Module # 5

SEPARATION EQUIPMENTS: GENERAL DESIGN CONSIDERATIONS OF CYCLONE SEPARATORS, CENTRIFUGES, SEPARATION EQUIPMENTS

Lecture 1: GENERAL DESIGN CONSIDERATION OF CYCLONE SEPARATORS

1. INTRODUCTION

Chemical processes consist of reaction stages and/or separation stages in which the process streams are separated and purified. Such separations involve physical principles based on differences in the properties of the constituents in the stream. Heterogeneous mixtures consist of two or more phases which have different composition. These mixtures consist of components that do not react chemically and have clearly visible boundaries of separation between the different phases. Components of such mixture can be separated using one or more appropriate techniques. These separation processes includes Gas-Liquid (vapor-liquid) separation, Gas-Solid separation (vapor-solid), Liquid-Liquid separation (immiscible), Liquid-Solid, and Solid-Solid separation etc. This separation can be done by exploiting the differences in density between the phases. Gravitational force or centrifugal force can be used to enhance the separation. The separation units can be either horizontal or vertical. The main techniques used to separate the phases, and the components within the phases, are discussed in details.

The principle methods for the separation of such mixtures could be classified as:

1. Cyclone separator, 2. Gas-Liquid separator, 3. Liquid-Liquid separator
4. Gravity separator, 5. Centrifugal separator, 6. High speed tubular centrifuge
7. Scrubbers 8. Electrostatic precipitator, 9. Hydro cyclone

2. CYCOLNE SEPARATOR

Cyclone separators provide a method of removing particulate matter from air or other gas streams at low cost and low maintenance. Cyclones are somewhat more complicated in design than simple gravity settling systems, and their removal efficiency is much better than that of settling chamber. Cyclones are basically centrifugal separators, consists of an upper cylindrical part referred to as the barrel and a lower conical part referred to as cone (figure 5.1). They simply transform the inertia force of gas particle flows to a centrifugal force by means of a vortex generated in the cyclone body. The particle laden air stream enters tangentially at the top of the barrel and travels downward into the cone forming an outer vortex. The increasing air velocity in the outer vortex results in a centrifugal force on the particles separating them from the air stream. When the air reaches the bottom of the cone, it begins to flow radially inwards and out the top as clean air/gas while the particulates fall into the dust collection chamber attached to the bottom of the cyclone.

Cyclones have no moving parts and available in many shapes and sizes, for example from the small 1 and 2 cm diameter source sampling cyclones which are used for particle size analysis to the large 5 m diameter cyclone separators used after wet scrubbers, but the basic separation principle remains the same.

Three different types of cyclone are shown in figure 5.2. First figure i.e. 5.2a shows a cyclone with a tangential entry. These types of cyclones have a distinctive and easily recognized form and widely used in power and cement plants, feed mills and many other process industries.

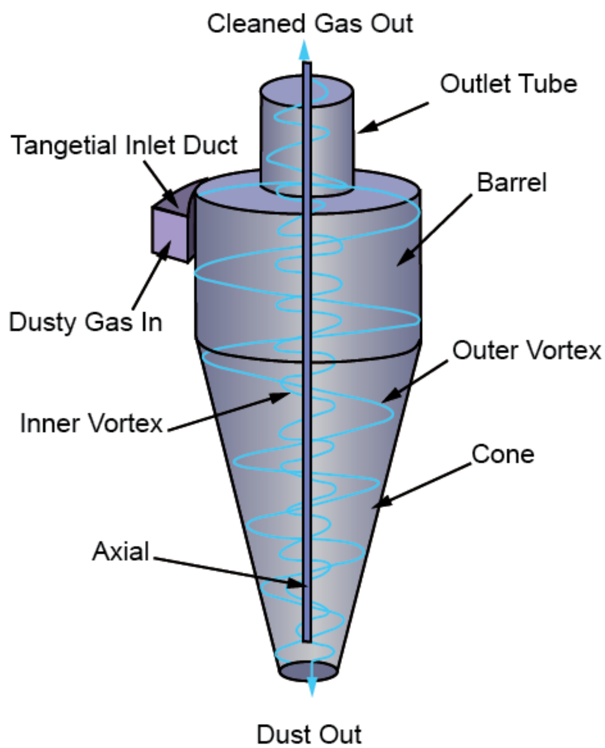


Figure 5.1: schematic diagram of cyclone separator

Figure 5.2b shows the axial entry cyclones, the gas enter parallel to the axis of the cyclone body. In this case the dust laden gases enter from the top and are directed into a vortex pattern by the vanes attached to the central tube. Axial entry units are commonly used in multi cyclone configuration, as these units provide higher efficiencies.

Another type of larger cyclonic separator shown in figure 5.2c is often used after wet scrubbers to trap particulate matter entrained in water droplets. In this type, the air enters tangentially at the bottom, forming vertex. Large water droplets are forced against the walls and are removed the air stream.

Cyclone collectors can be designed for many applications, and they are typically categorized as high efficiency, conventional (medium efficiency), or high throughput (low efficiency). High efficiency cyclones are likely to have the highest-pressure drops of the three cyclone types, while high throughput cyclones are designed to treat large volumes of gas with a low-pressure drop. Each of these three cyclone types have the same basic design. Different levels of collection efficiency and operation are achieved by varying the standard cyclone dimensions.

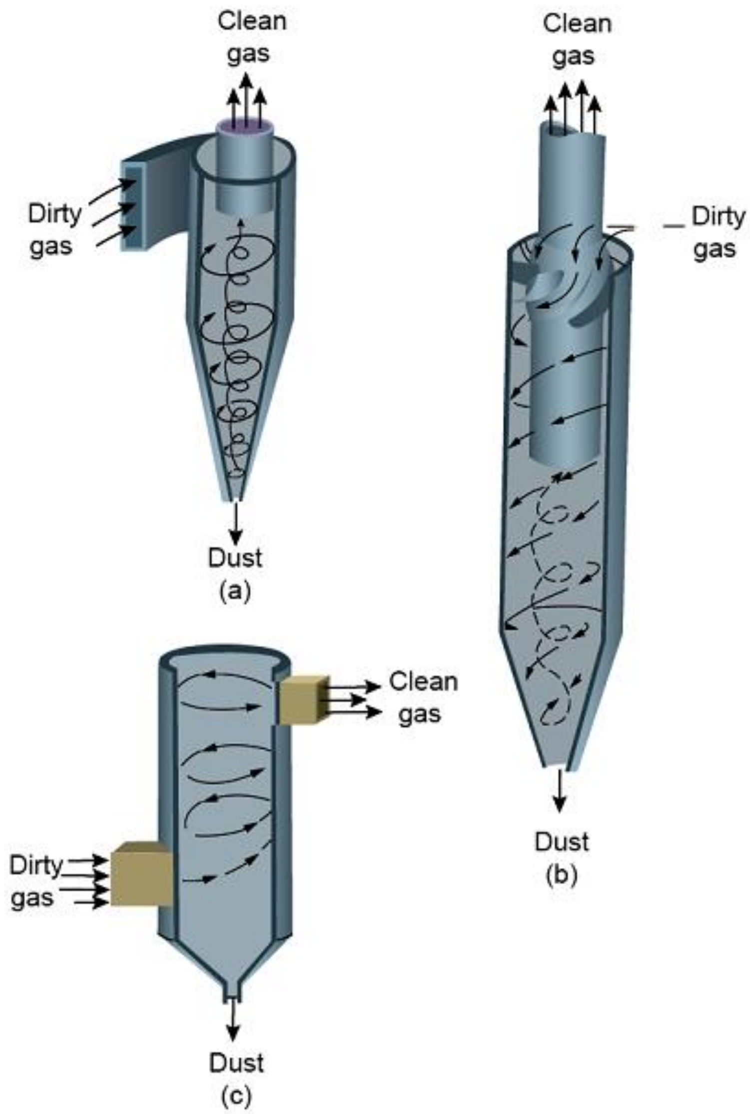


Figure 5.2: different types of cyclone

The collection efficiency of cyclones varies as a function of density, particle size and cyclone design. Cyclone efficiency will generally increase with increases in particle size and/or density; inlet duct velocity; cyclone body length; number of gas revolutions in the cyclone; ratio of cyclone body diameter to gas exit diameter; inlet dust loading; smoothness of the cyclone inner wall.

Similarly, cyclone efficiency will decrease with increases in the parameters such as gas viscosity; cyclone body diameter; gas exit diameter; gas inlet duct area; gas density; leakage of air into the dust outlet.

