

**Q.5** For a mixing tank operating in the laminar regime, the power number varies with the Reynolds number (Re) as

(a) $Re^{-1/2}$ (b) $Re^{1/2}$

(c) Re (d) $Re^{-1}$

**(GATE 2009)**

**Q.6** In a mixing tank operating at very high Reynolds number (> $10^4$), if the diameter of the impeller is doubled (other conditions remaining constant), the power required increases by a factor of

(a) 1 / 32 (b) 1 / 4

(c) 4 (d) 32

**(GATE 2012)**

**Q.7** An agitated cylindrical vessel is fitted with baffles and flat blade impellers. The power number for this system is given by $N_p = \frac{P}{\rho n^3 D^5}$ where P is the power consumed for the mixing, ρ is the density of the fluid, n is the speed of the impeller and D is the diameter of the impeller. The diameter of the impeller is 1/3rd the diameter of the tank and the height of liquid level is equal to the tank diameter. The impeller speed to achieve the desired degree of mixing is 4 rpm. In a scaled up design, the linear dimensions of the equipment are to be doubled, holding the power input per unit volume constant. Assuming the liquid to be Newtonian and to be independent of Reynolds number, what is the impeller speed (in rpm) to achieve the same degree of mixing in the scaled up vessel?

(a) 0.13 (b) 1.26

(c) 2.52 (d) 3.82

**(GATE 2016)**

**Q.8** A propeller (diameter D = 15 M) rotates at N = 1 revolution per second (rps). To understand the flow around the propeller, a lab scale model is made. Important parameters to study the flow are velocity of the propeller tip $(V = \pi ND)$, diameter D and acceleration due to gravity (g) . The lab scale model is 1/100$^{th}$ of the size of the actual propeller.

The rotation speed of the lab scale model, to the nearest integer, should be _____rps

**(GATE 2017)**

| ANSWER KEY | | | | | | | |
|---|---|---|---|---|---|---|---|
| 1 | 2 | 3 | 4 | 5 | 6 | 7 | 8 |
| (d) | (b) | (c) | (d) | (d) | (d) | (c) | 10 |

# EXPLANATIONS

**Q.1 (d)**

Power Number, $N_P = \frac{P}{\rho N^3 D^5}$

Given, $N_\rho, P, \rho$ are constants.

$$N_1^3 D_1^5 = N_2^3 D_2^5$$

Given $D_2 = 1.2D_1$

$$\Rightarrow \frac{N_2}{N_1} = \left(\frac{D_1}{D_2}\right)^{5/3}$$

$$\Rightarrow N_2 = \frac{N_1}{(1.2)^{5/3}}$$

**Q.2 (b)**

We know that

Power No., $N_P = \frac{P}{\rho N^3 D^5}$

Given, $N_\rho, D, \rho$ are constants.

$P \propto N^3$

Given that N increases by two times

Therefore power requirement will increase by 8 times.

**Q.3 (c)**

Power No. = $\frac{P}{\rho N^3 \cdot D^5}$

$P \alpha N^3$

$$\frac{p_2}{p_1} = \left(\frac{N_2}{N_1}\right)^3$$

$$p_2 = p_1 \times 3^3$$
$$= 27p_1$$

**Q.4 (d)**

In turbulent regime, Power no.,

$$N_{Po} = \frac{P}{\rho N^3 D_i^5} = \text{constant}$$

To keep equal rate of mass transfer, power per unit volume must be constant,

$$\frac{P}{V} = \frac{N_{Po}\rho N^3 D_i^5}{\pi D^3/4} = \text{constant} \qquad \left\{\begin{matrix}\text{For tank of } H=D \\ V=\pi D^3/4\end{matrix}\right\}$$

$$\Rightarrow \frac{4N_{Po}\rho N^3 D_i^5}{\pi D^3} = \text{constant}$$

$$\Rightarrow \frac{N^3 D_i^5}{D^3} = \text{constant}$$

$$\Rightarrow \frac{N_2^3}{N_1^3} = \left(\frac{D_i^5}{D^3}\right)_1 \Big/ \left(\frac{D_i^5}{D^3}\right)_2$$

For scale up, all geometrical dimensions are scaled up with same ratio, thus

$$\frac{D_{i2}}{D_{i1}} = \frac{D_2}{D_1}$$

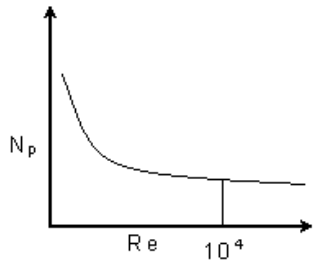
Therefore, $\Rightarrow \frac{N_2^3}{N_1^3} = \left(\frac{D_1}{D_2}\right)^2 \quad \Rightarrow \frac{N_2}{N_1} = \left(\frac{D_1}{D_2}\right)^{2/3}$

**Q.5 (d)**

**Q.6 (d)**

At high Reynolds number, the power number is nearly remains constant.

The Re v/s $N_p$ curve can be plotted as below




We know that power number is given by

$$N_p = \frac{P}{\rho N^3 D_a^5}$$

$$\therefore \quad \frac{P_1}{\rho N_1^3 D_{a1}^5} = \frac{P_2}{\rho N_2^3 D_{a2}^5}$$

$\because \rho = \text{Constant}$ and $N_1 = N_2$

$$\Rightarrow \frac{P_2}{P_1} = \left(\frac{D_{a2}}{D_{a1}}\right)^5 = 2^5 = 32$$

$$\Rightarrow P_2 = 32\ P_1$$

**Q.7 (c)**

We have,

$$\text{Power No} = \frac{P}{P\ N^3\ D_i^5}$$

For constant power no.

$$\frac{P}{PN^3\ D_i^5} = \text{const}$$

$$P \propto N^3\ D_i^5 \qquad \text{(I)}$$

$$\text{Volume of tank} = \frac{\pi}{4} D_V^2 H$$

$$D_V = H \qquad \text{(given)}$$

$$\Rightarrow V \propto D_V^3 \qquad (II)$$

From equation (I) and (II)

$$\frac{P}{V} = \frac{N^3\ D_i^5}{D_v^3}$$

$$\text{For same } \frac{P}{V} = \frac{N_1^3\ D_{i_1}^5}{D_{v1}^3} = \frac{N_2^3\ D_{i_2}^5}{D_{V_2}^3}$$

If all linear dimensions have to be doubled

$$\frac{D_{i_2}}{D_{i_1}} = \frac{D_{V_2}}{D_{V_1}} = 2$$

$$\Rightarrow N_2^3 = N_1^3 \left(\frac{D_{V_1}}{D_{V_2}}\right)^2$$

$$\Rightarrow N_2 = N_1 \left(\frac{D_{V_1}}{D_{V_2}}\right)^{2/3} \qquad (N_1 = 4\,\text{rpm})$$

$$\Rightarrow N_2 = 4 \times \left(\frac{1}{2}\right)^{2/3} = 2.52\ \text{rpm}$$

**Q.8 10**

# 4 SOLID FLUID SEPARATION

Solid – Fluid separation consists of separation of solids from liquids and gases through various methods.

Solid – liquid separation is an important unit operation, used for the recovery and processing of solids or for purification of liquids or for separating the two phases for environmental reasons.

## 4.1 SEDIMENTATION

Sedimentation, also known as settling, may be defined as the removal of solid particles from a suspension by settling under gravity.

Clarification is a similar term, which usually refers specifically to the function of a sedimentation tank in removing suspended matter from the water to give a clarified effluent. In a broader sense, clarification could include flotation and filtration.

Thickening in sedimentation tanks is the process whereby the settled impurities are concentrated and compacted on the floor of the tank and in the sludge-collecting hoppers.

Concentrated impurities withdrawn from the bottom of sedimentation tanks are called *sludge*, while material that floats to the top of the tank is called *scum*.

### 4.1.1 APPLICATION OF SEDIMENTATION PROCESSES

In water treatment, sedimentation is commonly used to remove impurities that have been rendered settle able by coagulation and flocculation, as when removing turbidity and color. Precipitates formed in processes such as water softening by chemical precipitation are also removed by sedimentation.

In municipal wastewater treatment, sedimentation is the main process in primary treatment, where it is responsible for removing 50 to 70% of the suspended solids (containing 25-40 per cent of the BOD) from the wastewater. After biological treatment, sedimentation is used to remove the biological floc produced by microorganisms in these processes, so that effluent quality will approach a standard suitable for discharge into inland waterways. The removal of grit in the preliminary stage of treatment is commonly carried out by means of a differential sedimentation process in which heavy grit is permitted to settle while lighter organic matter is retained in suspension. Further sedimentation after

coagulation may be used in tertiary treatment.

Sedimentation is also required where phosphorus removal is effected by chemical precipitation separately from primary or secondary treatment. Other less obvious applications of sedimentation are in the separation of digested sludge from supernatant liquor within secondary (unstirred) sludge digesters, and also in sludge lagoons.

An understanding of the principles governing the various forms of sedimentation behavior is essential to the effective design and operation of sedimentation tanks.

### 4.1.2 CLASSIFICATION OF SETTLING BEHAVIOR

Several cases of settling behaviour may be distinguished on the basis of the *nature* of the particles to be removed and their *concentration*. Thus, individual particles may be discrete (sand grains) or flocculent (most organic materials and biological solids). Particle concentrations may vary from very low through to high in which case adjacent particles are actually in contact. Common classifications of settling behaviour are:

Class I - Unlimited settling of discrete particles

Class II - Settling of dilute suspensions of flocculent particles

Class III - Hindered settling and zone settling

Class IV - Compression settling (compaction).

#### 4.1.2.1 SEDIMENTATION CLASS I - UNLIMITED SETTLING *OF* DISCRETE PARTICLES

Sedimentation is removal of discrete particles in such low concentration that each particle settles freely without interference from adjacent particles (that is, unhindered settling).

When a particle settles in a fluid it accelerates until the drag force due to its motion is equal to the submerged weight of the particle. At this point, the particle will have reached its terminal velocity, $V_p$.

A diagram for settling of an idealized spherical particle is shown below in Figure 4.1). $V_p$ is the particle settling velocity (m/s); D is the drag force; W is the submerged weight of the particle; d is the diameter of the particle (m); $A_p$ is the projected area of the particle normal to the direction of motion ($m^2$); $V_p$ is the volume of the particle ($m^3$); $\rho$ is the density of the particle ($kg/m^3$); $\rho_p$ is the fluid density ($kg/m^3$); $\mu$ is the dynamic viscosity of the fluid ($N.s/m^2$); and $C_D$ is the drag coefficient.

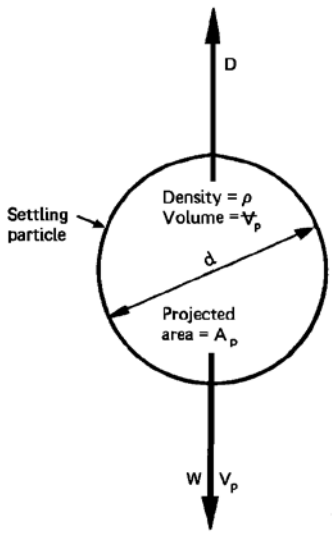



**Figure 4.1 Definition diagram for particle terminal settling velocity**

An expression for $V_p$ may be derived from the submerged weight of the settling particle, W, and the fluid drag force, D.

The drag force on a particle is given by

$$D = C_D \rho_l A_p v_s^2 / 2$$

The submerged weight of the particle can be expressed as

$$W = (\rho - \rho_l) g\, V_p$$

Since D = W, the above, after substituting $A_p$ and $\forall_p$ for functions in terms of particle diameter d and rearranging, results in the following expression for $v_s$.

$$V_s = \sqrt{\frac{4}{3}\frac{(\rho - \rho_l)}{\rho_l}\frac{gd}{C_D}}$$

In practice, it is found that $C_D$ is a function of the Reynolds Number, $R_e$, and, for spherical particles, it can be represented by the following expressions

$$R_e < 1, C_D = \frac{24}{R_e}$$

$$1 < R_e < 10^3, C_D = \frac{24}{R_e} + 0.44$$

$$10^3 < R_e < 10^5, C_D \approx 0.44$$

Substituting the above expression for $R_e$ < 1 (laminar flow) in Equation 2.3 and noting that

$R_e = \rho_l v_s d / \mu$, results in the following equation, known as Stoke's Law:

$$Vs = \frac{g}{18}\frac{(\rho - \rho_l)}{\mu} d^2$$

At high values of $R_e$, where $C_D \approx 0.4$, the equivalent expression is

$$V_s = \sqrt{3.33\frac{(\rho - \rho_l)}{\rho_l} gd}$$

The general conclusion, that $V_s$ depends on a particular diameter, particle density and, under some conditions, also on fluid viscosity and hence on temperature, is important in understanding sedimentation behavior. Furthermore, in practical sedimentation tanks, the terminal settling velocity is quickly reached, so, for non-flocculent particles and uniform fluid flow the settling velocity is constant throughout the settling time.

This fact can be usefully applied to a study of settling in an *ideal sedimentation tank* to provide an important design principle for sedimentation processes.

The ideal rectangular horizontal flow sedimentation tank is considered divided into four zones

a) Inlet zone - in which momentum is dissipated and flow is established in a uniform forward direction

b) Settling zone - where quiescent settling is assumed to occur as the water flows towards the outlet

c) Outlet zone - in which the flow converges upwards to the decanting weirs or launders

d) Sludge zone - where settled material collects and is moved towards sludge hoppers for withdrawal. It is assumed that once a particle reaches the sludge zone it is effectively removed from the flow.

The critical particle in the settling zone of an ideal rectangular sedimentation tank, for design purposes, will be one that enters at the top of the settling zone, at point A, and settles with a velocity just sufficient to reach the sludge zone at the outlet end of the tank, at point B. The velocity components of such a particle are $V_h$ in the horizontal direction and $V_s$, the terminal settling velocity, in the vertical direction.

From the geometry of the tank it is apparent that the time required for the particle to settle, $t_o$, is given by

$$t_o = \frac{H}{V_p} = L / V_s$$

but, since $V_s = Q/WH$, then $V_s = Q/WL$, where Q is the rate of flow, and L, W and H are the length, width and depth of the tank, respectively. Since the surface area of the tank, A, is WL, then

$$V_s = Q/A$$

According to this relationship, the slowest-settling particles, which could be expected to be completely removed in an ideal sedimentation tank would have a settling velocity of Q/A. Hence this parameter, which is called the *surface loading rate* or *overflow rate*, is a fundamental parameter governing sedimentation tank performance.

This relationship also implies that *sedimentation efficiency is independent of tank depth* - a condition that holds true only if the forward velocity is low enough to ensure that the settled material is not scoured and re-suspended from the tank floor.

In an *ideal upflow sedimentation tank*, it is apparent that a particle will be removed only if its settling velocity exceeds the water upflow velocity. In this case the minimum upflow velocity is given by the flow rate divided by the surface area of the tank (Q/A), so once again the minimum

settling velocity for a particle to be removed is $V_s = Q/A$.

In an ideal sedimentation tank with a horizontal or radial flow pattern, particles with settling velocities < $V_s$ can still be removed partially, but not in an ideal upflow tank.

#### 4.1.2.2 SEDIMENTATION CLASS II SETTLEMENT OF FLOCCULENT PARTICLES IN DILUTE SUSPENSION

It should be recognized that particles do collide and that this benefits flocculation and hence sedimentation. In a horizontal sedimentation tank, this implies that some particles may move on a curved path while settling faster as they grow rather than following the diagonal line in Figure 2.5a. This favors a greater depth as the longer retention time allows more time for particle growth and development of a higher ultimate settling velocity. However, if the same retention time were spread over a longer, shallower tank, the opportunity for collision would become even greater because the horizontal flow rate would become more active in promoting collisions. In practice, tanks need to have a certain depth to avoid hydraulic *short-circuiting* and are made 3-6 m deep with retention times of a few hours.

The advantage of low depths is exploited in some settling tanks by introducing baffles or tubes. These are installed at an angle, which permits the settled sludge to slide down to the bottom of the settler, even though any angle effectively increases the vertical displacement between two plates.

#### 4.1.2.3 SEDIMENTATION CLASS III - HINDERED SETTLING AND ZONE SETTLING AND SLUDGE BLANKET CLARIFIERS

As the concentration of particles in a suspension is increased, a point is reached where particles are so close together that they no longer settle independently of one another and the velocity fields of the fluid displaced by adjacent particles, overlap. There is also a net upward flow of liquid displaced by the settling particles. This results in a reduced particle-settling velocity and the effect is known as *hindered settling*.

The most commonly encountered form of hindered settling occurs in the extreme case where particle concentration is so high that the whole suspension tends to settle as a 'blanket'. This is termed *zone settling*, because it is possible to distinguish several distinct zones, separated by concentration discontinuities. Figure represents a typical batch-settling column test on a suspension exhibiting zone-settling characteristics. Soon after leaving such a suspension to stand in a settling column, there forms near the top of the column a clear interface separating the settling sludge mass from the clarified supernatant. This interface moves downwards as the suspension settles. Similarly, there is an interface near the bottom between that portion of the

suspension, which has settled and the suspended blanket. This interface moves upwards until it meets the upper interface, at which point settling of the suspensions is complete.

It is apparent that the slope of the settling curve at any point represents the settling velocity of the interface between the suspension and the clarified supernatant. This once again leads to the conclusion that in designing clarifiers for treating concentrated suspensions (Class III), the surface loading rate is a major constraint to be considered; unless the surface loading rate adopted is less than the zone-settling velocity ($v_{sz}$) of the influent suspension, solids will be carried over in the effluent.

An important application of zone settling is the final design of sedimentation tanks of activated sludge processes.

Hindered settling is also important in upflow clarifiers in water treatment. These units often operate with a high concentration of solids (consisting of chemically-formed floc and impurities) in suspension. By simple comparison with the ideal upflow sedimentation tank in figure, it may be seen that a suspension will be retained in an upflow clarifier only if the settling velocity of the suspension interface, $v_i$, is equal to the upflow velocity of the water, $v_u$. This is important because many practical settling tanks are designed to maintain a high concentration of solids in suspension in order to take advantage of the increased opportunity for particles to collide and agglomerate. This assists in removing many of the very fine particles that might otherwise be carried over in the effluent.

In both these cases involving hindered settling of concentrated suspensions, the settling velocity of the suspension, $v_p$ is dependent on its concentration; as this increases, $v_p$ decreases.

The relationship between the velocity of settling $v_p$ and the volumetric concentration of the particles in the suspension has not yet been determined analytically for hindered settling situations. It is therefore necessary to use empirical equations to define an approximate relationship.

Many equations for this relationship have been proposed. The combined advantages of mathematical simplicity and reasonable accuracy over a wide range of concentrations are features of the empirical equation proposed by Richardson:

$$v_p = v_s.\varepsilon^n$$

where $\varepsilon = (1 - c)$, the porosity of the suspension; c is the proportion of the total suspension volume occupied by particles; and n is an index depending on the Reynolds number and the size and shape of the particles.

For smooth spheres, the value of n varies from 4.65, for fully laminar flow conditions, to about 2.5 for turbulent flow

conditions around the particles. For irregular particles, it is impracticable to determine volumetric concentration. The relationship may be modified to

$$v_p = v_s\left(1 - kc'\right) \qquad [2.7]$$

where c' is the concentration in mass units; k and n are parameters so chosen that the resulting formula closely approximates the performance of the particular suspension.

Suitable values of k and n may be selected by plotting experimental values of log $v_p$ against values of log (1 - kc') for different selected values of k until a value is found to give approximation to a straight line over an appropriate range. The corresponding value of n may be calculated from the slope of the line. The value of $v_s$ is given by the extrapolated value of V when c = 0. Computer techniques can be useful in selecting (or fitting) appropriate values of k and n to give a least-squares best fit.

#### *4.1.2.4 SEDIMENTATION CLASS IV - COMPRESSION SETTLING (COMPACTION)*

Very high particle concentrations can arise as the settling particles approach the floor of the sedimentation tanks and adjacent particles are actually in contact. Further settling can occur only by adjustments within the matrix, and so it takes place at a reducing rate. This is known as *compression settling* or *consolidation* and is illustrated by the lower region of the zone-settling diagram (Fig. 2.6). Compression settling occurs as the settled solids are compressed under the weight of overlying solids, the void spaces are gradually diminished and water is squeezed out of the matrix.

Compression settling is important in gravity thickening processes. It is also particularly important in activated-sludge final settling tanks, where the activated sludge must be thickened for recycling to the aeration tanks and for disposal of a fraction of the sludge.

### 4.1.3 DESIGN OF SEDIMENTATION TANKS

Sedimentation theory, predicts that, in the case of ideal settling, the main design parameter to be considered is surface loading rate, Q/A, because it represents the critical particle settling velocity for complete removal. Practical Class II settling likewise requires that adequate depth, H, or detention time, t, be provided in order to allow flocculation to take place. Uniform flow distribution cannot always be assumed in practice owing to density currents, inadequate dissipation of momentum at the tank inlet and drawdown effects at the effluent weirs. As a result of all these effects, surface loadings and detention times derived from theory should be multiplied by a suitable safety factor, typically 1.7 to 2.5, for practical design.

These considerations apply to all three types of tank commonly used for Class II sedimentation, namely rectangular

horizontal flow tanks, circular radial flow tanks and square upflow tanks.

In the case of Class III sedimentation, it was also shown that the surface loading rate is the major parameter to be considered in design. Most of the following development of theory therefore applies to the design of both Class II and Class III sedimentation tanks.

The design of sedimentation tanks for a given flow rate Q, involves the selection of the surface loading rate, Q/A, from which the required tank surface area may be calculated, and either tank depth, H, or detention time, t. The relationships between the various parameters concerned can be expressed as shown below.

For Q in $m^3/h$ and A in $m^2$, the particle settling velocity, $V_p$ (m/h) is given by

$$V_p = Q/A$$

Detention time (hours) is

$\frac{AH}{Q}$, where H is depth (m)

The task of proportioning the tank, once values of the major parameters are chosen, can be simplified by using a simple design chart based on the above equations. Alternative designs may be quickly compared using this diagram, and effects of flow variations on critical loading parameters be determined.

The *forward velocity* must also be considered in *rectangular tanks*, as excessive velocity may result in the scouring and re-suspension of settled sludge. This requirement influences the choice of length-to-width ratio for such tanks.

Forward velocity, $V_h$ (m/h), is given by

$V_h = \frac{Q}{WH}$, where W is width of tank (m)

or,

$V_h = \frac{L}{t}$, where L is length of tank

Where L = length of tank (m)

t = detention time (hours)

This expression is represented in the upper left-hand quadrant of and, when read in conjunction with the lower left-hand quadrant, gives the relationship between $V_h$ and L:W ratio for rectangular tanks. Values of L/W in practice range from 3 to 6, with a value of 4 being common. For conversion, m/h x $\frac{1000}{3600}$ = mm/s.

*Weir loading rate*, $Q/L_w$, is important in *rectangular tanks*. A single weir across the end of these tanks is considered too short to prevent the influence of the approach current generated by the weir from extending upstream into the settling zone, with possible disruption of the flow pattern through the tank. The length of the weir can be doubled by placing a collection trough in the tank at the surface just before the end of the tank so that the water can flow into the trough from both sides. If

this is still insufficient for larger tanks, the weir length can be increased by providing multiple suspended weir troughs, designed to limit the maximum weir loading rate to about 12 $m^3/m.h$ (4 - 8 more typical). The troughs usually take the shape of square fingers, projecting into the tank for a short distance in the direction facing the oncoming water flow. Water can then flow into the troughs from all perimeters and the length of the trough is greatly extended by these so-called "fingers". The fingers all connect to an end trough or the final end weir, from where the water flows to the sand filters.

In *circular radial flow tanks,* the weir loading rate on a single perimeter weir, is usually within the normal range of values, so that suspended weirs are not necessary for small circular tanks. The water then runs over the weir into a collection trough all along the whole perimeter of the tank. From there, it would run into one or more pipes or channels to take the water to the sand filters. In larger radial tanks, the trough is placed within the tank, a little distance from the outer edge. It is placed at such a depth as to then make both sides of the trough overflow weirs into the trough. This almost doubles the length of the weir compared with a single, peripheral weir.

Precautions must be taken with outlet weirs in large tanks because the very small depth of the of flow over the weir under low flow conditions may be completely biased towards one end of the settling tank due to slight construction inaccuracies (1mm can be significant) or even wind pressures. These problems can be counteracted by having a sawtooth pattern on the weir to increase local depth, without affecting overall weir loading rates and possible scouring of sediments.

*Inlets* should be designed to dissipate the momentum and accurately distribute the incoming flow in such a way as to establish the required flow pattern in the tank.

The diverging flow which occurs in circular outward flow tanks is inherently less stable than the uniform forward flow in rectangular tanks, so that the design of the inlet stilling box is important in circular tanks. Excessive turbulence must be avoided in the inlet region, since the sludge collection hopper in most types of tank is located immediately beneath the inlet.

*Sludge scrapers* must be provided in modern rectangular and large circular sedimentation tanks, since it is not practicable to slope the floor sufficiently for gravitational self-cleaning.

One of the distinctive features of square hopper-bottomed tanks is that their sides are steeply sloped so that they are self-cleaning. Sludge moves down the walls by gravity to collect at the bottom of the hopper, from where it can be drawn off under hydrostatic head. The operating simplicity and lack of mechanical parts, which are features of these tanks have led to their widespread use in small treatment plants. They are not generally economical for larger plants, however, because of the

rapid increase in hopper depth and the corresponding cost increases which large tanks entail.

#### 4.1.3.1 HORIZONTAL FLOW SEDIMENTATION TANK

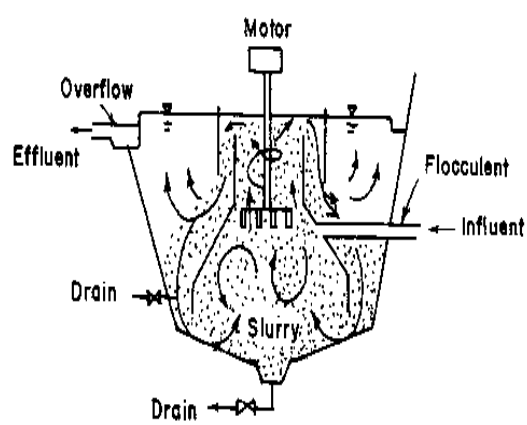
Some practical design data are provided for based on practical experiences. Various features must be incorporated into the design to obtain an efficient sedimentation process. The inlet to the tank must provide uniform distribution of flow across the tank. If more than one tank exists, the inlet must provide equal flow to each tank. Baffle walls are often placed at the inlet to distribute even flow, by use of 100-200 mm diameter holes evenly spaced across the width of the wall. Table 4.1 lists typical values.

**Table 4.1 Horizontal flow sedimentation tanks**

| Parameter | Design value |
|---|---|
| Surface loading rate ($m^3/m^2.d$) | 20-60 |
| Mean horizontal velocity (m/min) | 0.15-0.90 |
| Water depth (m) | 3-5 |
| Detention time (min) | 120-240 |
| Weir loading rate ($m^3$/m-d) | 100-200 |

#### 4.1.3.2 SLUDGE BLANKET CLARIFIERS

The SBC is quite flexible and can be adopted for use in almost all site conditions. Figure 2.9 shows an interesting SBC system with plate settler arrangement. Here the usual clariflocculator is also equipped with plate settler, thereby facilitating the solid removal. The plate settler sits on top of the clariflocculator. Therefore, in most situations, this unit does not need a subsequent filter unit.




(a) Slurry circulation type

**Sludge blanket clarifiers**

There are other forms of the SBC such as a combined slurry recirculation type with paddle flocculator or a SBC with plate

settlers. SBCs with flocculator are also sometimes called "clariflocculator" or other brand names. Another variation is where pulsed flow is used to induce the required velocity gradients in the clarifier to aid flocculation and such clarifiers are then called "pulsator" or "superpulsator" clarifiers.

The clariflocculator seems to have some clear advantages, even though it looks slightly sophisticated. It is the complicated theory that is sophisticated, but not the reactor itself. Some established detail designs of the SBC are available and could easily be incorporated into any new designs. There are modified versions incorporating a plate settler and filter to achieve entire solid-liquid separation in the same unit itself. Most of these modifications are made to suit the need of developing nations. Its use is highly recommended in developing countries - especially in small community water supply schemes - because of its flexibility in capacity and its ability to take up widely varying turbidity loads. It is highly suitable for package treatment plants, which are useful in remote areas as well as in congested urban areas.

**Example 4.1** Design a sedimentation tank for a flow (Q) of 1000 $m^3/d$. Determine the dimensions of the tank and the outflow weir length. Assume suitable design criteria.

**Solution:** Assume an overflow rate (OFR) of 20 $m^3/m^2.d$ as a typical value.

$$\text{Area} = \frac{Q}{OFR} = \frac{1000}{20} = 50\ m^2$$

Assuming a detention time (DT) of 2 h,

$$\text{Volume} = Q \times DT = 1000 \times \frac{2}{24} = 83.3\ m^3$$

$$\text{Depth} = \frac{V}{A} = \frac{83.3}{50} = 1.7\ m$$

If width (W) to length (L) ratio is 1:3, then

$A = 3W^2 = 50$

$W = 4.1$ m

$L = 3W = 12.3$ m

Assuming a weir loading rate (WLR) of 160 $m^3/m.d$,

$$\text{Minimum weir length} = \frac{Q}{WLR} = \frac{1000}{160} = 6.3m$$

In order to accommodate this required weir length, a double trough at the end of the tank would suffice, having a length of 8m.