The efficiency of a cyclone collector is related to the pressure drop across the collector. This is an indirect measure of the energy required to move the gas through the system. The pressure drop is a function of the inlet velocity and cyclone diameter. Form the above discussion it is clear that small cyclones are more efficient than large cyclones. Small cyclones, however, have a higher pressure drop and are limited with respect to volumetric flow rates. Another option is arrange smaller cyclones in series and/or in parallel to substantially increase efficiency at lower pressure drops. These gains are somewhat compensated, however, by the increased cost and maintenance problems. Also these types of arrangements tend to plug more easily. When common hoppers are used in such arrangements, different flows through cyclones can lead to reentrainment problems. A typical series arrangement is shown in figure 5.3. In such arrangements large particle can be arrested in the first cyclone and a smaller, more efficient cyclone can collect smaller particles. Due to that it reduces dust loading in the second cyclone and avoids problems of abrasion and plugging. Also, if the first cyclone is plugged, still there will be some collection occurring in the second cyclone. The additional pressure drop produced by the second cyclone adds to the overall pressure drop of the system and higher pressure can be a disadvantage in such series system design. Cyclone efficiency can also be improved if a portion of the flue gas is drawn through the hopper. An additional vane or lower pressure duct can provide this flow. However, it may then become necessary to recirculate or otherwise treat this as purge exhaust to remove uncollected particulate matter.

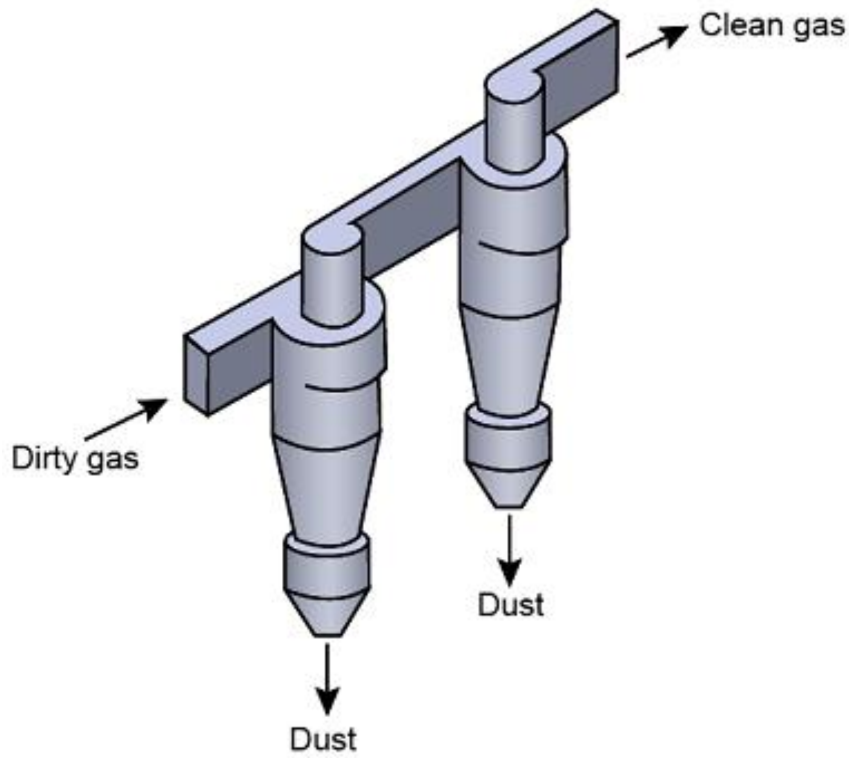


Figure 5.3: Typical series arrangement

2.1 Cyclone performance

Cyclones are basically centrifugal separators. They simply transform the inertia force of gas particle to a centrifugal force by means of a vortex generated in the cyclone body. The particle laden gas enters tangentially at the upper part and passes through the body describing the vortex. Particles are driven to the walls by centrifugal forces (an expression for this force is given blow eq. 5.1), losing its momentum and falling down to the cyclone leg. In the lower section, the gas begins to flow radially inwards to the axis and spins upwards to the gas outlet duct.

$$F = \frac{\rho_p d_p^3 v_p^2}{r} \qquad (5.1)$$

ρ_p = particle density, (kg/m^3)
d_p = particle diameter, inches (µm)
v_p = particle tangential velocity (m/s)
r = radius of the circular path, (m)

The main variables describing the cyclone performance are pressure drop, efficiency and cut diameter. Equations involving each of these parameters are provided in this section.

2.1.1 Cut diameter

The cut diameter of the cyclone is defined as the size of the particles collected with 50% collection efficiency. It is an indicator of the size range of particles that can be collected. It is a convenient way of defining as it provides information on the effectiveness for a particle size range. A frequently used expression for cut off diameter is

$$d_{pc} = \left(\frac{9\mu B_c}{2\pi N v_i \left(\rho_p - \rho \right)} \right)^{1/2} \qquad (5.2)$$

μ = viscosity (Pa.s); B_c = inlet width (m)
N = effective number of turns (5-10 for common cyclone)
v_i = inlet gas velocity (m/s)
ρ_p = particle density (kg/m^3); ρ = gas density (kg/m^3)

A value of N, number of turns, must be known in order to solve equation (5.2) for d_{pc}. Given the volumetric flow rate, inlet velocity, and dimension of the cyclone, N can be easily calculated. Values of N can vary from 1 to 10, with typical values in the 4-5 range.

2.1.2 Collection efficiency

The collection or separation efficiency is most properly defined for a given particle size. As mentioned, fractional efficiency is defined as the fraction of particles of a given size collected in the cyclone, compared to those of that size going into the cyclone. Experience shows that collection efficiency of cyclone separator increases with increasing particle mean diameter and density; increasing gas tangential velocity; decreasing cyclone diameter; increasing cyclone length; extraction of gas along with solids through the cyclone legs.

Several equations have been developed to predict the collection efficiency in cyclones through correlation equations. The following section describes two methods of calculating cyclone efficiency. First the theory proposed by Leith and Licht (1973) for calculating fractional efficiency will be discussed and then a convenient graphical method developed by Lappel (1951) will be presented.

The fractional efficiency equation of Leith and Licht is given as:

$$E_i = 1 - e^{\left[-2(C\psi)^{1/(2n+2)} \right]} \qquad (5.3)$$

Where C = cyclone dimension factor
ψ = impaction parameter
n = vortex exponent

$$\psi = \frac{\rho_p d_p^2 v_i}{18 \mu D_c}(n+1) \qquad (5.4)$$

In this expression, *c* is a factor that is a function only of the cyclone's dimensions. The symbol ψ expresses characteristics of the particles and gas and is known as *inertia or impaction parameter*. The value of *n* is dependent on the cyclone diameter and temperature of the gas stream. And ρ_p times v_i expresses the particle's initial momentum. Although the calculation involved in this method are tedious but are straightforward.

A more popular and an older method of calculating cyclone fractional efficiency and overall efficiency was developed by Lappel (1951). He first computed the ratios $d_p/[d_p]_{cut}$ from equation 5.2 and it is observed that cyclone efficiency correlates in a general way with this ratio. For a typical cyclone, efficiency will increase as the ratio increases. The preceding correlation has been found to agree well with experimental data. To calculate fractional efficiency, the following procedure given below should be used. The sum of the products in the rightmost column will give the overall efficiency.

Lappel calculation procedure				
d_p range	Wt fraction in range	$d_p/[d_p]_{cut}$	E_i for each d_p from experiment or Lappel's method, %	Wt fraction × E_i

Lecture 2: GENERAL DESIGN CONSIDERATIONS

2.1.3 Pressure drop

Pressure drop across the cyclone is of much importance in a cyclone separator. The pressure drop significantly affects the performance parameters of a cyclone. The total pressure drop in a cyclone will be due to the entry and exit losses, and friction and kinetic energy losses in the cyclone. Normally most significant pressure drop occurs in the body due to swirl and energy dissipation. There have been many attempts to predict pressure drops from design variables. The idea is that having such an equation, one could work back and optimize the design of new cyclones. The empirical equation given by Stairmand (1949) can be used to estimate the pressure drop.

$$\Delta P = \frac{\rho_f}{203}\left\{u_1^2\left[1+\ 2\phi^2\left(\frac{2r_t}{r_e}-1\right)\right]+2\ u_2^2\ \right\} \qquad (5.5)$$

ΔP = cyclone pressure drop

ρ_f = gas density; u_1 = inlet duct velocity; u_2 = exit duct velocity

r_t = radius of circle to which the centre line of the inlet is tangential; r_e = radius of exit pipe

ϕ = cyclone pressure drop factor

$\Psi = f_c\ (A_s/A_1)$

f_c= friction factor, taken as 0.005 for gases

A_s = surface area of cyclone exposed to the spinning fluid

For design purposes this can be taken as equal to the surface area of a cylinder with the same diameter as the cyclone and length equal to the total height of the cyclone

A_t = area of inlet duct

Above equation is for the gas flowing alone, containing no solids. The presence of solid will increase the pressure drop over that calculated using equation (5.5), depending on the solids loading.