**Example 4.2** Design a coagulation sedimentation tank with a continuous flow for treating water for a population of 45,000 persons with an average daily consumption of 135 L/person. Assume a surface loading rate of 0.9 $m^3m^{-2}h^{-1}$ and that the weir loading rate is within acceptable limits.

**Solution:**

Average consumption = 135 x 45,000 = 6,075,000 L/d.

Allow 1.8 times for maximum daily consumption:

maximum daily consumption = 1.8 x 6,075,000 = 10,935$m^3$/d.

Therefore, required surface area of the tank = (10,935/24)/0.9 = 506 $m^2$.

Assume minimum depth of tank = 3.5 m.

Therefore, (settling) volume of the tank = 506 x 3.5 = 1772 $m^3$.

Assume a length to width ratio of the tank of 3.5:1. Therefore the width would be

= 506/3.5$w^2$ m = 12m

Therefore, length of tank = 3.5 x 12 = 42 m.

Assuming a bottom slope of 1 in 60.

Depth of the deep end (at the influent end) = 3.5 + (1/60) x 42 = 4.2 m.

A floc chamber should be provided, at the entry to the tank, the capacity of which is assumed to be 1/16 of the settling chamber, i.e. = 1772/16 = 110.8 $m^3$.

If the depth of floc chamber is 2.5 m, then

the area of the floc chamber = 110.8/2.5 = 44.3 $m^2$.

The flocculation chamber also has a width equal to the sedimentation chamber, ie 12m. Therefore, length of floc chamber = 45.562/12 ≈ 3.8m.

It should be considered to add this length to the settling tank as it would otherwise reduce the settling volume by 3.8/42 = 9%. However, considering that we have already provided amply for maximum flow conditions, we could still fit the flocculation unit within the tank.

### 4.1.4 FACTORS AFFECTING SEDIMENTATION

Several factors affect the separation of settle able solids from water. Some of the more common types of factors to consider are:

#### 4.1.4.1 PARTICLESIZE

The size and type of particles to be removed have a significant effect on the operation of the sedimentation tank. Because of their density, sand or silt can be removed very easily. The velocity of the water-flow channel can be slowed to less than one foot per second, and most of the gravel and grit will be removed by simple gravitational forces. In contrast, colloidal material, small particles that stay in suspension and make the water seem cloudy, will not settle until the material is coagulated and flocculated by the addition of a chemical, such as an iron salt or aluminum sulfate.

The shape of the particle also affects its settling characteristics. A round particle, for example, will settle much more readily than a particle that has ragged or irregular edges.

All particles tend to have a slight electrical charge. Particles with the same charge tend to repel each other. This repelling

action keeps the particles from congregating into flocs and settling.

#### 4.1.4.2 WATER TEMPERATURE

Another factor to consider in the operation of a sedimentation basin is the temperature of the water being treated. When the temperature decreases, the rate of settling becomes slower. The result is that as the water cools, the detention time in the sedimentation tanks must increase. As the temperature decreases, the operator must make changes to the coagulant dosage to compensate for the decreased settling rate. In most cases temperature does not have a significant effect on treatment. A water treatment plant has the highest flow demand in the summer when the temperatures are the highest and the settling rates the best. When the water is colder, the flow in the plant is at its lowest and, in most cases; the detention time in the plant is increased so the floc has time to settle out in the sedimentation basins.

#### 4.1.4.3 CURRENTS

Several types of water currents may occur in the sedimentation basin:

- Density currents caused by the weight of the solids in the tank, the concentration of solids and temperature of the water in the tank.

- Eddy currents produced by the flow of the water coming into the tank and leaving the tank.

The currents can be beneficial in that they promote flocculation of the particles. However, water currents also tend to distribute the floc unevenly throughout the tank; as a result, it does not settle out at an even rate.

Some of the water current problems can be reduced by the proper design of the tank. Installation of baffles helps prevent currents from short circuiting the tank.

### 4.1.5 SEDIMENTATION BASIN ZONES

Under ideal conditions, the sedimentation tank would be filled with the water that has been coagulated, and the floc would be allowed to settle before any additional water is added. That is not possible for most types of water treatment plants.

Most sedimentation tanks are divided into these separate zones:

#### 4.1.5.1 INLET ZONE

The inlet or influent zone should provide a smooth transition from the flocculation zone and should distribute the flow uniformly across the inlet to the tank. The normal design includes baffles that gently spread the flow across the total inlet of the tank and prevent short circuiting in the tank. (Short circuiting is the term used for a situation in which part of the influent water exits the tank too quickly, sometimes by flowing across the top or along the bottom of the tank.) The baffle could include a wall across the inlet, perforated with holes across the width of the tank.

#### 4.1.5.2 SETTLING ZONE

The settling zone is the largest portion of the sedimentation basin. This zone provides the calm area necessary for the suspended particles to settle.

#### 4.1.5.3 SLUDGE ZONE

The sludge zone, located at the bottom of the tank, provides a storage area for the sludge before it is removed for additional treatment or disposal.

Basin inlets should be designed to minimize high flow velocities near the bottom of the tank. If high flow velocities are allowed to enter the sludge zone, the sludge could be swept up and out of the tank.

Sludge is removed for further treatment from the sludge zone by scraper or vacuum devices which move along the bottom.

#### 4.1.5.4 OUTLET ZONE

The basin outlet zone or launder should provide a smooth transition from the sedimentation zone to the outlet from the tank. This area of the tank also controls the depth of water in the basin. Weirs set at the end of the tank control the overflow rate and prevent the solids from rising to the weirs and leaving the tank before they settle out. The tank needs enough weir length to control the overflow rate, which should not exceed 20,000 gallons per day per foot of weir.

### 4.1.6 SELECTION OF BASIN

There are many sedimentation basin shapes. They can be rectangular, circular, and square

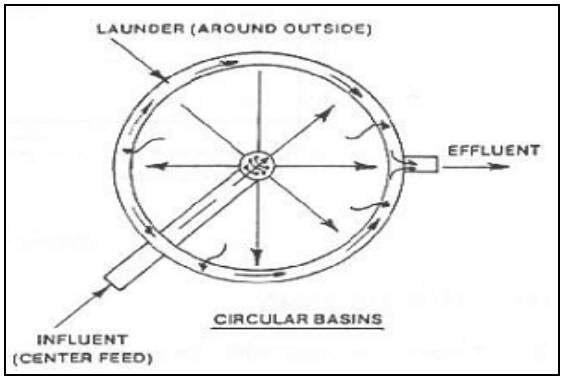
#### 4.1.6.1 RECTANGULAR BASINS

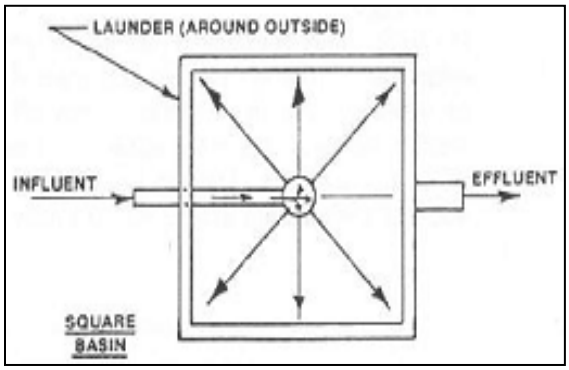
Rectangular basins are commonly found in large-scale water treatment plants. Rectangular tanks are popular as they tend to have:

- High tolerance to shock overload
- Predictable performance
- Cost effectiveness due to lower construction cost
- Lower maintenance
- Minimal short circuiting

#### 4.1.6.2 CIRCULAR AND SQUARE BASINS

Circular basins are frequently referred to as clarifiers. These basins share some of the performance advantages of the rectangular basins, but are generally more prone to short circuiting and particle removal problems. For square tanks the design engineer must be certain that some type of sludge removal equipment for the corners is installed.




### 4.1.7 SOLIDS CONTACT UNITS

A solids contact unit combines the coagulation, flocculation, and sedimentation basin in one unit. These units are also called up flow clarifiers or sludge-blanket clarifiers. The solids contact unit is used primarily in the lime-soda ash process to settle out the floc formed during water softening. Flow is usually in an upward direction through a sludge blanket or slurry of flocculated suspended solids.

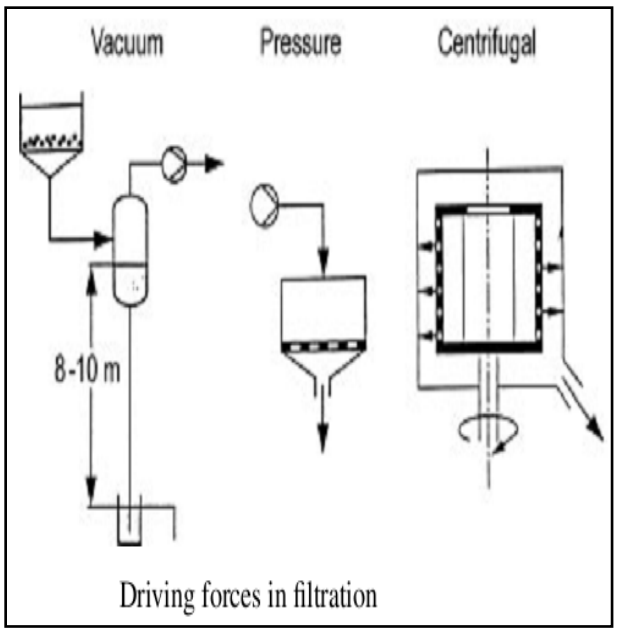
## 4.2 FILTRATION

Filtration is the separation of solid particles or liquid ones (droplets) from liquids and gases with the help of a filter medium also called a septum, which is essentially permeable to only the fluid phase of the mixture being separated. In earlier times, this process was carried out with felts, and the word “filter” has a common derivation with “felt”. Often however, purification of a liquid or gas is called filtration even when no semi-permeable medium is involved (as in electro-kinetic filtration).

The liquid more or less thoroughly separated from the solids is called the filtrate, effluent, permeate or, more rarely, clean water. As in other separation processes, the separation of phases is never complete: Liquid adheres to the separated solids (cake with residual moisture) and the filtrate often contains some solids (solids content in the filtrate or turbidity).

### 4.2.1 MECHANISM OF FILTRATION PROCESS

Filtration is effected by application of vacuum, pressure, or of centrifugal force (as shown in figure given below).Vacuum filtration requires a vacuum pump. The pump evacuates the gas from a filtrate receiver, where the filtrate is separated from the gas. The filtrate is drained either by a barometric leg of at least 8 to 10mor by a pump that is able to “run on snore” (i.e. with a deficiency of feed liquid so that it tends to draw in air). Pressure filtration typically only requires a pump for delivering of the suspension and the filter is placed within a pressure vessel, hence less easily accessible. Centrifugal filtration is done in perforated centrifuge rotors (→Centrifuges, Filtering).



Driving forces in filtration

Vacuum filters have the great advantage that the cake is freely accessible. This facilitates automatic cake handling. However, vacuum filters cannot handle hot liquids, or solvents with high vapor pressure. The pressure difference across vacuum filters is very limited, and the residual moisture of the filter cake is higher than with pressure filters. Pressure filters in turn are preferred when the product must be kept in a closed system for safety reasons, or if the residual moisture content is important. The handling of the filter cake is obviously more difficult in a pressure filter. Filtration by centrifugal force requires more technical equipment, but as a general rule it yields solids with lower residual moisture (→Centrifuges, Filtering).

### 4.2.2 FILTRATION MODELS

Various models can describe the physical process of filtration. There are four idealized filtration models as depicted in given figure.

#### 4.2.2.1 CAKE FILTRATION

Cake filtration is the most frequently used model. Here it is assumed that the solids are deposited on a filter medium as a homogeneous porous layer with a constant permeability. Thus, if the flow rate dV/dt is constant, the pressure drop will increase lin-early, proportional to the quantity of solid deposited. This model can be applied particularly well for all hard, particulate solids.

#### 4.2.2.2 BLOCKING FILTRATION

The pressure drop is caused by solid particles blocking pores. Soft, gelatinous particles retained by a sieve exhibit such a behavior. If the flow rate dV/dt is constant, the pressure drop increases exponentially with the quantity filtered, the number of open pores asymptotically approaching zero. The pores may belong to a filter medium (screen or filter layer) or it may be pores within a filter cake of coarse particles, which are blocked by migrating fine particles.

#### 4.2.2.3 DEEP BED OR DEPTH FILTRATION

Solid particles are retained in a deep filter layer. This takes place for example in sand filters for clarification of drinking water, which retain even colloidal particles. The typical effect of deep bed filtration is adhesion of solids to the grains of the filter layer, comparable to char-coal adsorption. Only rather big particles are retained by

the screening effect. When the filter bed has been saturated with solids, the solids concentration in the filtrate leaving the bed progressively approaches that of the in-coming suspension.

#### 4.2.2.4 CROSS-FLOW FILTRATION

In cross-flow filtration the suspension flows with high speed tangentially to the filter surface, preventing the formation of a cake. Only a small flow of liquid passes through the filter medium. A certain layer of solids accumulates in the boundary layer on the filter surface, and reduces the flow of filtrate. After an initial period, a dynamic equilibrium is established between convective transport of solids to the filter surface and removal of solids by turbulence and by diffusion.

- **SURFACE FILTRATION**

Surface filtration is the antonym to depth filtration. The solids are retained on the surface of a filter medium. Generally the models of cake filtration or of blocking filtration can be applied.

- **SCREENING**

Screening designates a classification process, which retains the particles below a certain size and let's passes the smaller ones (→Screening). Often the term screening filtration is also used to designate a surface filtration with a screen as a filter medium. Its mode of action resembles screening (or straining) as long as the filter medium is clean, but it is clearly a cake filtration as soon as a layer of solids has formed.

### 4.2.3 TYPES OF FILTERS

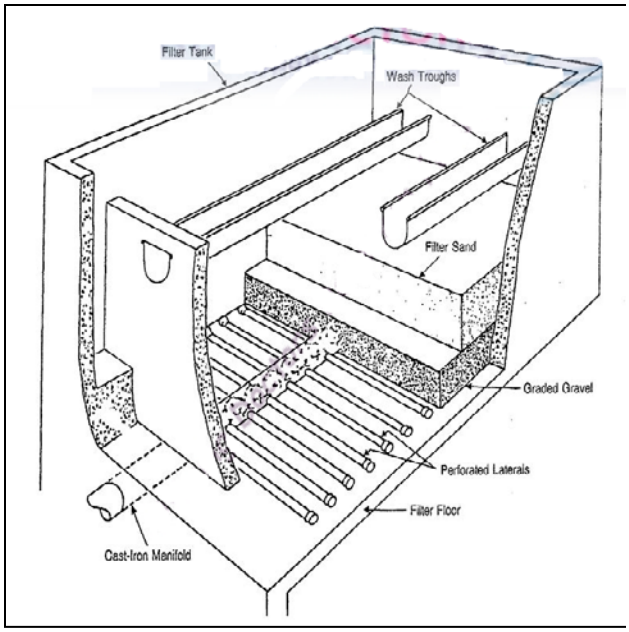
Several types of filters are used for water treatment. The earliest ones developed were the slow sand filters. They typically have filter rates of around 0.05 gpm/ft$^2$ of surface area. This type of filter requires large filter areas. The top several inches of the sand has to be removed regularly--usually by hand--due to the mass of growing material ("schmutzdecke") that collects in the filter. The sand removed is usually washed and returned to the filter. They may also be used as a final step in wastewater treatment.

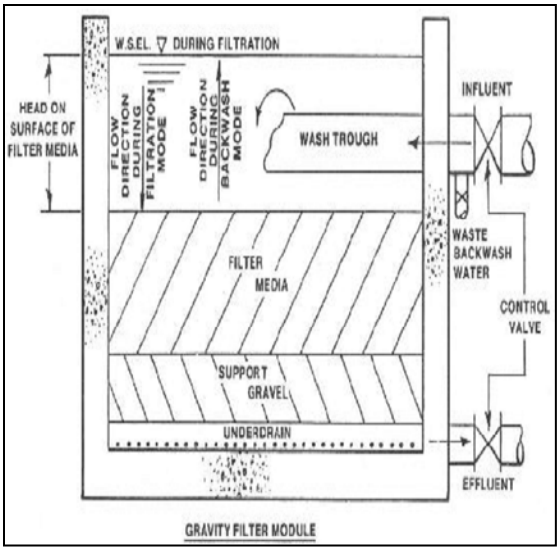
Most filters are classified by filtration rate, type of filter media, or type of operation into:

- Gravity Filters
- Rapid Sand Filters
    - High Rate Filters
        - Dual media
        - Multi-media
- Pressure Filters
    - Sand or Multi-media

#### 4.2.3.1 RAPID SAND FILTERS

Rapid sand filters can accommodate filter rates 40 times those of slow sand filters. The major parts of a rapid sand filter are:



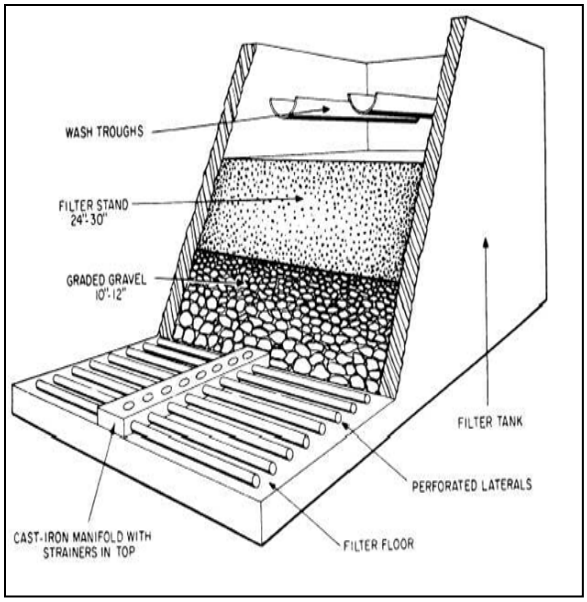




Filter tank or filter box Filter sand or mixed-media Gravel support bed under drain system Wash water troughs Filter bed agitators.

The filter tank is generally constructed of concrete and is most often rectangular. Filters in large plants are usually constructed next to each other in a row, allowing the piping from the sedimentation basins to feed the filters from a central pipe gallery. Some smaller plants are designed with the filters forming a square of four filters with a central pipe gallery feeding the filters from a center well.

### 4.2.3.2 SAND FILTERS

The filter sand used in rapids and filters is manufactured specifically for the purpose of water filtration. Most rapid sand filters contain 24-30 inches of sand, but some newer filters are deeper. The sand used is generally 0.4 to 0.6 mm in diameter. This is larger than the sand used in slow rate filtration. The coarser sand in the rapid filters has larger voids that do not fill as easily.




### 4.2.3.3 WASH WATER TROUGHS

Wash water troughs placed above the filter media collect the backwash water and carry it to the drain system. Proper placement of these troughs is very important to ensure that the filter media is not carried into the troughs during the backwash and removed from the filter. The wash troughs must be installed at the same elevation so that they remove the backwash evenly from the filter and so that an even head is maintained across the entire filter. These backwash troughs are constructed from concrete, plastic, fiberglass, or other corrosion-resistant materials.

- **SURFACE WASH**

During the operation of a filter, the upper six-to-ten inches of the filter media remove most of the suspended material from the water. It is important that this layer be thoroughly cleaned during the backwash cycle. Normal backwashing does not, in most cases, clean this layer completely; therefore, some method of agitation is needed to break up the top layers of the filter and to help the backwash water remove any material caught there.

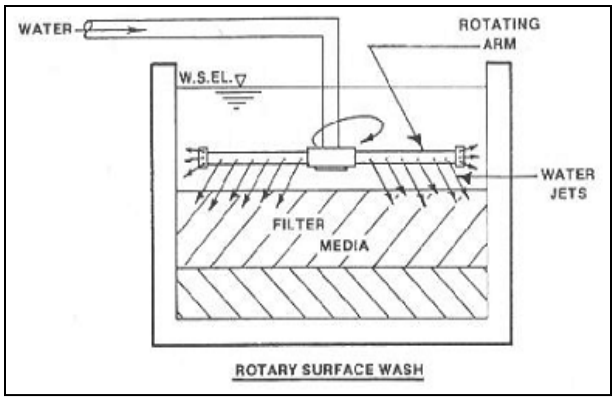
The surface wash system consists of a series of pipes installed in the filter that introduce high velocity water or air jet action into the upper layer of the filter. This jet action will generally be supplied by rotating arms that are activated during the backwashing of the filter.

A newer design of surface wash uses compressed air to mix the upper layer and loosen the particles from the sand so that the backwash water can remove the particles more easily. This air wash generally is turned on before the backwash cycle. If both are used at the same time, some sand may be washed away. The compressed air rate can be two-to-five cubic feet per minute per square foot (cfm / $ft^2$) of filter surface, depending on the design of the filter.

### 4.2.3.4 HIGH RATE FILTERS

High rate filters, which operate at a rate three-to-four times that of rapid sand filters, use a combination of different filter media, not just sand. The combinations vary with the application, but generally they are sand and anthracite coal. Multi-media or mixed-media filters use three or four different materials, generally sand, anthracite coal, and garnet.

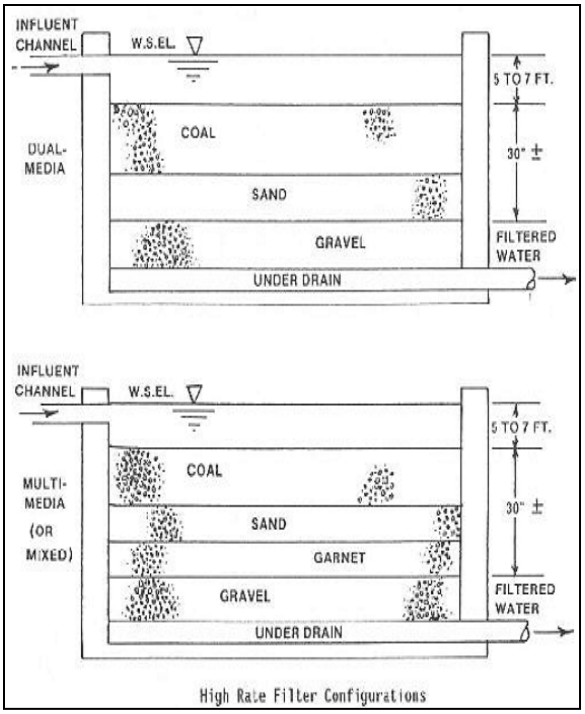
In rapid sand filters, finer sand grains are at the top of the sand layer with larger grains farther down into the filter. As a result, the filter removes more suspended material in the first few inches of the filter. In the high rate filter, the media size



ROTARY SURFACE WASH

decreases. The top layers consist of a coarse material with the finer material farther down, allowing the suspended material to penetrate deeper into the filter.

The material in a filter bed forms layers in the filter, depending on their weight and specific gravities. In the coarse layer at the top, the larger suspended particles are removed first, followed by the finer materials. This allows for longer filter runs at higher rates than is possible with rapid sand filters.

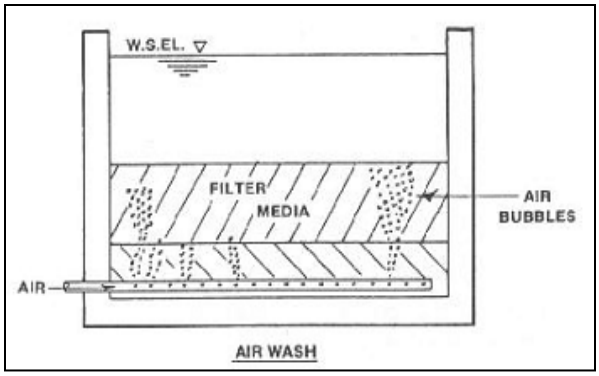


High Rate Filter Configurations

The type of filter media used in a high rate filter depends on many factors, including the raw-water quality, raw-water variations, and the chemical treatment used. Pilot studies help the operator evaluate which material, or combination of materials, will give the best result.

### 4.2.3.5 PRESSURE FILTERS

Pressure filters fall into two categories: pressure sand and diatomite filters.



AIR WASH

- **PRESSURE SAND FILTERS**

This type of filter is used extensively in iron and manganese removal plants.

A pressure sand filter is contained under pressure in a steel tank, which may be vertical or horizontal, depending on the space available. As with gravity filters, the media is usually sand or a combination of media. Filtration rates are similar to gravity filters.

These filters are commonly used for iron and manganese removal from groundwater, which is first aerated to oxidize the iron or manganese present, then pumped through the filter to remove the suspended material.

Because the water is under pressure, air binding will not occur in the filter. However, pressure filters have a major disadvantage in that the backwash cannot be observed; in addition, cracking of the filter bed can occur quite easily, allowing the iron and manganese particles to go straight through the filter. When using pressure filters for iron and manganese removal, the operator must regularly measure the iron and manganese concentration of the filter effluent and backwash the filter before breakthrough occurs. Because of these limitations, pressure filters must not be used to treat surface water.

- **DIATOMACEOUS EARTH FILTER**

This type of filter is commonly used for the treatment of swimming pools. The process was developed by the military during World War II to remove microorganisms that cause amoebic dysentery from water used in the field.

### 4.2.4 CALCULATION FOR PRESSURE DROP

#### 4.2.4.1 CAKE FILTRATION

The resistance to flow of a filter cake can be de-scribed by Darcy's law (as shown in figure given below). Consider a liquid flowing through a filter cake (or a stream of water percolating through soil as considered by Darcy). The pressure drop Δp of this flow is proportional to:

1) The flow rate per unit area V/A
2) The cake thickness H
3) The viscosity η of the liquid
4) A constant $\alpha_H$ describing the "specific filter resistance" of the cake:

$$\boxed{\Delta p_1 = \left(\frac{\dot{V}}{A}\right) \cdot H \cdot \eta \cdot \alpha_{\mathrm{H}}}$$

The SI units are:

$$\left[\mathrm{Pa} = \left(\frac{\mathrm{m}^3/\mathrm{s}}{\mathrm{m}^2}\right) \cdot \mathrm{m} \cdot \mathrm{Pa} \cdot \mathrm{s} \cdot \mathrm{m}^{-2}\right]$$

The unit of $\alpha_H$ must therefore be $m^{-2}$ in order to satisfy the Darcy equation. The reciprocal of the filter resistance $\alpha_H$ is also called permeability *k* of the filter cake:

$$\boxed{k = \frac{1}{\alpha_H} \left[\mathrm{m}^2\right]}$$

Sometimes it is more convenient to define cake thickness in terms of solid mass per unit area (unit kg/m$^2$). This leads to a slightly different definition of filter resistance, the factors being:

1) The flow rate per unit area V/A

2) The cake thickness m/A

3) The viscosity η of the liquid

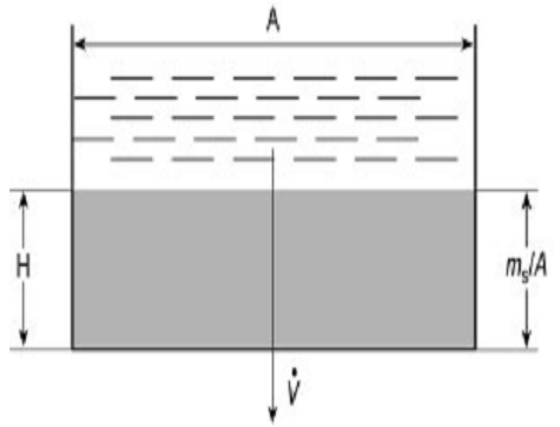
4) A constant $\alpha_m$ with the unit m/kg describing the resistance of the cake.

Then the following expression is obtained

$$\Delta p_1 = \left(\frac{\dot{V}}{A}\right) \cdot \left(\frac{m}{A}\right) \cdot \eta \cdot \alpha_m$$

For practical reasons the viscosity η is very often not measured separately. Then it is legitimate to include it in a term $\alpha_H\eta$ (unit mPa·s/m$^2$) or $\alpha_m\eta$ (unit mPa·s·m/kg), respectively.

Using this latter term $\alpha_H\eta$ or $\alpha_m\eta$ filter resistances lie between $10^{11}$ mPa·s/m$^2$ (filtering very rapidly) and $10^{16}$ mPa·s/m$^2$ (nearly un-filterable), or between $10^8$ and $10^{13}$ mPa·s·m/kg, respectively.




Definition of filter resistance
$\eta\alpha_H$ = Cake thickness H related filter resistance;
$\eta\alpha_m$ = Solid mass related filter resistance

### 4.2.4.2 THE CAKE FILTER EQUATION

The pressure drop in a filter is composed of a pressure drop$\Delta$p1across the cake according to Equation and a pressure drop $\Delta$p2 across the filter medium, which can be written as:

$$\Delta p_2 = \beta \cdot \eta \cdot \left(\frac{\dot{V}}{A}\right)$$

Where b (unit m$^{-1}$) is the resistance of the filter medium.