Alternative design equation for cycl ones occasionally used is:

$$\Delta P = \frac{0.0027\ \mathrm{q}^2}{\mathrm{k_c} D_c^2 B_c H_c (L_c / D_c)^{1/3} (Z_c / D_c)^{1/3}} \qquad (5.6)$$

Where q = volumetric flow rate

B_c = inlet width, H_c = inlet height, D_c = outlet diameter, Z_c = cone length, L_c = cylinder length

In this equation k_c is a dimensionless factor express of cyclone inlet vanes.
K_c = 0.5 for cyclones without vanes; K_c = 1.0 for cyclone vanes that do not expand the entering gas or touch the outer wall; however K_c = 2.0 for cyclone vanes that expand and touch the outlet all. The above equation when compared with experimental data shows poor correlation coefficient (Theodore, 2008).

Another form of the empirical pressure drop an equation which is being widely used:

$$\Delta P = K_c \rho v_i^2 \qquad (5.7)$$

K_c = a proportionality factor, if the ΔP is measured in inches, it varies from 0.013 to 0.024, with 0.024 the norm.

Design considerations

- Select either the high efficiency or high throughput design, depending on the performance required
- Obtain an estimate of the particle size distribution of the solids in the stream to be treated.
- Calculate the number of cyclone needed in parallel.
- Estimate the cyclone diameter for an inlet velocity of say 15 m/s. Then obtain the other cyclone dimensions from the graphs (refer to page 452, Sinnott, 2005)
  Then estimate the scale up factor for the transposition of the figure. (refer to page 452 and 453, Sinnott, 2005)
- Estimate the cyclone performance and overall efficiency, if the results are not satisfactory try small diameter.
- Calculate the cyclone pressure drop and check if it is within the limit or else redesign.
- Estimate the cost of the system and optimize to make the best use of the pressure drop available (Sinnott, 2005).

Example 5.1: Estimate the cut diameter and overall collection efficiency of a cyclone given the particle size distribution of dust from cement kiln. Particle size distribution and other pertinent data are given below (refer: Theodore, 2008).

Avg particle size in range d_p, µm	**1**	**5**	**10**	**20**	**30**	**40**	**50**	**60**	**>60**
Wt percent	03	20	15	20	16	10	06	03	07

Gas viscosity = 0.02 Cp; Specific Gravity of the particle = 3.0

Inlet gas velocity of cyclone = 48 ft/sec

Effective number of turns within cyclone = 5

Cyclone diameter = 8 ft

Cyclone inlet width = 2 ft

Solution:

Cut size d_{pc} can be calculated from the following equation

$$d_{pc} = \left[\frac{9\mu B_c}{2\pi N u_i (\rho_p - \rho)} \right]^{1/2}$$

and collection efficiency as a function of the ratio of particle diameter to cut diameter can be obtained by

$$E = \frac{1}{1 + (d_{pc} / d_p)^2}$$

First determine the value of

$$\rho_p - \rho = \rho_p = 3(62.4) = 187.2 \text{ lb/ft}^3$$

$$d_{pc} = \left[\frac{9\mu B_c}{2\pi N u_i (\rho_p - \rho)} \right]^{1/2}$$

$$= \left[\frac{9 \times 0.02 \times 6.72 \times 10^{-4} \times 2}{2\pi \times 5 \times 48 \times 187.2} \right]^{1/2}$$

$= 2.92 \times 10^{-5}$ft

$= 8.9$ µm

Table 5.1:

d_p, µm	w_i	d_p/d_{pc}	E_i, %	w_iE_i%
1	0.03	0.11	0	0.0
5	0.20	0.55	23	4.6
10	0.15	1.11	55	8.25
20	0.20	2.22	83	16.6
30	0.16	3.33	91	14.56
40	0.10	4.44	95	9.5
50	0.06	5.55	96	5.7
60	0.03	6.66	98	2.94
>60	0.07	-	100	7.0

The overall collection efficiency is therefore

$$E = \sum w_i E_i = 0 + 4.6 + 8.25 + 16.6 + 14.56 + 9.5 + 5.7 + 2.94 + 7.0$$

$$= 69.15\% = 0.6915$$

3. GAS LIQUID SEPARATOR

A gas-liquid separator is a vessel into which a liquid and gas mixture is fed and wherein the liquid is separated by gravity, falls to the bottom of the vessel, and is withdrawn. The gas travels upward at a design velocity which minimizes the entrainment of any liquid droplets in the gas as it exits the top of the vessel. The gas liquid feed mixture can be under pressure at the inlet of the separator. In ideal case gas liquid separator should give the maximum liquid recovery and should have low pressure of feed stream to atmospheric pressure at the discharge from the separator. This may not be accomplished in a single stage and may need number of stages. Usually from economic point of view, number of stages can be 3 to 4. As far as possible the gas and liquid should flow without any application of motive power.

While designing it is assumed that gas and liquid phase are in thermodynamic equilibrium with each other

Let the feed of gas liquid mixtures: f kg moles

Let f_1, f_2, f_3 are kg moles of the components in feed mixtures

$F = f_1 + f_2 + f_3$

Let the moles of gas separated be: g kg moles/hr

g_1, g_2, g_3 are kgmoles of components in gas

$G = g_1 + g_2 + g_3$

Let the moles of liquid separated be: l kg moles/hr

l_1, l_2, l_3 are kgmoles of components in liquid

$L = l_1 + l_2 + l_3$

$l_1 = f_1 – g_1$; $l_2 = f_2 – g_2$ and so on……..

$g_1 = G.k_1 (l_1/L)$; $g_2 = G.k_2 (l_2/L)$ and so on……

k_1, k_2, k_3 are defined as

$$K = \left(\frac{\text{vapour pressure of the component}}{\text{total pressure}}\right)$$

Therefore,

$$g_1 = \frac{k_1 f_1}{\left(\frac{L}{G} + k_1\right)} \quad ; \qquad g_2 = = \frac{k_2 f_2}{\left(\frac{L}{G} + k_2\right)} \qquad \text{and so on}$$

The important dimensions of a gas liquid separator are diameter and height. The diameter of the vertical gas liquid separator depends on maximum allowable velocity. The limiting velocity is the terminal velocity of maximum size droplet with permissible entrainment.

$$\text{Cross sectional area} = A = \frac{V.\rho_g}{G} = \frac{\pi D^2}{4} \qquad (5.8)$$

Whereas in case of horizontal separator cross sectional area estimated is only that of the free space above the liquid. The liquid seal depth has also considers while estimating the diameter of the vessel. Height of gas phase above the liquid should be sufficient to allow the disengagement of entrained droplets. The depth of the liquid level in the vessel depends upon the retention time and is given by:

$$\text{liquid depth} = \frac{\text{liquid flow rate} \times \text{retention time}}{\text{cross sectional area of the separator}}$$

4. LIQUID-LIQUID SEPARATOR

Separating liquid-liquid dispersions can be difficult depending on the physical properties of the two liquid phases. The specific gravity, viscosity and interfacial tension (IFT) of the two liquid phases are important parameters in determining how easily the two liquids can be separated. In liquid –liquid separator interface position can be fixed at the desired level by fixing the difference in elevation ΔH between liquid phase overflow and heavy phase under flow. The high specific gravity liquid prevents the light phase from leaving the separator via the bottom outlet. The position of interface in the liquid-liquid separator depends upon the difference in the elevation ΔH between the light and heavy phase (figure 5.4). Pressure change in the separator can affect the static pressure difference between the two layers and this change in the pressure affects the interface position in the separator.

Size of this type of separator depends upon the retention time needs for each phase to settle within the separator. Horizontal unit is most popular to carry out the separation process efficiently and avoids the mixing of the two phases (light and heavy). If the phases can be easily separated, vertical unit also shows good performance.

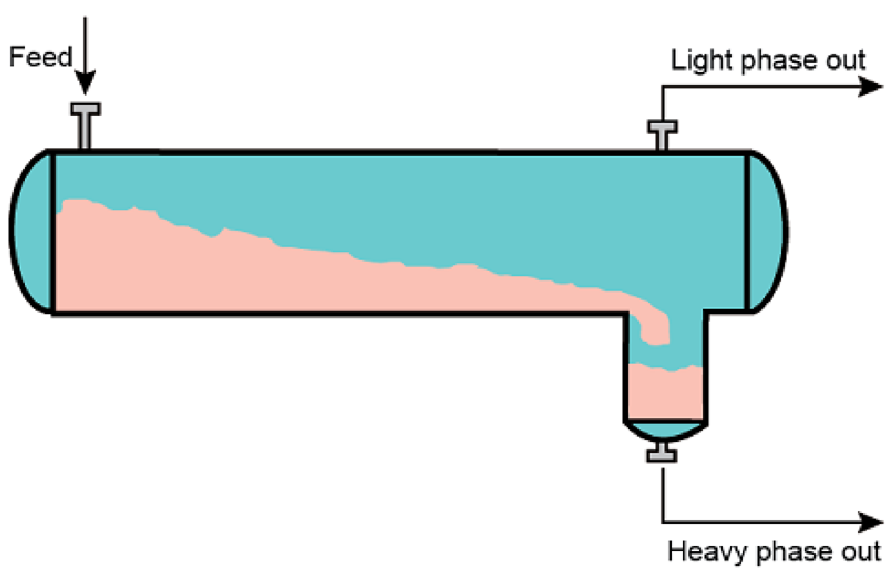


Figure 5.4: Two phase (Liquid-Liquid) separator

Let ρ_L; ρ_h specific gravity of light and heavy phase respectively (*also refer the figure given in the example 5.2*)

H_1 and H_2 height of heavy and light phase

C – Distance from the bottom of the separator to the reference line

X – light phase liquid head in the separator

Y – heavy phase liquid head in the separator.