The total pressure drop is therefore:

$$\Delta p = \Delta p_1 + \Delta p_2 = \alpha_H \eta \cdot H \cdot \frac{\dot{V}}{A} + \beta\eta\frac{\dot{V}}{A}$$

Or

$$\Delta p = \Delta p_1 + \Delta p_2 = \alpha_m \eta \cdot \frac{m \cdot \dot{V}}{A^2} + \beta\eta\frac{\dot{V}}{A}$$

If the suspension is homogeneously mixed, the cake height H (or m / A) will be

proportional to the quantity of filtrate. The concentration is de-scribed by a factor K:

$$K_{\mathrm{H}} = \frac{H \cdot A}{V}$$

Or

$$K_{\mathrm{m}} = \frac{m}{V}$$

This gives

$$\Delta p = \frac{\alpha_{\mathrm{H}} \eta K_{\mathrm{H}}}{A^2} \cdot V \cdot \frac{\mathrm{d}V}{\mathrm{d}t} + \frac{\beta \eta}{A} + \frac{\mathrm{d}V}{\mathrm{d}t}$$

Or

$$\boxed{\Delta p = \frac{\alpha_{\mathrm{m}} \eta K_{\mathrm{m}}}{A^2} \cdot V \cdot \frac{\mathrm{d}V}{\mathrm{d}t} + \frac{\beta \eta}{A} + \frac{\mathrm{d}V}{\mathrm{d}t}}$$

The above equations are identical under the generalization $\alpha_H \cdot K_H = \alpha_m \cdot K_m$. Nevertheless the distinction between both equations is useful for clarity.

The differential Equations can be integrated either for constant flow rate or for constant pressure. Integration for constant flow rate dV / dt =const gives the trivial solution:

$$\boxed{\Delta p = \frac{\eta \cdot \dot{V}}{A} \cdot \left( \frac{\alpha_{\mathrm{H}} K_{\mathrm{H}}}{A} \cdot V + \beta \right)}$$

Or

$$\Delta p = \frac{\eta \cdot \dot{V}}{A} \cdot \left( \frac{\alpha_{\mathrm{m}} K_{\mathrm{m}}}{A} \cdot V + \beta \right)$$

For Δp = constant, the integration yields:

$$\mathrm{d}t = \frac{\alpha \eta K}{A^2 \cdot \Delta p} \cdot V \cdot \mathrm{d}V + \frac{\beta \eta}{A \cdot \Delta p} \cdot \mathrm{d}V$$

$$t = \frac{\alpha_{\mathrm{H}} \eta K_{\mathrm{H}}}{2 A^2 \cdot \Delta p} \cdot V^2 + \frac{\beta \eta}{A \cdot \Delta p} \cdot V$$

Or

$$\boxed{t = \frac{\alpha_{\mathrm{m}} \eta K_{\mathrm{m}}}{2 A^2 \cdot \Delta p} \cdot V^2 + \frac{\beta \eta}{A \cdot \Delta p} \cdot V}$$

#### 4.2.4.3 COMPRESSIBLE CAKE FILTRATION

Most filter cakes are compressible, which means that their resistance increases with growing pressure. An x-fold increase in filtration pressure therefore normally gives rise to a less than x-fold increase in flow rate. The compression of the cake is caused by the compressive stress $p_s$ on the solids which is caused by the drag of the flowing liquid (see below figure). The loss in liquid pressure $p_L$ translates into solids pressure $p_S$, the sum of both being constant:

$$\partial p_S = -\partial p_L$$

$$p_{\mathrm{S}} + p_{\mathrm{L}} = p_0$$

The local filter resistance $\alpha_{loc}$ in the cake is a function of the local compressive stress $p_S$ and it is therefore low at the surface of the cake and high near the interface with the filter medium. This $\alpha_{loc}$ =f($p_S$) can be measured in the Compression

Permeability cell (CP cell), where a sample of the cake is subject to a known compressive stress.

However, only the average resistance $\alpha_{av}$ of a filter cake, all layers combined, is important for practical scale-up purposes. It can be calculated from $\alpha_{loc}$ measured in the CP-cell.

$$\alpha_{av} = \frac{\Delta p}{\int_0^p \frac{dp}{\alpha_{loc}}}$$

Normally however it is much easier to measure $\alpha_{av}$ directly with filtration experiments in a pressure filter. The uneven distribution of local porosity and filter resistance is then ignored.

The dependence of filter resistance $\alpha_{av}$ or $\alpha_{loc}$ on pressure can be approximated over a limited pressure range by

$$\alpha = \alpha_0 \cdot \left( \frac{p_S}{\text{unit pressure}} \right)^n$$

From the differential form of filtration equation, we know that

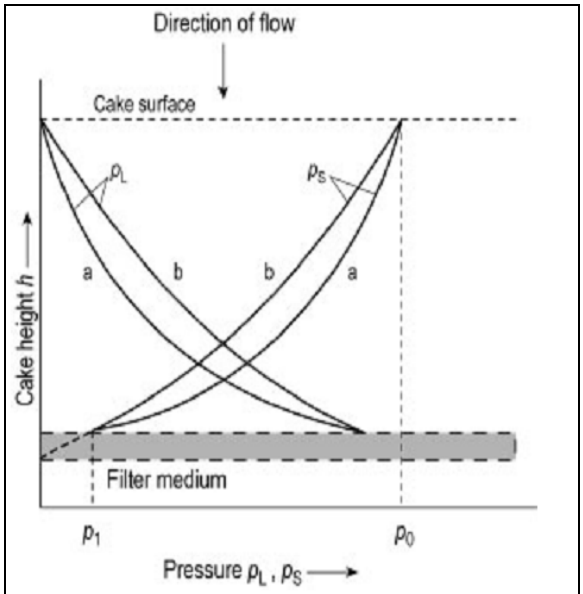
$$\alpha_m \eta = \Delta \left( \Delta p \cdot \frac{dt}{dV} \right)_e \cdot \frac{A^2}{m_e}$$

Where $\alpha_0$ is the resistance at unit pressure drop and n is a compressibility index (equal to zero for incompressible cakes). Both parameters can be determined easily from a logarithmic plot $\alpha = f(p_S)$, where the slope indicates the compressibility factor n. Introducing $\alpha_{loc}$ according to this approximation into Equation , it can be shown that for 0 < n < 1.

$$n_{av} = n_{loc}$$

$$\alpha_{0,av} = (1 - n) \cdot \alpha_{0,loc}$$

Thus the approximation Equation has the advantage that it applies to both the local and the average resistance. However, such an approximation is valid only for a limited range of pressures and Equations (18) and (19) are restricted to 0 < n < 1.




Hydraulic pressure $p_L$ and solids pressure $p_S$ as function of cake height. Curve (a) represents a more compressible cake than curve (b)

### 4.3 GAS SOLID SEPARATION

Gas – Solid separation is an unit operation used to control air pollution and to prevent loss of materials. Gas – solid separation can be achieved by the application of the principles of electrostatic effects, gravitational settling, filtration, wet scrubbing or centrifugation.

The removal of dust from gas, termed *gas cleaning,* is often necessary for a number of reasons:

(i) To control air pollution
(ii) To prevent dust from entering a machine
(iii) To prevent wastage of valuable materials
(iv) To reduce the maintenance of equipments

### 4.3.1 FACTORS AFFECTING THE GAS SOLID OPERATION

During gas- solid separation, the factors which affect this operation are:

(i) The properties of solid particles
(ii) The quality of dust to be handled
(iii) The moisture content of gas and dust
(iv) The temperature of the gas – solid system

### 4.3.2 MECHANISMS OF GAS SOLID OPERATION

The principal separation mechanisms which are involved in gas – solid separation are

(i) Gravitational settling
(ii) Inertial separation
(iii) Washing with a liquid (Scrubbing)
(iv) Electrostatic deposition
(v) Centrifugal separation

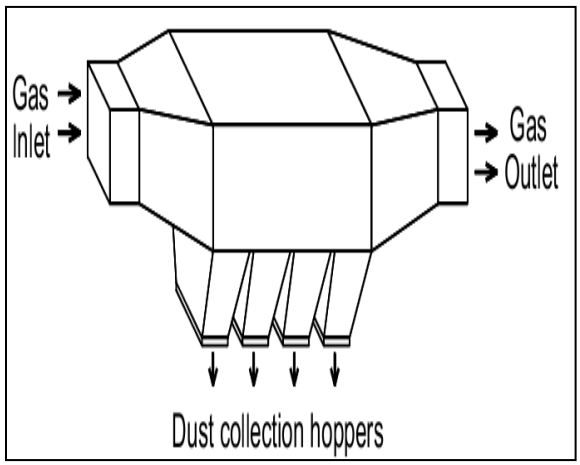
### 4.3.3 GAS CLEANING EQUIPMENTS

Gas cleaning equipments, also known as *gas collectors,* vary widely in design, operation, effectiveness, space requirement, construction and capital and operating and up – keep costs. The selection of a dust collector depends on the following factors:

(i) Dust concentration and particle size
(ii) Air / gas stream characteristics
(iii) Dust characteristics
(iv) Degree of dust removal desired
(v) Method of dust disposal

#### 4.3.3.1 GRAVITY SETTLING CHAMBER

This is the simplest of all separation equipment. You have a big box, with the inlet and outlet streams way up at the top, and as the fluid flows through, the particles fall out. If the box is long enough, then all the particles should fall out. This brings us back to the settling velocity; if the flow rate through is too fast for the size of the box you are using, then not all the particles, if any, are going to settle.



Despite their simple design and economical mode of operation, these units re seldom used because of their large space requirements and low efficiency.

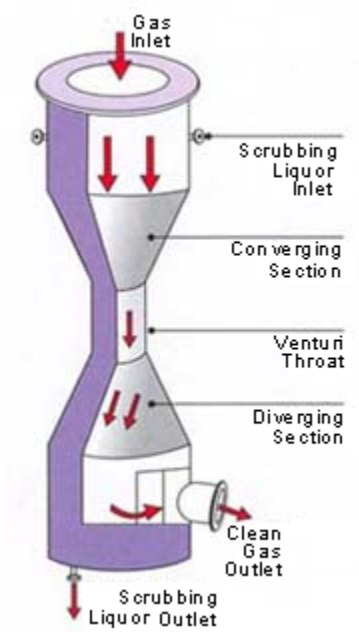
#### 4.3.3.2 WET SCRUBBERS

Wet scrubbers, commonly known as *wet collectors,* are a class of gas cleaning devices in which a scrubbing liquid (usually water) is used for the separation of dust particles from the gas stream. The separation efficiency largely depends on the degree of contact between the gas and the liquid streams. There are very large variety of wet scrubbers available and one such design is the *venturi scrubber,* which is also the most widely used type of wet scrubber.

Venturi scrubbers are high – energy type of wet scrubbers and work on the principle that the energy from the inlet gas stream is being utilized to atomize the scrubbing liquid.

Venturi scrubbers consist of a venturi-shaped inlet and separator. The dust-laden gases venturi scrubbers enter through the venturi and are accelerated to speeds between 12,000 and 36,000 ft/min (60.97-182.83 m/s). These high-gas velocities immediately atomize the coarse water spray, which is injected radially into the venturi throat, into fine droplets. High energy and extreme turbulence promote collision between water droplets and dust particulates in the throat. The agglomeration process between particle and droplet continues in the diverging section of the venturi. The large agglomerates formed in the venturi are then removed by an inertial separator.

Venturi scrubbers achieve very high collection efficiencies for respirable dust. Since efficiency of a venturi scrubber depends on pressure drop, some manufacturers supply a variable-throat venturi to maintain pressure drop with varying gas flows.




#### 4.3.3.3 CYCLONE SEPARATORS

Cyclonic separation is a method of removing particulates from an air, gas or liquid stream, without the use of filters, through vortex separation. Rotational effects and gravity are used to separate mixtures of solids and fluids. The method can also be used to separate fine droplets of liquid from a gaseous stream.

A high speed rotating (air) flow is established within a cylindrical or conical container called a cyclone. Air flows in a helical pattern, beginning at the top (wide end) of the cyclone and ending at the

bottom (narrow) end before exiting the cyclone in a straight stream through the center of the cyclone and out the top. Larger (denser) particles in the rotating stream have too much inertia to follow the tight curve of the stream, and strike the outside wall, then falling to the bottom of the cyclone where they can be removed. In a conical system, as the rotating flow moves towards the narrow end of the cyclone, the rotational radius of the stream is reduced, thus separating smaller and smaller particles. The cyclone geometry, together with flow rate, defines the *cut point* of the cyclone. This is the size of particle that will be removed from the stream with 50% efficiency. Particles larger than the cut point will be removed with a greater efficiency and smaller particles with a lower efficiency.

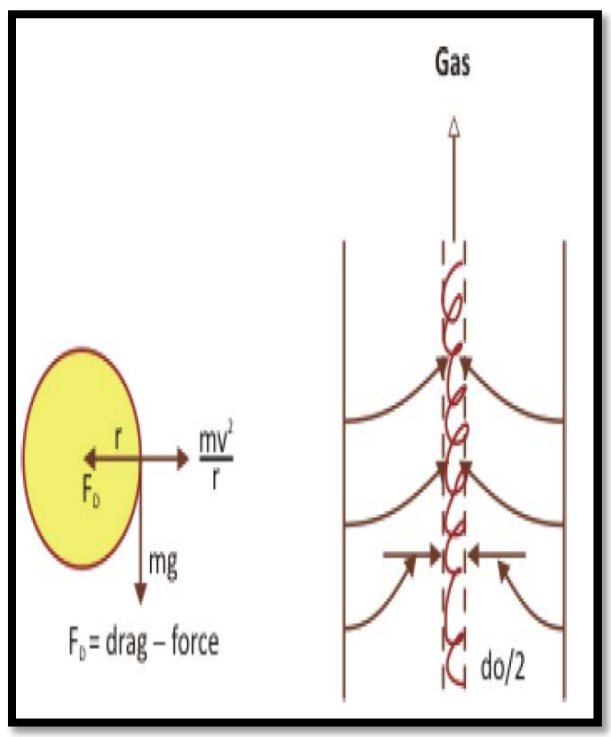
a. The real trajectory of gas and particles is difficult to analyze. The particles laden gas enters the cyclone from the sideway (see top view) at a high flow rate and moves downward in a swirling/ spiral path.

b. Solid particles are thrown outward radially due to centrifugal force. They strike the walls of cyclone and settle down. Gas, on the other hand, will move radially inward, then upward through the least hydrodynamically resistance – path to the exit.

c. Gas moving in spiral reaches the apex of the cone, then moves upward in a smaller spiral (~ $d_o$ / 2 ) path to the exit at the top, as the opening at the bottom is filled with solid particles. For the gas, the least resistance – path is the exit at the top. For the particles, the least resistance- path is the exit at the bottom.

d. Mechanistically, if the centrifugal force acting on the particles is larger than the drag (inward) by the gas, the particles will strike the walls and settle down, else they will move inward along with the gas. At a radius r, where these two forces are equal, particle will rotate in equilibrium and move downward till they hit the slant walls and are collected. Gas on the other hand has a very high upward flow rate at the center, typically in the core-diameter of $d_o$ / 2.