Static head exerted by point ‘a and b’ over the reference line are equal.

$H_1 \rho_h = X\rho_L + (Y + C) \rho_h$

Similarly, $H_2 = (X + Y + C)$

The difference between overflow and under flow head

$\Delta H = H_2 - H_1$

$$\Delta H = X + Y + C - \frac{[X\rho_L + (Y + C)\rho_h]}{\rho_h}$$

In case if heavy phase is water

$\rho_h = 1.0 \qquad \Delta H = (1-\rho_L)\, X$

Lecture 3: CENTRIFUGES

Example 5.2: Design a liquid-liquid gravity separator which can handle a two phase liquid stream of 0.5 $m^{3/}$min. The feed contains 45% by volume of light phase and 55% by volume of a heavy phase. Densities of light (ρ_l) and heavy phase (ρ_h) are 900 and 1150 kg/m^3 respectively. Required settling time of light phase is 5 min while the settling time for heavy phase is 4 min.

Solution:

Feed rate to separator = 0.5 m^3/min

Light phase in feed = 0.5*0.45 = 0.22m^3/min

Heavy phase in feed = 0.5*0.55 = 0.28 m^3/min

Settling time for light phase = 5 min

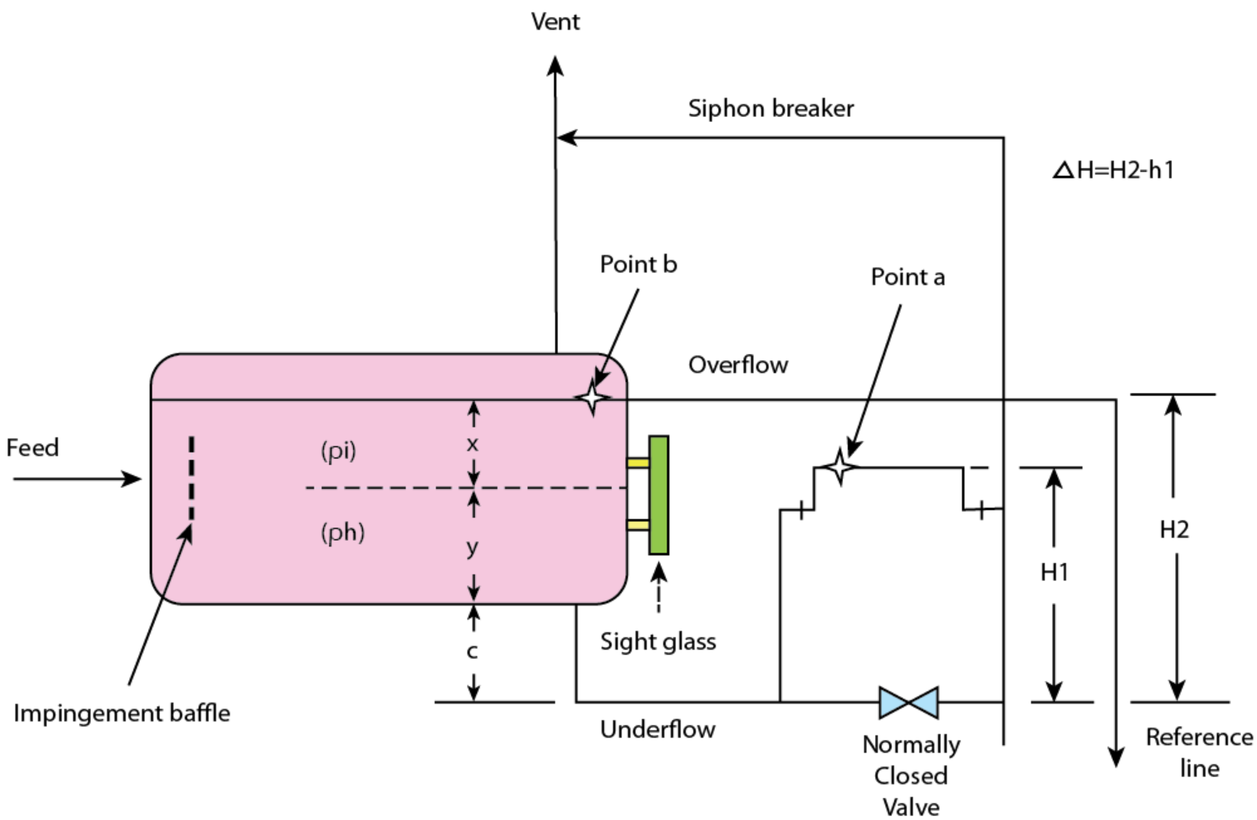
Volume of light phase in separator = 0.22*5 = 1.1 m^3

Design Volume can be taken as 20% more

Design volume of light phase = 1.2*1.125 = 1.32 m^3

Volume of heavy phase = 0.27*4 = 1.08 m^3

Design volume of heavy phase = 1.2*1.08 = 1.3m^3



Total design volume of separator = 1.32+1.3 = 2.62 m^3
Assume free board volume = 0.288 m^3
Separator vessel volume = 2.62+0.288 = 2.908 m^3
For a horizontal tank with elliptical heads having major to minor axis ratio 2:1

$$\left[\frac{\pi}{4}D^2L+\frac{\pi}{12}D^3\right]=2.4$$

If we use L = 1.3*D we have:

$$\left[\frac{\pi}{4}D^2 1.3D+\frac{\pi}{12}D^3\right]=2.4$$

i.e. D = 1.23 m

Length of cylindrical vessel = 1.23*1.3 = 1.6 m

The interface for the required volumes of the light and heavy phase in the separator should be:

X = 0.46; Y = 0.54

Let the distance from the bottom of the separator to the reference line = C = 0.6 m

Light phase overflow height from the reference line

H_2 = X+Y+C = 0.41+0.59+0.6

The distance between overflow leg and underflow leg

$$\Delta H=\left[1-(\frac{\rho_l}{\rho_h})\right]X$$

$$\Delta H=\left[1-(\frac{0.9}{1.15})\right]0.46 \ = \ 0.1 \text{ m}$$

Heavy phase underflow height

$$H_1=\left[\frac{X\rho_l+(Y+C)\rho_h}{\rho_h}\right]$$

$$H_1=\left[\frac{0.46*0.9+(0.54+0.6)1.15}{1.15}\right] = 1.5 \text{ m}$$

For practical reasons and to compensate for fabrication problem. Height H_1 should be designed 25 mm short and make the adjustable spool with 50 mm offset. Final heavy phase underflow height

$$H_1 = 1.5 - 0.025 + 0.05 = 1.525 \text{ m}$$

Let the interface in the separator is located 0.15 m below the designed evaluation during the initial run. To raise the interface level, raise the adjustable spoul piece by

$$\Delta H = \left[1 - \left(\frac{\rho_l}{\rho_h}\right)\right] X$$

$$\Delta H = \left[1 - \left(\frac{0.9}{1.15}\right)\right] 0.15 = 0.032 \text{ m}$$

Thus for adjusted ΔH value of 0.032 m, the interface position in the separator moves by 0.15m in the same direction.

5. GRAVITY SEPARATION

Separation of two immiscible liquids or slurry containing fine solids can also be carried out by making use of the density difference. In case of these two phase system the continuous phase is a liquid while the disperse phase is solid or other immiscible liquid. Solid-liquid separation processes are generally based on either one or a combination of gravity settling, filtration, and centrifugation principles. In gravity settling separation, solid particles will settle out of a liquid phase if the gravitational force acting on the droplet or particle is greater than the drag force of the fluid flowing around the particle (sedimentation).