Any particle in the zone will be carried upward.




Theoretical 'cut-size' of a cyclone is the particle size above which all particles will

be collected. A theoretical expression considering drag and centrifugal forces on a particle, has been obtained to estimate the 'cut size' of cyclone. The calculation takes into account the experimental observation that the equilibrium rotation-radius of all captured particles in cyclone is $\geq \frac{1}{2}\left(\frac{do}{2}\right)$ or 0.25do, where do is the diameter of the nozzle at the top of the cyclone though which the gas exits.

- The settling velocity of captured particles,

$$v_o = \frac{0.25 \times do \times G \times g}{\pi \times Z \times D_c \times \rho_f \times U_{to}^2}$$

Where

$d_o$ = diameter of gas exit at nozzle, m

G = gas flow rate, $m^3$ / s

G = gravity, $m^2$ / s

Z = height of cyclone, m

$D_c$ = cyclone diameter, m

$\rho_f$ = gas density, kg / $m^3$

$U_{to}$ = G / $\rho A_i$, m / s

$A_i$ = cross sectional area of inlet pipe, $m^2$

Form $v_o$, the theoretical cut–diameter, $d_p$ is determined from the settling velocity equation:

$$v_o = \frac{d_p^2 g\left(\rho_g - \rho_f\right)}{18\mu_f}$$

Let assumed that particles settle in Stoke's regime.

- All particles having diameter < $d_p$ will have equilibrium radius within 0.5 do so that they will be carried away with the gas.

All particles having diameter > $d_p$ will be captured in cyclone.

- Cyclones are very effective in removing particles from gas. Disadvantages are large flow rate required and large pressure–drop because of the tortuous path of the gas.
- $\Delta p \simeq 8\rho_f \frac{v_c^2}{2}$, where $v_c$ = gas velocity at the inlet
- Separation factor of a cyclone, s is defined as

$$S = \frac{F_c}{F_g} = \frac{mv^2/r}{mg} = \frac{v^2}{rg}$$

- Cyclones are effective typically for particle size > $5\ \mu m$
- Efficiency (capturing) of cyclone, $\eta_c$

$$\eta_c = \frac{\text{mass of collected particles of diameter, } d_p}{\text{mass of particles of diameter } d_p \text{ in the incoming gas}}$$

Design graphs are available to calculate the efficiency.

# GATE QUESTIONS

**Q.1** Match the systems in Group I with equipment used to separate them in Group II

| Group I | Group II |
|---|---|
| P gas – solid | 1 filter press |
| Q liquid – liquid | 2 cyclone |
| | 3 decanter |
| | 4 thickener |

(a) P-1, Q-2 (b) P-2, Q-3
(c) P-3, Q-4 (d) P-4, Q-1

**(GATE 2004)**

**Q.2** For a cyclone of diameter 0.2 m with a tangential velocity of 15 m/s at the wall, the separation factor is
(a) 2250 (b) 1125
(c) 460 (d) 230

**(GATE 2004)**

**Q.3** For a particle settling in water at its terminal settling velocity, which of the following is true?
(a) Buoyancy =weight + drag
(b) Weight =buoyancy + drag
(c) Drag = buoyancy + weight
(d) Drag = weight

**(GATE 2004)**

**Q.4** In constant pressure filtration,
(a) Resistance decreases with time
(b)Rate of filtration is constant
(c) Rate of filtration increases with time
(d) Rate of filtration decreases with time

**(GATE 2004)**

**Q.5** A centrifugal filtration unit operating at a rotational speed of ω has inner surface of the liquid (density $\rho_L$) located at a radial distance R from the axis of rotation. The thickness of the liquid film is δ and no cake is formed. The initial pressure drop during filtration is
(a) $\frac{1}{2}\omega^2 R^2\rho_L$ (b) $\frac{1}{2}\omega^2\delta\rho_L(2R + \delta)$
(c) $\frac{1}{2}\omega^2\delta^2\rho_L$ (d) $\frac{1}{2}\omega^2 R\rho_L(R + 2\delta)$

**(GATE 2004)**

**Q.6** $U_{mf}$ is the minimum fluidization velocity for a bed of particles. An increase in the superficial gas velocity from, $2U_{mf}$ to 2.5Umf results in (all velocities are smaller than the entrainment velocity of the particles) no change in
(a) Drag on particles
(b) Drag on column walls
(c) The bed height
(d) The bed voidage

**(GATE 2004)**

**Q.7** The Kozney-Carman equation, rewritten in terms of non-dimensional numbers, gives $\left(\frac{\Delta P}{\rho u^2}\right)$ proportional to

(a) $\left(\frac{L/D_p}{Re}\right)$ (b) $\left(\frac{Re}{L/D_p}\right)$

(c) $\left(\frac{L/D_p}{Re^2}\right)$ (d) $\left(\frac{Re^2}{L/D_p}\right)$

**(GATE 2004)**

**Q.8** A centrifuge of diameter 0.2 m in a pilot plant rotates at a speed of 50 Hz in order to achieve effective separation. If this centrifuge is scaled up to a diameter of 1 m in the chemical plant, and the same separation factor is to be achieved, what is the rotational speed of the scaled up centrifuge?

(a) 15 Hz (b) 22.36 Hz

(c) 30 Hz (d) 44.72 Hz

**(GATE 2005)**

**Q.9** What is the terminal velocity in m/s, calculated from Stokes law, for a particle of diameter $0.1 \times 10^{-3}$ m, density 2800 kg/m$^3$ settling in water of density 1000 kg/m$^3$ and viscosity $10^{-3}$ kg/m-s)? (Let g = 10 m/s$^2$)

(a) $2 \times 10^{-2}$ (b) $4 \times 10^{-3}$

(c) $10^{-2}$ (d) $8 \times 10^{-3}$

**(GATE 2005)**

**Q.10** A bed fluidized by water is used for cleaning sand contaminated with salt. The particles of sand and salt have the same shape and size but different densities ($\rho_{sand}$ = 2500 kg/m$^3$and $\rho_{salt}$ = 2000 kg/m$^3$). If the initial volume fraction of the salt in the mixture is 0.3 and if the initial value of the minimum fluidization velocity ($U_{mf}$) is 0.9 m/s, find the final value of the $U_{mf}$ (in m/s) when the sand is washed free of the salt. Assume that the bed characteristics (bed porosity and solid surface area per unit volume) do not change during the operation and that the pressure drop per unit length is directly proportional to the fluid velocity

(a) 0.70 (b) 0.90

(c) 1.00 (d) 1.46

**(GATE 2006)**

**Q.11** Two spherical particles have the same outer diameter but are made of different materials. The first one (with material density $\rho_1$) is solid, whereas the second (with material density $\rho_2$) is a hollow sphere with the inner shell diameter equal to half the outer diameter. If both the spheres have the same terminal velocity in any fluid, then the ratio of their material densities, $\rho_2/\rho_1$, is

(a) 1 (b) 8/7

(c) 2 (d) 8

**(GATE 2006)**

**Q.12** A filtration is conducted at constant pressure to recover solids from dilute slurry. To reduce the time of filtration, the solids concentration in the feed slurry is increased by evaporating half the solvent. If the resistance of the filter medium is negligible, the filtration time will be reduced by a factor of

(a) 1 (b) 2

(c) 4 (d) 8

**(GATE 2006)**

**Q.13** In constant pressure filtration, the rate of filtration follows the relation

($V$: filtrate volume, t: time, $k$ and c: constants).

(a) $\frac{dv}{dt} = kv + C$ (b) $\frac{dv}{dt} = \frac{1}{kv + C}$

(c) $\frac{dv}{dt} = kv$ (d) $\frac{dv}{dt} = kv^2$

**(GATE 2007)**

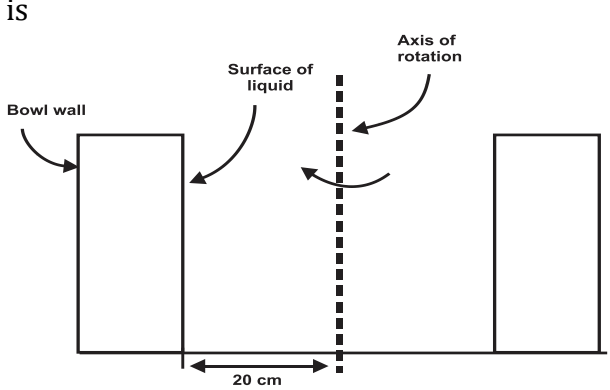
**Q.14** In the Stokes regime, the terminal velocity of particles for centrifugal sedimentation is given by

$U_t = \omega^2 r (\rho_p - \rho) d^2_p / 18 \mu$

where, $\omega$: angular velocity; $r$ : distance of the particle from the axis of rotation;

$\rho_p$: density of the particle; $\rho$: density of the fluid; $d_p$ : diameter of the particle and $\mu$ : viscosity of the fluid.

In a Bowl centrifugal classifier operating at 60 rpm with water ($\mu$ = 0.001 kg/m.s), the time taken for a particle ($d_p$ = 0.0001 m, sp.gr = 2.5) in seconds to traverse a distance of 0.05 m from the liquid surface is




(a) 4.8 (b) 5.8

(c) 6.8 (d) 7.8

**(GATE 2007)**

**Linked Answer Questions 15 & 16**

A fluidized bed (0.5m dia, 0.5m high) of spherical particles (diameter = 2000 μm, specific gravity = 2.5) uses water as the medium. The porosity of the bed is 0.4. The Ergun eqn for the system is

$\Delta P/L = 4 \times 10^5 U_{mf} + 1 \times 10^7 U_{mf}^2$

(Sl unit, $U_{mf}$ in m/s).

**Q.15** ΔP/L (SI unit) at minimum fluidization condition is (g = 9.8 m/sec$^2$).

(a) 900 (b) 8820

(c) 12400 (d) 17640

**(GATE 2007)**

**Q.16** The minimum fluidization velocity (mm/sec) is

(a) 12.8 (b) 15.8

(c) 24.8 (d) 28.8

**(GATE 2007)**

**Q.17** For laminar flow conditions, the relationship between the pressure drop ($\Delta P_c$) across an incompressible filter cake and the specific surface area ($S_o$) of the particles being filtered in given by ONE of the following:

(a) $\Delta P_c$ is proportional to $S_o$

(b) $\Delta P_c$ is proportional to $1/S_o$

(c) $\Delta P_c$ is proportional to $S_o^2$

(d) $\Delta P_c$ is proportional to $1/S_o^2$

**(GATE 2008)**

**Q.18** Two identically sized spherical particles A and B having densities $\rho_A$ and $\rho_B$, respectively; are settling in a fluid of density ρ. Assuming free settling under turbulent flow conditions, the ratio of the terminal settling velocity of particle A to that of particle B is given by

(a) $\sqrt{\frac{(\rho_A - \rho)}{(\rho_B - \rho)}}$ (b) $\sqrt{\frac{(\rho_B - \rho)}{(\rho_A - \rho)}}$

(c) $\frac{(\rho_A - \rho)}{(\rho_B - \rho)}$ (d) $\frac{(\rho_B - \rho)}{(\rho_A - \rho)}$

**(GATE 2008)**

**Q.19** The terminal settling velocity of a 6 mm diameter glass sphere (density: 2500 kg/m$^3$) in a viscous Newtonian liquid (density: 1500 kg/m$^3$) is 100 μm/s. If the particle Reynolds number is small and the value of acceleration due to gravity is 9.81 m/s$^2$, then the viscosity of the liquid (in Pa. s) is

(a) 100 (b) 196.2

(c) 245.3 (d) 490.5

**(GATE 2009)**

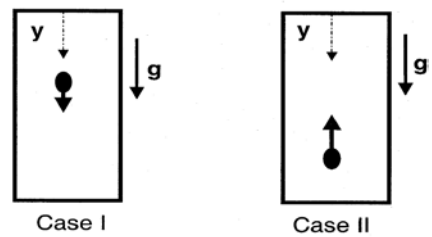
**Q.20** The height of a fluidized bed at incipient fluidization is .075 m and corresponding voltage of 0.38. If the voltage of the bed increases to 0.5, then the height of the bed would be:-

(a) .058 m (b) .061 m

(c) .075 m (d) .093 m

**(GATE 2010)**

**Q.21** Consider the following two cases of movement of particles. In Case I, the particle moves along the positive y-direction and in Case II, the particle moves along negative y-direction. Gravity acts along the positive y-direction. Which ONE of the following options corresponds to the CORRECT directions of buoyancy acting on the particles?




(a) Positive y – direction for both the cases,

(b) Negative y – direction for Case I, positive y – direction for Case II

(c) Negative y – direction for both the cases

(d) Positive y – direction for Case I, negative y – direction for Case II

**(GATE 2011)**

**Common data question 22 – 23.**

For liquid flowing through a packed bed the pressure drop per unit length of then bed $\frac{\Delta p}{L}$ is

$$\frac{\Delta P}{L} = \frac{150\mu V_0}{(\Phi_s d_p)^2}\frac{(1-\varepsilon)^2}{\varepsilon^3} + \frac{1.75\rho V_0^2(1-\varepsilon)}{\Phi_s d_p \varepsilon^3}$$

Where $V_0$ is the superficial liquid velocity, $\varepsilon$ is the bed porosity, $d_p$ is average particle size, $\Phi_s$ is particle sphericity, $\rho$ is liquid density and $\mu$ is liquid viscosity.

Given data : $d_p = 1\times10^{-3}$ m, $\Phi_s = 0.8$, $\rho = 1000$ kg/m3, $\mu = 1\times10^{-3}$ kg/ms

particle density $\rho$ = 2500 kg/m and acceleration due to gravity g = 9.8 m/s$^2$.

**Q.22** When velocity is 0.005 m/s and $\varepsilon = 0.5$ which ONE of the following is the CORRECT value for the ratio of the viscous loss to the kinetic energy loss?