The solid phase is assumed to having the higher density and its percentage should not be more than 25% by volume. Because upto the concentration of 1.0 -2.0% (vol.), hindered settling is negligible but above 10% (approx.), hindered settling becomes more dominant. The particle size or droplet to be separated should be less than 500 µm. In this separation, the sedimentation velocity or terminal settling velocity or sedimentation rate can be estimated using:

$$u_g = \frac{d_p^2(\rho_s - \rho_L)}{18\mu} \qquad (5.9)$$

Above equation shows that if the density difference is negative. The particle will move upwards with a velocity u_g toward the surface of the liquid i.e. the particle will move in the opposite direction to that in which the force of gravity. The sedimentation rate depends upon the particle density, density of continuous phase and the viscosity of the continuous phase.

While in case of batch operation (i.e. separation by gravity) time required to achieve complete separation for the particle which is initially at the surface must be sufficient enough to travel through the entire depth of the liquid (figure 5.5a). Time required for the particle sedimentation can be reduced by decreasing the depth of the chamber. Decrease in depth can be achieved by increasing the length of the chamber and keeping same residence time, but with decrease in depth operation may become uncontrollable.

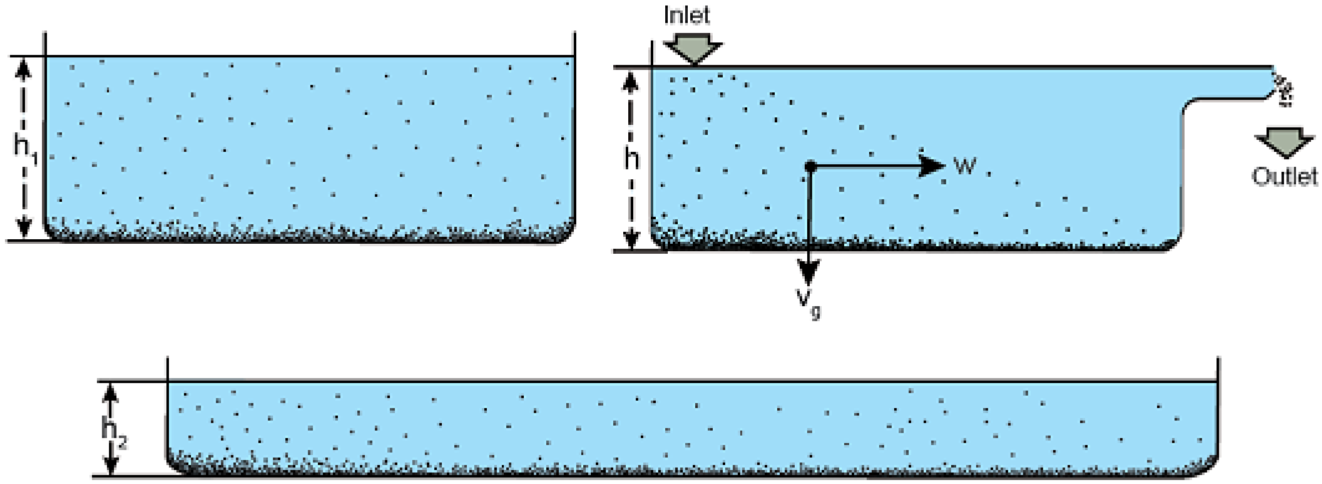


Figure 5.5a:

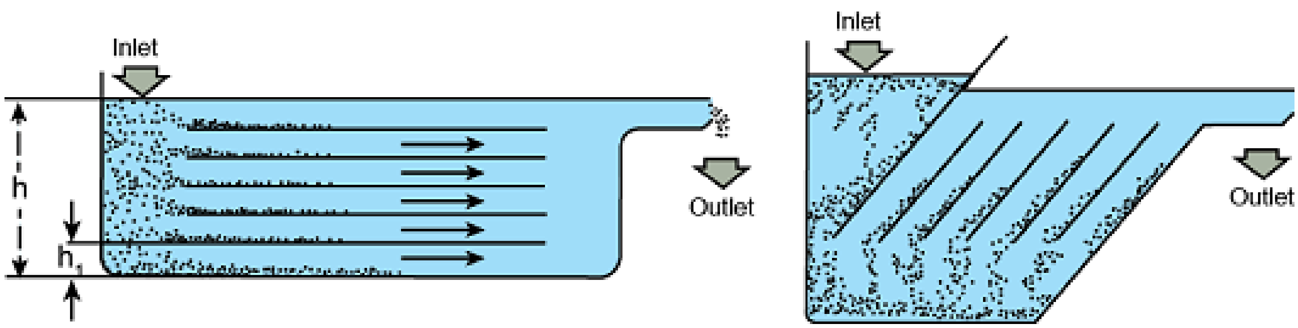


Figure 5.5b:

For continuous operation time provided for separation (by gravity) in the vessel is given by:

$$t = \frac{V}{Q} = \frac{W\,h\,L}{Q} \qquad (5.10)$$

V - volume of tank $= W \times h \times L$

W - width of tank ; h- height of tank; L – length of tank

The desired particle size should be able to settle to the bottom of the tank

$$t = \frac{h}{u_{lim}}$$

$$t = \frac{W.\,h\,.L}{Q} = \frac{h}{u_{lim}}$$

$$\text{i.e.} \quad \frac{1}{u_{lim}} = \frac{W\,L}{Q} = \frac{A}{Q}$$

WL - surface area of the tank A.

Q - throughout rate

$Q = [u_{lim}\,.\,A]$

Above equation gives the maximum possible throughout for the desired particle size d_{lim} having a sediment velocity u_{lim}. From this it is observed that the height of the tank does not affect the throughput rate. Only alternative is throughput can be increased by increasing the surface area and it (surface area) can be increased by providing the horizontal or inclined plates in the tank as shown in figure 5.5b.

$Q = u_{lim}.\ N.\ A$

Where N - number of separation channels of area A.

In the above equation u_{lim} is fixed by the particle size and density, thus the throughput is proportional to total area N A.

6. CENTRIFUGAL SEPARATION

Centrifugal separation is similar to sedimentation, but some forces are added in order to get a better separation. In a centrifugal device, the centrifugal force is generated to increase the force acting on the particles. Utilization of centrifugal action for the separation of materials of different densities and phases, might be built in stationary and rotary types of equipment. In centrifugal separators, centrifugal forces act on particles is several times greater than gravity as it enters a cylindrical separator. This results in a much shorter separation time than could be accomplished solely by gravity. It has mainly been used to separate fluids in static state, i.e. specific volumes, which needed to be separated.

The centrifugal force acting on a particle is : $C_f = m\,w^2\,r$

Where, m – mass of particle; w – angular velocity of the particle; r – radial distance from axis of rotation

The influence of centrifugal force on the particle varies as the particle moves in the separation vessel. The relative centrifugal force can be defined as the force acting upon a particle in a centrifugal field in terms of multiple of its own weight in the gravitational field.

$$(C_f)_r = \frac{w^2 r}{g} \qquad (5.11)$$

$$w = 2\,\pi\,N/60$$

where, N is revolution per minute

$$(C_f)_r = 1.11 \times 10^{-5} \quad N^2.r \qquad (5.12)$$

here, r is radius from the axis of rotation,

the relative mass of particle or effective mass of particle $= V\,(\rho_P - \rho_f)$

V – volume of particle; ρ_P – density of particle; ρ_f – density of fluid.

For a spherical particle: $V = \left(\frac{\pi d_p^3}{6}\right)$; d_p – particle diameter

The force producing the motion:

$$F = \frac{\pi d_p^3}{6}\left(\rho_P - \rho_f\right)\, w^2.r \qquad (5.13)$$

The force resisting the motion of particle: $F = 3\,\pi.\mu.d_P.u$

μ - viscosity of liquid; u – velocity of particle

The particle accelerates until it reaches a terminal velocity where the two forces are equal

$$\frac{\pi d_p^3}{6}\left(\rho_P - \rho_f\right)\, w^2.r = 3\,\pi.\mu.d_P.u \qquad (5.14)$$

$$u = \frac{d_p^2\left(\rho_p - \rho_f\right) w^2 r}{18\mu} \qquad (5.15)$$

The terminal velocity u is proportional to radius r. The true terminal velocity, u_s is reached after a short initial interval of time. If the density of particle is greater than density of fluid, the particle moves outward from the axis of rotation (i.e. acceleration is +ve), and it the density of particle is less than the density of fluid, the particle moves in ward toward the axis (i.e. retardation is +ve). For the centrifuges carrying relatively thin layers of liquid, the velocity of particle can be considered as approximately constant across such a layer. The radial distance travelled by a particle in time t seconds is x cm.

$$x = u.t = \frac{d_p^2\left(\rho_p - \rho_f\right) w^2.r.\,t}{18\mu} \qquad (5.16)$$

t = V/Q where, V – volume of liquid in the bowl at any given moment

Q – volumetric flow rate

Therefore, $\left(r_2 - r_1\right) = \dfrac{d_p^2\left(\rho_p - \rho_f\right)w^2.\,r}{18\mu}.\left(\dfrac{V}{Q}\right)$ here $(r_2 - r_1) = S$, is liquid layer thickness

Therefore,

$$Q = \frac{1}{18}\times\frac{d_p^2\left(\rho_p - \rho_f\right)}{\mu}.\left(\frac{w^2.\,r.V}{S}\right) \qquad (5.17)$$

From this equation, ideal performance of a centrifuge can be estimated

$$Q = k\ X\ Y, \quad \text{here } k \text{ is constant}$$

$X = \dfrac{d_p^2\left(\rho_P - \rho_f\right)}{\mu}$ this eqn. measured distance normal to the disc surface

&

$Y = \dfrac{w^2.r.V}{S}$ this eqn measured distance parallel to the disc surface

$$\frac{w^2.r.V}{S} = (C_f)_r\left(\frac{V}{S}\right) \qquad (5.18)$$

$$Q = \frac{1}{18}\times\frac{d_P^2\left(\rho_P - \rho_f\right)}{\mu}(C_f)_r\left(\frac{V}{S}\right) \qquad (5.19)$$

Above equation valid for the system follows Stoke's law and it predicts the maximum throughput at which a particle of given size is eliminated from the liquid stream.

Lecture 4: CENTRIFUGES (CONT.)

6.1 High speed tubular centrifuge

In case of high speed tubular centrifuge the velocity of the particle for size upto 10 micron does not perceptibility exceed 1 m/sec. The Reynolds number can be considered between 10^{-4} to 10 and this is the region where Stoke's law is applicable.

$$u = \frac{d_p^2\left(\rho_p - \rho_r\right)}{18\mu} w^2.r \qquad (5.20)$$

Above equation can be applied for high speed tubular bowl centrifuge. In this type of centrifuge the residence time estimated by V/Q introduces an error and because of that the performance evaluation on its basis may not be perfect. Another approach to evaluate the performance it to assume viscous flow in the bowl and use for its length some fraction of geometric length estimated empirically.

In case of a bowl of length L, having an internal wall radius r_b. The bowl is filled so that the liquid surface radius is r_L.

$$L = \frac{18}{\pi}\frac{Q}{d_p^2\left(\rho_p - \rho_r\right)}\frac{\mu}{w^2}\left[\frac{2.r_b^2\ \ln\left(r_b/r_L\right) - \left(r_b^2 - r_L^2\right)}{\left(r_b^2 - r_L^2\right)^2}\right] \qquad (5.21a)$$

$$Q = \frac{\pi}{18} d_p^2\left(\frac{\rho_p - \rho_r}{\mu}\right) w^2 L\left[\frac{\left(r_b^2 - r_L^2\right)^2}{2.r_b^2\ \ln\left(r_b/r_L\right) - r_b^2 + r_L^2}\right] \qquad (5.21b)$$

For the fallacious block flow pattern:

$$Q = \frac{\pi}{18} d_p^2\left(\frac{\rho_p - \rho_r}{\mu}\right) w^2 L\left[\frac{r_b^2 - r_L^2}{\ln\left(r_b/r_L\right)}\right] \qquad (5.22)$$

From the above equation it is clear that for a given centrifuge rotor with interior radius r_b, it is obvious that depth of the liquid layer may be varied i.e. different value of r_L can be used, which will have a great influence on the centrifuge performance.

If we consider the influence of the magnitude of r_L, one can write the expression for volumetric flow rate through the bowl in m^3/sec.

$$Q = \frac{\pi}{18} d_p^2\left(\frac{\rho_p - \rho_r}{\mu}\right) w^2 L.\ Z \qquad (5.23)$$

Z – Velocity parallel to the disc surface (cm/s) and its value can be calculated for various values of r_b/r_L ratio. For tubular bowl centrifuge the value ratio r_b/r_L lies between 0.3 to 0.5. The value of Z can be calculated using equation (5.21a,b), the ratio Z_1/Z_2 = 0.75 which shows the magnitude of the error caused by block flow condition. In general way we can assume an effective length of bowl as 0.5 times the geometric length, L.

Thus the practical value for the performance of a high speed tubular bowl centrifuge can be estimated as follows:

$$Q = \frac{\pi}{18} d_p^2 \left(\frac{\rho_p - \rho_r}{\mu} \right) w^2 L. \ (0.75 \times 0.5) \ \frac{\left(r_b^2 - r_L^2 \right)}{\ln\left(r_b / r_L \right)} \qquad (5.24)$$

Substitute $w = (2\pi N/60)$

and $\ln (r_b/r_L) = 2.303 \log_{10} (r_b/r_L)$

$d_p = (\eta/10^4)$

η – diameter of particle in micron

N – rpm of motor

Therefore the performance equation:

$$Q = 2.8 \times 10^{-12} \ \eta^2 \left(\frac{\rho_p - \rho_r}{\mu} \right) N^2 L. \left(\frac{\left(r_b^2 - r_L^2 \right)}{\log_{10}\left(r_b / r_L \right)} \right) \qquad (5.25)$$

Example 5.3:

Calculate the sedimentation rate in gravity separation and centrifugal separation for the particle size limiting to d_{lim}= 7 µm. The particle density = 1040 kg/m^3; liquid density = 1000 kg/m^3; Viscosity of continuous phase = 1x10^{-3} N-s/m^2

Solution:

For gravity separation

$$u_g = \frac{d_p^2(\rho_s - \rho_l)}{18\mu} g = \frac{(7\times10^{-6})^2(1040-1000)}{18\times0.001}\times9.81$$

$$= 1\text{x}10^{-6} \text{ m/sec}$$

In centrifugal separation, the sedimentation rate

$$u = \frac{r\omega^2}{g} u_g ; \omega = \frac{2\pi N}{60} = \frac{2\times3.14\times5000}{60} = 523.60$$

$$u = \frac{0.2\times523.6^2}{9.81}\times1\times10^{-6}$$

$$u = 0.558 \text{ x } 10^{-2} \text{ m/sec}$$

Ratio of sedimentation rate in centrifugal to gravity separator = $\dfrac{0.558\times10^{-2}}{1\times10^{-6}}$ = 5580

7. SCRUBBEERS

Scrubbers are most often used as an air pollution control device to remove particulate matters and chemicals from waste gas streams of stationary point source. They are also applied where the slurry is used in other parts of the process or where the mixture is in a slurry form. In some scrubbers are applied so that chemical reaction will be generated within the scrubbing action.

7.1 Orifice scrubbers

In an orifice scrubber also known as an impaction scrubber, the gas stream flows over the surface of a pool of scrubbing liquid. As the air impinges on the water surface, it entrains droplets of the liquid. The waste gas stream then flows upward and enters an orifice with a narrower opening than the duct. The orifice brings turbulence in the flow which atomizes the entrained droplets. The atomized droplets capture the particulate matters in the air stream. Air velocity in the orifice scrubbers can be controlled by using adjustable orifices. The main advantage of this type of scrubber is that it does not need the recirculation pump for the scrubbing liquid, which is the major contributor to operating costs for most of the scrubbers. The disadvantage is difficulty in removing the sludge generated during the scrubbing process. In most scrubber design waste continuously drains from the bottom. Orifice scrubber consists of a static pool for scrubbing liquids, so waste generated is removed with a sludge ejector, which operates like a conveyor belt.

7.2 Venturi scrubbers

Venturi scrubber has a converging-diverging section, in this type of system the cross sectional area of the channel decreases then increases along the length of the channel. The narrowest area in the channel is referred as the throat. In venturi scrubber, liquid is introduced slightly upstream of the throat or directly into the throat section. In the converging section the decrease in area causes the waste gas stream velocity and turbulence to increase (figure 5.6). The scrubbing liquid is atomized by high velocity air stream and improves the air –liquid contact. Further the air –liquid mixtures decelerate as it moves through the diverging section, which helps to creates particle droplet impacts and agglomeration of the droplets. The separation of the liquid droplets from the air stream takes place in the entrainment section. The entrainment section usually consists of cyclonic separator and mist eliminator. For venturi scrubber collection efficiency for the fine particulate matter is higher but the equipment is more expensive than spray tower, cyclonic or tray tower scrubber. High air velocity and turbulence in the venture scrubber throat result in high collection efficiencies ranging from 70 to 99% for particles larger than 1 μm in diameter and greater than 50% for submicron particles. Increasing the pressure drop increases the collection efficiency, but the system’s energy demand also increases leading to higher operational cost.