(a) 0.09 (b) 1.07

(c) 10.71 (d) 93

**(GATE 2011)**

**Q.23** On further increasing, incipient fluidization is achieved. Assuming that the porosity of the bed remains unaltered, the pressure drop per unit length (in pa/m) under incipient fluidization condition is

(a) 3675 (b) 7350

(c) 14700 (d) 73501

**(GATE 2011)**

**Q.24** Taking the acceleration; due to gravity to be 10 m/s$^2$, the separation factor of a cyclone 0.5 m in diameter and having a tangential velocity of 20 m/s near the wall is____

**(GATE 2013)**

**Q.25** A cylindrical packed bed of height 1 m is filled with equal sized spherical particles. The particles are nonporous and have a density of 1500 kg/m$^3$ . The void fraction of the bed is 0.45. The bed is fluidized using air (density 1kg/m$^3$ ). If the acceleration due to gravity is 9.8m/s$^2$ , the pressure drop (in pa) across the bed at incipient fluidization (up to one decimal place) is _____

**(GATE 2015)**

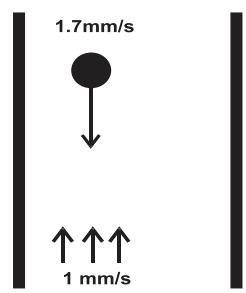
**Q.26** A typical batch filtration cycle consists of filtration followed by washing, one such filtration unit operating at constant pressure difference first filters a slurry during which 5 liters of filtrate is collected in 100 s. This is followed by washing. Which is done for $t_w$ seconds and uses 1 liter of wash water. Assume the following relation to be applicable between the applied pressure drop ΔP, cake thickness L at time t, and volume of liquid V collected in time t.

$$\frac{\Delta P}{L} = k_1 \frac{dV}{dt}; \qquad L = k_2 V \text{, if L is changing.}$$

$k_1$ and $k_2$ can be taken to be constant during filtration and washing. The wash time tw, in seconds (up to one decimal place) is _____

**(GATE 2015)**

**Q.27** A spherical solid particle of 1mm diameter is falling with a downward velocity of 1.7mm/s through a liquid (viscosity 0.04 Pa·s) at a low Reynolds number (Strokes regime). The liquid is flowing upward at a velocity of 1 mm/s. All velocities are with respect to a stationary reference frame. Neglecting the wall effects, the drag force per unit projected area of the particle, in Pa, (up to two decimal places) is _____




**(GATE 2015)**

**Q.28** In a cyclone separator used for separation of solid particles from a dust laden gas, the separation factor is defined as the ratio of the centrifugal force to the gravitational force acting on the particle. Sr denotes the separation factor at a location (near the wall) that is at a radial distance r from the centre of the cyclone. Which one of the following statements is INCORRECT?

(a) Sr depends on mass of the particle
(b) Sr depends on the acceleration due to gravity
(c) Sr depends on tangential velocity of the particle
(d) Sr depends on the radial location (r) of the particle

**(GATE 2016)**

**Q.29** Consider a rigid solid sphere falling with a constant velocity in a fluid. The following data are known at the conditions of interest: viscosity of the fluid = 0.1 Pa s, acceleration due to gravity = 10 m $s^{-2}$, density of the particle = 1180 kg $m^{-3}$ and density of the fluid = 1000 kg $m^{-3}$. The diameter (in mm, rounded off to the second decimal place) of the largest sphere that settles in the Stokes' law regime (Reynolds number 0.1), is _____

**(GATE 2016)**

**Q.30** A gas bubble (gas density $\rho_g = 2\ kg/m^3$; bubble diameter D = $10^{-4}$ m) is rising vertically through water (density $\rho = 1000\ kg/m^3$; viscosity $\mu = 0.001$ Pa.s).

Force balance on the bubble leads to the following equation

$$\frac{dv}{dt} = -g\frac{\rho_s - \rho}{\rho_g} - \frac{18\mu}{\rho_g D^2}v$$

Where v is the velocity of the bubble at any given time t. assume that the volume of the rising bubble does not change. The value of g = 9.81 m/$s^2$

The terminal rising velocity of the bubble (in cm/s), rounded to 2 decimal places, is ___ cm/s

**(GATE 2017)**

**Q.31** The terminal velocity of a spherical particle in gravitational settling under Stokes regime varies
(a) linearly with the particle diameter
(b) linearly with the viscosity of the liquid
(c) directly with the square of particle diameter
(d) inversely with the density of particle

**(GATE 2018)**

| ANSWER KEY | | | | | | | |
|---|---|---|---|---|---|---|---|
| **1** | **2** | **3** | **4** | **5** | **6** | **7** | **8** |
| (b) | (d) | (b) | (d) | (b) | (c) | (a) | (b) |
| **9** | **10** | **11** | **12** | **13** | **14** | **15** | **16** |
| (c) | (c) | (b) | (c) | (b) | (c) | (b) | (b) |
| **17** | **18** | **19** | **20** | **21** | **22** | **23** | **24** |
| (c) | (a) | (b) | (d) | (c) | (c) | (b) | 160 |
| **25** | **26** | **27** | **28** | **29** | **30** | **31** | |
| 8079.61 | 40 | 1.296 | (a) | 2.15 | 0.54 | (d) | |

# EXPLANATIONS

**Q.1 (b)**
For gas- solid separation cyclone is used while for liquid liquid operation, decanter is used.

**Q.2 (d)**
We know that

Separation factor, $S = \frac{V_0^2}{gR}$

Where, $V_0$ tangential velocity =15m/s

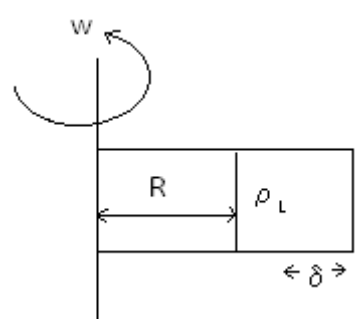
R= radius = 0.1m

$$S = \frac{(15)^2}{9.81 \times 0.1} = 229.59 \approx 230$$

**Q.3 (b)**
At terminal settling velocity
Weight =buoyancy + drag

**Q.4 (d)**

**Q.5 (b)**




$$\Delta P = \frac{1}{2}\rho_L \omega^2 \left[ (R+\delta)^2 - R^2 \right]$$

$$\Rightarrow = \frac{1}{2}\rho_L \omega^2 \left[ R^2 + \delta^2 + 2R\delta - R^2 \right]$$

$$\Rightarrow \Delta P = \frac{1}{2}\rho_L \omega^2 \left[ \delta^2 + 2R\delta \right]$$

$$\Rightarrow \Delta P = \frac{1}{2}\rho_L \omega^2 \delta (\delta + 2R)$$

**Q.6 (c)**
As the superficial gas velocity increases, the pressure drop increases but the

particles do not move and the bed height remains same.

**Q.7 (a)**

Kozeny- Carman equation is

$$\frac{\Delta P}{L} = \frac{150\mu\,\bar{u}(1-\epsilon)^2}{\phi_s^2 D_P^2\,\epsilon^3}$$

Now, $$\frac{\Delta p}{\rho u^2} = \frac{150\,\mu\bar{u}L}{\phi_S^2 D_P^2 \rho \bar{u}^2}\frac{(1-\epsilon)^2}{\epsilon^3}$$

$$\frac{150(10\,\epsilon)^2}{\phi_s^2\,\epsilon^3}\frac{(L/\Delta P}{\left(\frac{\delta D_P\,\bar{u}}{\mu}\right)} = k.\frac{L/D\rho}{Re}$$

$$\Rightarrow \quad \frac{\Delta p}{\delta\mu^2} \propto \frac{(L/D_P)}{Re}$$

**Q.8 (b)**

To maintain same separation factor

$\omega_1^2 r_1 / g = \omega_2^2 r_2 / g$

On Putting values, we get:

$\omega_2 = 22.36$ Hz

**Q.9 (c)**

From Stoke's law,

$$u = \frac{gD_P^2(\rho_l - \rho)}{18\mu}$$

$$u = \frac{10\times(0.1\times10^{-3})^2(2800-1000}{18\times10^{-3}}$$

$$u = 10^{-2}\,m/s$$

**Q.10 (c)**

We know that

Initially, sand: salt = 0.7: 0.3

Given, $\rho_{sand} = 2500\ kg/m^3$

$\rho_{salt} = 2000\,kg/m^3$

Applying force balance,

pressure force = Weight of the bed – Buoyancy Force

$$\Delta P\times A = \left[\rho_{sand}\cdot AL(1-\varepsilon)\cdot 0.7 + \rho_{salt}\cdot AL(1-\varepsilon)\cdot 0.3\right]g - \rho_{water}\cdot AL(1-\varepsilon)\cdot g$$

$$\Rightarrow \Delta P = \left[2500\cdot L(1-\varepsilon)\cdot 0.7 + 2000\cdot L(1-\varepsilon)\cdot 0.3\right]g - 1000\cdot L(1-\varepsilon)\cdot g$$

$$\Rightarrow \frac{\Delta P}{L} = 1350(1-\varepsilon)\cdot g \qquad \text{(i)}$$

At final condition, all salt is washed away.

Applying force balance,

pressure force = Weight of the bed – Buoyancy Force

$$\Delta P'\times A = \left[\rho_{sand}\cdot AL(1-\varepsilon)\right]g - \rho_{water}\cdot AL(1-\varepsilon)\cdot g$$

$$\Rightarrow \Delta P' = \left[2500\cdot L(1-\varepsilon)\right]g - 1000\cdot L(1-\varepsilon)\cdot g$$

$$\Rightarrow \frac{\Delta P'}{L} = 1500(1-\varepsilon)\cdot g \qquad \text{(ii)}$$

From equation (i) and (ii),

$$\Rightarrow \quad \frac{\Delta P}{\Delta P'} = \frac{1350}{1500} = 0.9$$

Given, $\Delta P \propto U$,

thus, $$\frac{\Delta P}{\Delta P'} = \frac{U}{U'} = 0.9$$

$$\Rightarrow \frac{0.9\,m/s}{U'} = 0.9 \Rightarrow U' = 1\,m/s$$

**Q.11 (b)**

We know, $$u_t = \sqrt{\frac{2g(\rho_p - \rho)m}{A_P\rho_p\times C_D\times\rho}}$$

Therefore, $u_t \propto \sqrt{m}$

For solid sphere, let diameter = D

$$\Rightarrow \text{Volume of Sphere} = \frac{4}{3}\times\pi\left(\frac{D}{2}\right)^3 = \frac{\pi}{6}\times D^3$$

$$\Rightarrow \text{Mass of Sphere} = \frac{\pi}{6}\times D^3\times\rho_1 \quad \left[\because \rho_1 = \text{density}\right]$$

For Hollow sphere

internal diameter, $D_i = \frac{1}{2}D_0 = \frac{1}{2}D \quad \left[\because D_0 = D\right]$

$$\Rightarrow \text{Volume of Hollow sphere} = \frac{4}{3}\pi\left[D_0^3 - D_i^3\right]$$

$$= \frac{4}{3}\pi\left[D^3 - \frac{D^3}{8}\right] = \frac{7\pi}{6}\times D^3$$

$\Rightarrow$ Mass of Sphere $= \frac{7\pi}{6} D^3 \times \rho_2$

Since terminal velocity is same, i.e., $U_{t_1} = U_{t_2}$

$$\Rightarrow \frac{U_{t_1}}{U_{t_2}} = \frac{\sqrt{m_1}}{\sqrt{m_2}}$$

$$\Rightarrow \sqrt{m_1} = = \sqrt{m_2}$$

$$\Rightarrow m_1 = m_2$$

$$\Rightarrow \frac{\pi}{6} \times D^3 \times \rho_1 = \frac{7\pi}{6} D^3 \times \rho_2$$

$$\Rightarrow \frac{\rho_1}{\rho_2} = \frac{7}{8} \text{ or } \frac{\rho_2}{p_1} = \frac{8}{7}$$

**Q.12 (c)**

For constant filtration, $\frac{t}{V} = KV + c$

Since, filter medium resistance in negligible, $c = 0$

$\Rightarrow t = KV^2 \Rightarrow t \propto V^2$

$$\Rightarrow \frac{t_2}{t_1} = \left(\frac{V_2}{V_1}\right)^2 = \left(\frac{0.5V_1}{V_1}\right)^2 = \frac{1}{4}$$

$$\therefore \quad t_2 = \frac{t_1}{4}$$

**Q.13 (b)**

**Q.14 (c)**

Given, $N = 60\text{rpm} = 1\text{rps}$

$\omega = 2\pi N = 2\pi \times 1 = 2\pi$ rad / s,

$\rho_p = 2500$ kg / m$^3$, $\rho = 1000$ kg / m$^3$,

$d_p = 0.0001$ m, $\mu = 0.001$ Pa – s

we have,

$$U_t = \frac{\omega^2 r(\rho_p - \rho) d_p^2}{18\mu}$$

$$\Rightarrow \frac{dr}{dt} = \frac{\omega^2 r(\rho_p - \rho) d_p^2}{18\mu}$$

$$\Rightarrow dt = \frac{18\mu}{\omega^2(\rho_p - \rho) d_p^2} \frac{dr}{r}$$

$$\Rightarrow \int_0^t dt = \frac{18\mu}{\omega^2(\rho_p - \rho) d_p^2} \int_{0.2}^{0.25} \frac{dr}{r}$$

$\left\{\begin{array}{l} \because \text{ particle is at r = 0.20 m at t = 0} \\ \text{and let time taken by particle to traverse} \\ \text{0.05 m is t} \end{array}\right\}$

$$\Rightarrow \int_0^t dt = \frac{18 \times 0.001}{(2\pi)^2 (2500 - 1000)(0.0001)^2} \int_{0.2}^{0.25} \frac{dr}{r}$$

$t = 6.78$ sec

**Q.15 (b)**

from force balance,

pressure force = Weight of the bed – Buoyancy Force

$$\Delta P \times A = \left[\rho_{solid} \cdot AL(1-\varepsilon)\right] g - \rho_{water} \cdot AL(1-\varepsilon) \cdot g$$

$$\Rightarrow \frac{\Delta P}{L} = (\rho_{solid} - \rho_{water})(1-\varepsilon) \cdot g$$

$$\Rightarrow \frac{\Delta P}{L} = (2500 - 1000)(1 - 0.4) \times 9.8 = 8820 \text{ Pa}$$

**Q.16 (b)**

The minimum fluidization velocity:

$$\frac{\Delta P}{L} = 4 \times 10^5 U_{mf} + 1 \times 10^7 U_{mf}^2$$

Putting values,

$$8820 = 4 \times 10^5 U_{mf} + 1 \times 10^7 U_{mf}^2$$

On solving, we get:

$\Rightarrow U_{mf} = 15.8$ mm / s

**Q.17 (c)**

**Q.18 (a)**

For turbulent flow region terminal settling vel.