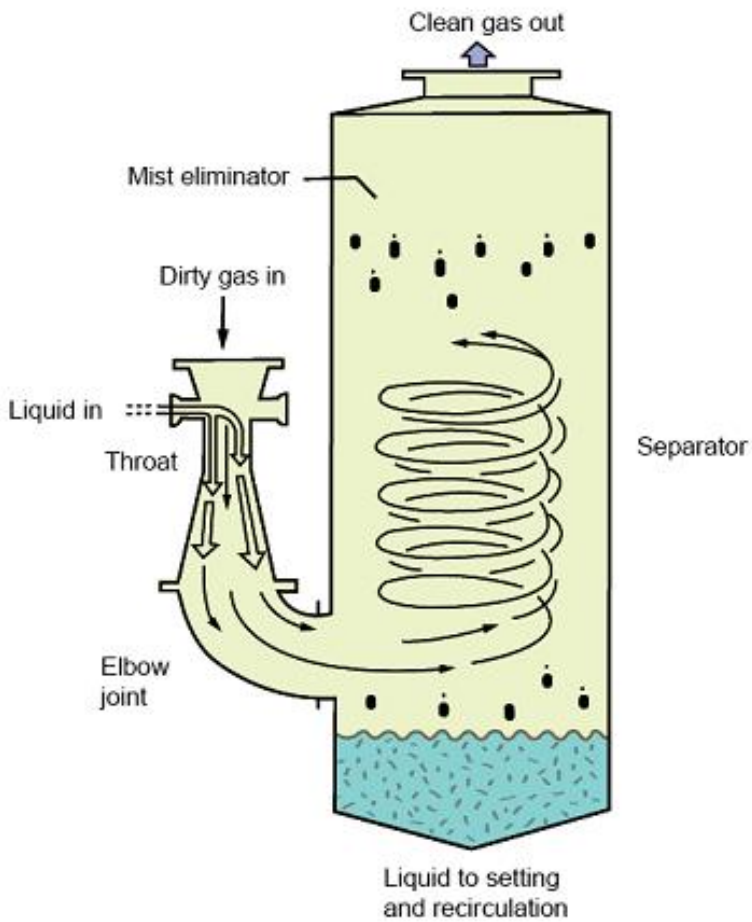


Figure 5.6: Venturi scrubber with cyclone separator and eliminator

7.3 Jet scrubber

In this scrubber, water flow is used in jet ejector to aspirate dusty air and to provide droplets for collecting particulates. Jet scrubber is used when it is not economical to used fan for a dust collection system (figure 5.7). Also, it can be used as a gas absorber.

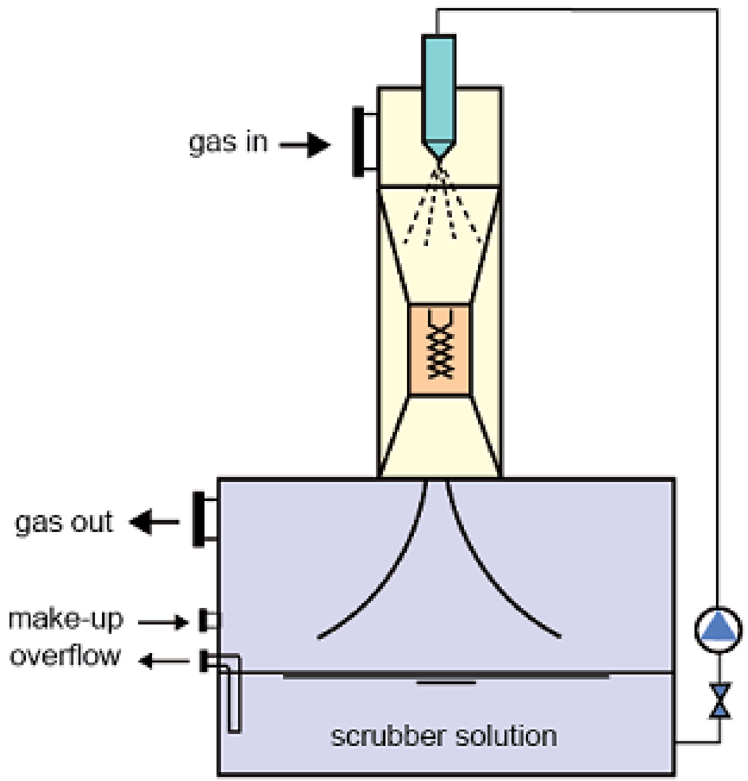


Figure 5.7: Jet scrubber

7.4 Dynamic scrubber

This type of scrubber is similar to spray towers, but only difference is that the dynamic scrubber uses power driven rotor that breaks the scrubbing liquids into finely dispersed droplets. Dynamic scrubbers are also known as mechanically –aided scrubbers or disintegrator. Liquid is sprayed into the suction of a fan and the wetted impeller and casing captures dust particles. Most dynamic scrubber systems humidify the waste air upstream of the rotor to reduce evaporation and particle deposition in the rotor area. This type of scrubber efficiently removes fine particulate matter, but the use of rotar in the system increases the maintenance cost. Pretreatment device, such as cyclone often used before dynamic scrubbers to remove the large particulate matter from the waste air stream. Collection efficiencies for dynamic scrubbers are similar to those for cyclonic spray towers.

Lecture 5: SEPARATION EQUIPMENTS

8. ELECTROSTATIC PRECIPITATOR

Electrostatic precipitator generally used to separate particulate matter that is easily ionized from a gas stream by using electrostatic charges. The principal actions are the charging of dust particles and forcing them to the collector plates. According to e.m.f. gradient the charge particle migrates and is attracted to collecting electrodes. Negatively charged particle are attracted towards the positive electrode and positively charged particles to the negative electrodes.

There are two types of electrostatic precipitators

1. Single stage unit 2. Two stage unit

Single stage electrostatic precipitator is very common. In this unit ionization and collection are combined. In two stage unit ionization is followed by collection. Each operation occurs in different part of the apparatus.

8.1 Single stage unit

Single stage unit precipitator consists of a rectangular shell or casing in which a number of grounded plates are suspended parallel to each other and has equal spacing between plates to form channels through which the gas flows (figure 5.8). High voltage discharge electrodes are suspended vertically between the plates from an insulated mounting frame. The distance between the grounded plates is in the 4 to 6 inch range. The voltage on the electrodes is between 40,000 to 60,000 volts. This voltage causes the gases to ionize and when this occurs the dust particle becomes negatively charged, the strength of this charge is a function of the dielectric characteristics of the dust. Some of the dust particles will have a high charge and the forces attract it to the grounded collecting plates. The time interval is determined by the distance the dust particle has traveled to the grounded collector plate and the magnitude of the charged dust particle. Some dust particles have higher forces that attract them to the collection plates at a greater efficiency rate than others. This may include other gases present in the process stream. For instance, some sulfur compounds in boiler gas increases collection efficiency. Apart from that the velocity of the gas passing through the plates also affects the efficiency of collection. For example, at 17 m/min gas velocity, only fifty percent of the particles may reach the collecting plates with an associated collection efficiency of 50%. Say at 9 m/min, the efficiency might be 95% and at 4 m/min it might be 99%. The pressure drop across the precipitator collection section usually in the range of 0.2 to 0.5 inches of water.

If the velocity varies, the efficiency will be lower across the sections with higher velocity, and the collection efficiency might be much lower than the predicted efficiency based on the average velocity. While designing electrostatic collector for single stage high voltage design, it is necessary to design the distribution baffles very carefully.

The two fundamental operations involved in electrostatic precipitation are: 1) electrical charging of the particles; 2) drift of the charged particles to the collector electrode under the influence of an electric field.