$$= u = \sqrt{\frac{4 d_p (\rho_p - \rho) g}{3 C_D \rho}}$$

Diameter of both practicable is same

$$\frac{u_1}{u_2} = \sqrt{\frac{(\rho_A - \rho)}{(\rho_B - \rho)}}$$

**Q.19 (b)**

From Stoke's law:

$$U_t = \frac{g\ D^2(\rho_p - \rho)}{18\mu}$$

putting values

$$\therefore\ 100 \times 10^{-6} = \frac{9.81 \times (6 \times 10^{-3})^2 (1800 - 1000)}{18 \times \mu}$$

$$\mu = 196.2\ \text{Pa} \cdot \text{s}$$

**Q.20 (d)**

$$L2 = L_1 = \left[\frac{1-\varepsilon_1}{1-\varepsilon_2}\right]$$

$$L_2 = .075\left(\frac{1-.38}{1-.5}\right) = .093\text{m}$$

**Q.21 (c)**

**Q.22 (c)**

$$\frac{\Delta P}{L} = \frac{150\mu V_0 (1-\varepsilon)^2}{(\Phi_s d_p)^2}\frac{(1-\varepsilon)^2}{\varepsilon^3} + \frac{1.75\rho V_0^2 (1-\varepsilon)}{\Phi_s d_p \varepsilon^3}$$

$$\frac{\Delta P}{L} = \text{viscous lodd} + \text{kinetic energy loss}$$

$$\frac{\text{viscous loss}}{\text{kinetic energy loss}} = \frac{150\mu V_0}{(\Phi_s d_p)}\frac{(1-\varepsilon)^2}{\varepsilon^2}$$

$$\frac{150\mu(1-\varepsilon)}{1.75\rho V_0 \Phi_s d_p}$$

$$= \frac{150 \times 1 \times 10^{-3} \times (1-05)}{1.75 \times 1000 \times 0.005 \times 0.8 \times 1 \times 10^{-3}}$$

$$= 10.71$$

**Q.23 (b)**

At incipient fluidization

$$\frac{\Delta p}{L} = g\ (1-\varepsilon)\ (\rho_p - \rho_f)$$

$$= 9.8 \times (1 - 0.5) \times 2500 - 1000) = 7350$$

**Q.24 160**

The separation factor is given as

$$S = \frac{v^2}{g \times r}$$

Where, $v = 20\ m/s, g = 10\ m/s^2$ and $r = 0.25\ m$

Thus,

$$S = \frac{(20)^2}{10 \times 0.25}$$

$$\boxed{S = 160}$$

**Q.25 8079.61**

We know that,

For incipient fluidization,

$$\frac{\Delta P}{L} = g(1-\varepsilon)(\rho_p - \rho)$$

putting given values,

$$\Delta P = 9.8 \times (1 - 0.45)(1500 - 1) \times 1$$

$$\Delta P = 8079.61\ \text{Pa}$$

**Q.26 40**

We have

Given: $\frac{\Delta P}{L} = k_1 \frac{dV}{dt}$ and $L = k_2 V$

so, $\frac{dV}{dt} = \frac{1}{k_1}\frac{\Delta P}{k_2 V}$

$$\Rightarrow \int_0^5 V dV = \frac{\Delta P}{k_1 k_2}\int_0^{100} dt$$

$$\Rightarrow \frac{\Delta P}{k_1 k_2} = 0.125$$

also, final rate of filtration

$$\frac{dV}{dt} = \frac{\Delta P}{k_1 k_2 V} = \frac{0.125}{V} = \frac{0.125}{5} = 0.025$$

washing time,

$$t_w = \frac{\text{volume of wash water used}}{\text{rate of washing } (= \text{final rate of filtration})}$$

$$\Rightarrow \quad t_w = \frac{1}{0.025} = 40\ \text{sec}$$

**Q.27 1.296**

The drag force is given by

$$F_D = C_D \times A_P \times \frac{\rho_f v^2}{2}$$

For Laminar regime,

$$C_D = \frac{24}{N_{Re}} = \frac{24\mu_f}{d_p \rho_f v}$$

Thus

$$F_D = \frac{24\mu_f}{d_p \rho_f v} \times A_P \times \frac{\rho_f v^2}{2}$$

$$\Rightarrow \frac{F_D}{A_P} = \frac{12\mu_f v}{d_p} \quad \{\text{where, } v = \text{ velocity of particle in fluid}\}$$

$$\Rightarrow \frac{F_D}{A_P} = \frac{12 \times 0.04 \times (1.7 - (-1)) \times 10^{-3}}{1 \times 10^{-3}} = 1.296 \frac{N}{m^2}$$

The correct range of drag force per unit area
(in Pa) is given between: 1.25 – 1.35.

**Q.28 (a)**

Sr depends on mass of the particle, is the correct answer.

**Q.29 2.15**

We know that, for $(N_{Re} \le 0.1)$

Terminal setting velocity under stokes law regime is given by ,

$$v_t = \frac{d_P^2 \, g \, (\rho_p - \rho_f)}{18 \, \mu_f}$$

$$v_t = \frac{d_p^2 \text{ x}10 \text{ x } (1180 - 1000)}{18 \text{ x } 0.1}$$

$$v_t = 1000 \, d_p^2$$

In Stoke's law max value of particle Reynolds number is 0.1 and it is given by

$$\Rightarrow N_{Rep} = \frac{d_p \times v \times \rho_f}{\mu_f} = 0.1$$

Putting values,

$$\Rightarrow \frac{d_p \text{ x } (1000 \, d_p^2) \text{ x } 10^3}{0.1} = 0.1$$

$$d_p^3 = \frac{1}{10^8} m^3 x \left(\frac{10mm}{1m}\right)^3 = 10 \, mm$$

$$d_p = 2.15 \, mm$$

**Q.30 0.54**

At terminal missing velocity

$$\frac{dv}{dt} = 0 = -g\frac{\rho_g - \rho}{\rho_g} - \frac{18\mu}{\rho_g D^2} v$$

We get

$$v = \frac{gD^2(\rho - \rho_g)}{18\mu} = \frac{9.81 \times (10^{-4})^2 (1000 - 2)}{18(0.001)}$$

$$= 5.439 \times 10^{-3} \text{ m/s} = 0.54 \text{ cm/s}$$

**Q.31 (d)**

# 5 TRANSPORTATION OF SOLIDS

## 5.1 TRANSPORTATION EQUIPMENTS

The handling of solids and their transportation is also a very important part of any chemical industry. For the solids transportation within the industry, there are many methods and techniques but some of the useful and important techniques are as follows:

## 5.2 CONVEYORS

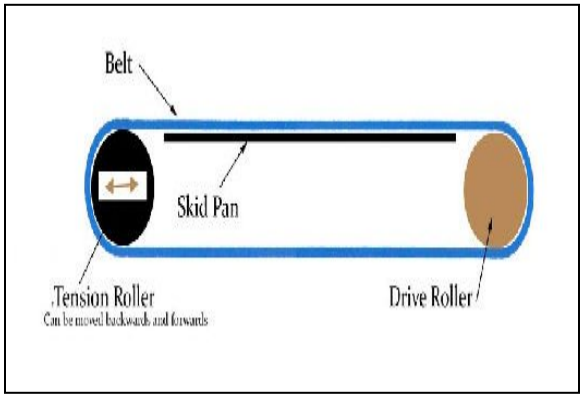
A conveyor system is a common piece of mechanical handling equipment that moves materials from one location to another. Conveyors are especially useful in applications involving the transportation of heavy or bulky materials. Conveyor systems allow quick and efficient transportation for a wide variety of materials, which make them very popular in the material handling and packaging industries. Many kinds of conveying systems are available, and are used according to the various needs of different industries. There are chain conveyors, belt conveyors, screw conveyors and roller conveyors.

### 5.2.1 BELT CONVEYORS

A conveyor belt (or belt conveyor) consists of two or more pulleys, with a continuous loop of material - the conveyor belt - that rotates about them. One or both of the pulleys are powered, moving the belt and the material on the belt forward. The powered pulley is called the drive pulley while the unpowered pulley is called the idler. There are two main industrial classes of belt conveyors; those in general material handling such as those moving boxes along inside a factory and bulk material handling such as those used to transport industrial and agricultural materials, such as grain, coal, ores, etc. generally in outdoor locations. Generally companies providing general material handling type belt conveyors do not provide the conveyors for bulk material handling. In addition there are a number of commercial applications of belt conveyors such as those in grocery stores.

The belt consists of one or more layers of material. They can be made out of rubber. Many belts in general material handling have two layers. An under layer of material to provide linear strength and shape called a carcass and an over layer called the cover. The carcass is often a woven fabric having a warp & weft. The most common carcass materials are polyester, nylon and cotton. The cover is often various rubber or plastic compounds specified by use of the belt. Covers can be made from more exotic materials for unusual applications

such as silicone for heat or gum rubber when traction is essential.




Belt conveyors are the most commonly used powered conveyors because they are the most versatile and the least expensive. Product is conveyed directly on the belt so both regular and irregular shaped objects, large or small, light and heavy, can be transported successfully. These conveyors should use only the highest quality premium belting products, which reduces belt stretch and results in less maintenance for tension adjustments. Belt conveyors can be used to transport product in a straight line or through changes in elevation or direction. In certain applications they can also be used for static accumulation or cartons.

### 5.2.2 CHAIN CONVEYOR

A chain conveyor is a type of conveyor system for moving material through production lines.

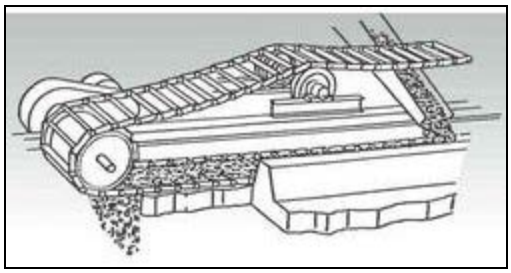
Chain conveyors utilize a powered continuous chain arrangement, carrying a series of single pendants. The chain arrangement is driven by a motor, and the material suspended on the pendants is conveyed. Chain conveyors are used for moving products down an assembly and / or around a manufacturing or warehousing facility.

Chain conveyors are primarily used to transport heavy unit loads, e.g. pallets, grid boxes, and industrial containers. These conveyors can be single or double chain strand in configuration. The load is positioned on the chains; the friction pulls the load forward. Chain conveyors are generally easy to install and have very minimum maintenance for users.

Many industry sectors use chain conveyor technology in their production lines. The automotive industry commonly use chain conveyor systems to convey car parts through paint plants. Chain conveyors also have widespread use in the white and brown goods, metal finishing and distribution industries. Chain conveyors are also used in the painting and coating industry, this allows for easier paint application. The products are attached to an above head chain conveyor, keeping products off of the floor allows for higher productivity levels.

### 5.2.3 SCREW CONVEYOR

A screw conveyor or auger conveyor is a mechanism that uses a rotating helical screw blade, called a "*flighting*",

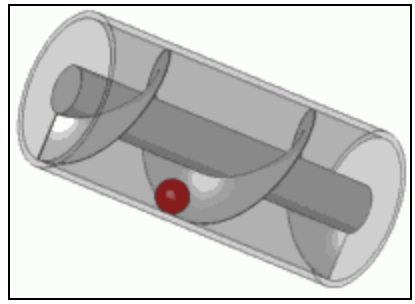


usually within a tube, to move liquid or granular materials. They are used in many bulk handling industries. Screw conveyors in modern industry are often used horizontally or at a slight incline as an efficient way to move semi-solid materials, including food waste, wood chips, aggregates, cereal grains, animal feed, boiler ash, meat and bone meal, municipal solid waste, and many others. The first type of screw conveyor was the Archimedes' screw, used since ancient times to pump irrigation water.

They usually consist of a trough or tube containing either a spiral blade coiled around a shaft, driven at one end and held at the other, or a "*shaft less spiral*", driven at one end and free at the other. The rate of volume transfer is proportional to the rotation rate of the shaft. In industrial control applications the device is often used as a variable rate feeder by varying the rotation rate of the shaft to deliver a measured rate or quantity of material into a process.

Screw conveyors can be operated with the flow of material inclined upward. When space allows, this is a very economical method of elevating and conveying. As the angle of inclination increases, the capacity of a given unit rapidly decreases.

The rotating part of the conveyor is sometimes called simply an auger.



**BUCKET ELEVATORS**

A bucket elevator, also called a grain leg, is a mechanism for hauling solids or flow-able bulk materials (most often grain or fertilizer) vertically.

It consists of:

1. Buckets to contain the material;
2. A belt to carry the buckets and transmit the pull;
3. Means to drive the belt;
4. Accessories for loading the buckets or picking up the material, for receiving the discharged material, for maintaining the belt tension and for enclosing and protecting the elevator.

A bucket elevator can elevate a variety of bulk materials from light to heavy and from fine to large lumps. A centrifugal discharge elevator may be vertical or inclined. Vertical elevators depend entirely on the action of centrifugal force to get the material into the discharge chute and must be run at speeds relatively high. Inclined elevators with buckets spaced apart or set close together may have the discharge chute set partly under the head pulley. Since they don't depend entirely on the centrifugal force to put the material into the chute, the speed may be relatively lower.

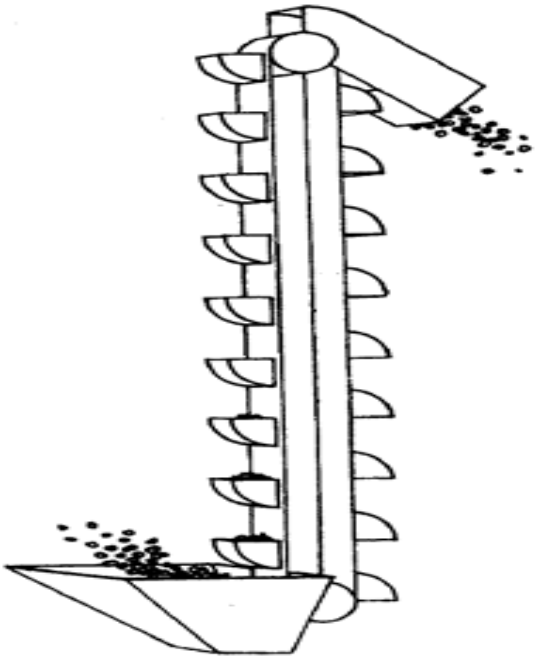
Nearly all centrifugal discharge elevators have spaced buckets with rounded

bottoms. They pick up their load from a boot, a pit, or a pile of material at the foot pulley.

The buckets can be also triangular in cross section and set close to on the belt with little or no clearance between them. This is a continuous bucket elevator. Its main use is to carry difficult materials at slow speed.

Early bucket elevators used a flat chain with small, steel buckets attached every few inches. Current construction uses a rubber belt with plastic buckets. Pulleys several feet in diameter are used at the top and bottom. The top pulley is driven by an electric motor.

Bucket Elevators



# GATE QUESTIONS

**Q.1** Sticky materials are transported by
(a) Apron conveyor (b) Screw conveyor
(c) Belt conveyor (d) Hydraulic conveyor
**(GATE 2007)**

| ANSWER KEY |
|---|
| **1** |
| (b) |

# EXPLANATIONS

**Q.1 (b)**

Use of ribbon flights in screw conveyor allow sticky materials to be handled.