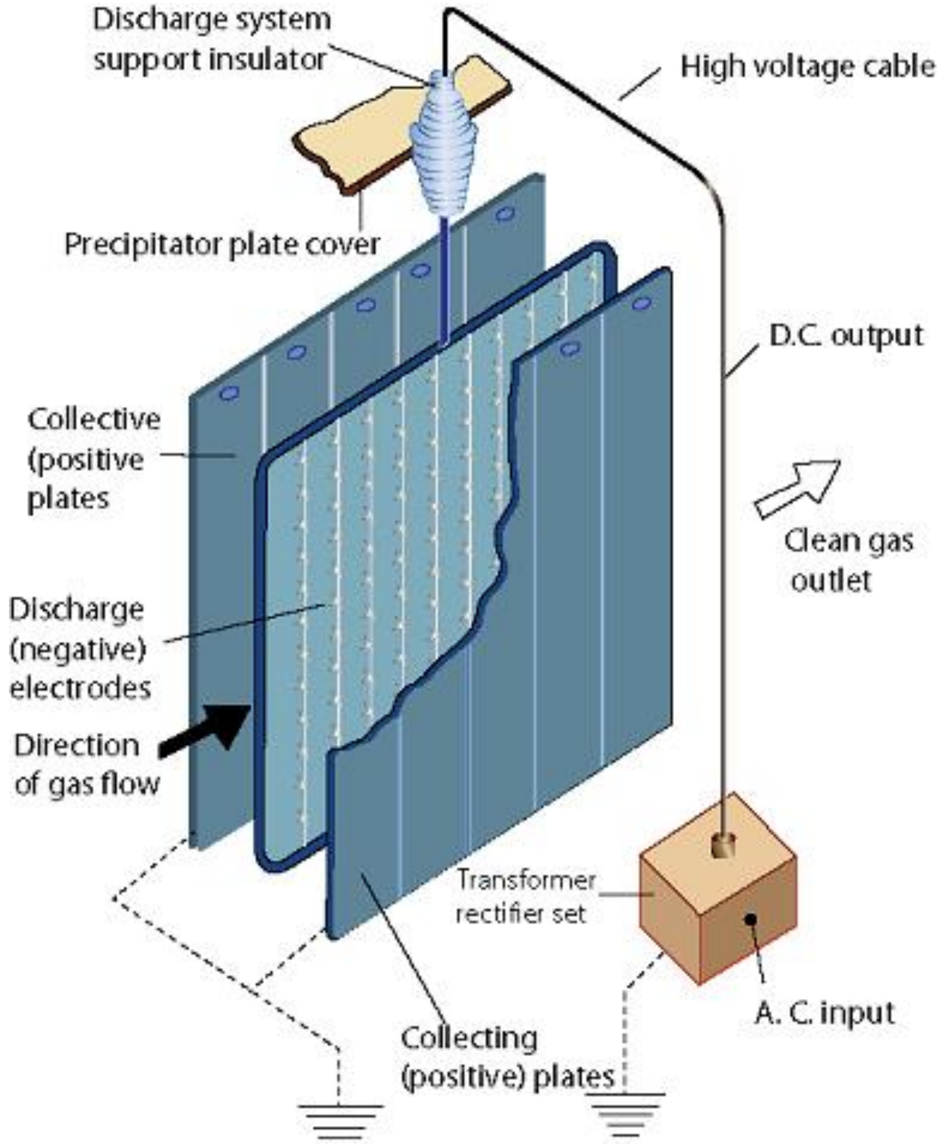


Figure 5.8: Single stage electrostatic precipitator

In the electrostatic precipitator, the electric field distribution in the inter electrode space is not uniform. In a wire and tube electrode system, it varies from a high value at the wire to a low value the tube surface. In this region suitable potential difference between such electrode results in electric break down of the gas close to the high electric field strength. A blue glow develops surroundings the wire which constitutes a corona, necessary for the operation of precipitator. A steady corona needs to be maintained for successful operation.

The necessary voltage can be calculated as:

$$V = 18\left[1 + \frac{0.3}{R_1^{0.5}}\right] R_1 \ln\left(\frac{R_2^2}{R_1}\right) \text{kV} \qquad (5.26)$$

R_1 – radius of wire electrode
R_2 – radius of collecting electrode
The velocity of collection for a conducting particle is given by:

$$w = 0.16\left[\frac{d_p^2\ V^2}{\mu}\right] \qquad (5.27)$$

μ - Viscosity of gas, N-s/m^2
The velocity of collection for a non conducting particle is given by:

$$w = 0.095\left[\frac{d_p^2\ V^2}{\mu}\right] \qquad (5.28)$$

The collection efficiency or precipitation efficiency of a electrostatic precipitation depends upon the time for which the gas remains in active electric field. Collection efficiency is usually from 90-99% due to economic reason. The collection efficiency in terms as:

$$\ln(1\text{-}\eta) = \left[\frac{R_2}{w.t}\right] \qquad (5.29)$$

Where, t- is time for which gas remains in electric field
w- velocity of particle collection.

8.2 Two stage unit

In two stage precipitator charging takes place in one section which is followed by a section consisting of alternately charged plates. The collecting electric field is established independently of the corona field and such precipitator are termed two stage. In this unit the ground plates are about an inch apart and have an intermediate plate that is also charged. But instead of 40,000-60,000 volt DC supply, the two stage precipitator has a 13,000-15,000 volt supply with the intermediate supply at 7,500 volts. This type of collector usually developed for heating and ventilating service and it provides very efficient dust collection and is designed with self cleaning washing system. The washing system is a light duty until designed for 250 cycles. As the usual cleaning is only required monthly, this unit exceeds the life of other components of heating and ventilating system. The high voltage electrodes consist of very fine wire stretched across springs and it is required for ionization. The plates have to maintain it at more precise distances. In this kind of service, the air distribution is usually very even, since the dust collecting filtering device operated at the same velocities as the heating and cooling coils.

9. Hydrocyclone

Hydrocyclones are used to carry out sedimentation of particles in a liquid medium under enhanced separating force. It is a device in which the necessary force is generated by rotating the feed slurry rapidly in the cylindrical conical section. The cylindrical conical section is provided at the top and bottom of the cyclone so that fine particles and coarse particles can be withdrawn continuously. Because of the operational similarities hydrocyclone is confused with a centrifuge. Hydrocyclones are simple, robust separating device which can separate both coarse and fine particles in a range of 4- 500 µm. While centrifuges are normally used for particle sizes of 100 µm or finer particles. Centrifuges require much higher centrifugal force than hydro cyclone. Centrifuges are more expensive and require higher operating cost. Schematic of hydro cyclone and its flow pattern is shown in (figure 5.1)

In a hydrocyclone, feed slurry enters at the top part of the cylindrical body, the entry can be tangential or involute type. As soon as the slurry starts to rotate in the cylindrical body, as a result coarse or heavy particles are thrown outwards towards the wall that move downward in a spiral path towards the underflow or an open orifice where they are ejected. During the process fine particles remain near the center of the cyclone, close to an air core which is always present, and eventually move in an upward spiral towards the overflow orifice. The orifice consists of a cylindrical section called as vortex finder, its lower end extends below the level of the feed port. The separation in a hydrocyclone is made on the basis of settling velocities, particle size, density and particle shape. If the slurry contains a mixture of solid particles of different densities, it is possible that large light particles may settle along with the small heavy particles. A specific gravity difference must exist between the solid and liquid phase in order that a separation can take place.

Size of cyclone and the estimation of the number of stages required for a particular application depend upon the stream flow rate. Selection of cyclone should be based on the recovery of a particle of specific size. The limit is decided based on the size for example, d_{50} particle diameter is the diameter of the particle, 50 % of which will appear in the overflow and the 50% in the under flow. It is referred as cut size and denoted as d_{50c}. Several correlations are available for determination of d_{50c}.

$$d_{50c} = 14.2 \frac{D^{0.46}\ d_i^{0.6}\ d_0^{0.21}\ \exp(0.063\ v)}{d_u^{0.71}\ h^{0.38}\ Q^{0.45}\ \left(\rho_s - \rho_f\right)^{0.5}} \qquad (5.30)$$

Q – slurry feed rate, m^3/hr; d_i – internal diameter of the inlet, cm; d_0 – internal diameter of overflow, cm; d_u – internal diameter of underflow, cm; D- diameter of cyclone, cm; v- volume fraction of solids in feed; h- distance between bottom of vortex finder to the top of underflow orifice, cm; ρ_s: ρ_f – density of solids and fluid respectively, kg/m^3

Inlet orifice area should be 6-8% of the cross sectional area of feed chamber, vortex finder diameter is 35-40% of cyclone diameter, cylindrical section length is equal to the diameter of feed chamber, angle of conical section is equal to 12^0 for cyclone diameter less than 300 mm and 20^0 for cyclone diameter greater than 300 mm, underflow orifice diameter is equal to 25% of vortex finder diameter.
If the hydro cyclone of above geometry has to be designed, then d_{50} can be estimated as:

$$d_{50} = \frac{13.2\ D^{0.675} \exp\left[-0.3 + 0.095\ v - 0.036\ v^2 + 0.000065\ v^3\right]}{\Delta P^{0.3}\left(\rho_s - \rho_f\right)} \quad (5.31)$$

ΔP – pressure drop across the cyclone (kPa)

The relationship between Q and ΔP is: $Q = 9.4 \times 10^{-3}\ \Delta P^{0.5}\ D^2$. The D_{50} base point can be estimated as: $D_{50b} = 2.86\ D^{0.66}$, where D – cyclone diameter, cm; D_{50b} – base diameter, µm.
The above equation is simple and indicates that D_{50b} increases as the cyclone diameter increases. The equation is applicable to standard hydrocyclone geometry. The standard configuration can be stated as: cyclone of diameter, D; inlet slurry diameter, 0.5D; vortex finder diameter d_0, 0.35 D; cylindrical section length, 0.35D; cone angle of $10\text{-}20^0$; apex orifice, (0.1-0.35) D. All these standard operating conditions are for water as a fluid, particles (quartz) of specific gravity 2.65 feed solid concentration < 1% v/v and pressure drop ΔP = 69 kPa. In case if the above condition cannot be satisfied then the correction factors have to be used.

Correction factor for feed concentration	Correction factor for pressure drop	Correction factor for specific gravity of solids
$C_1 = \left[\frac{53 - V^{-0.43}}{53}\right]$ V- volume % of solids in feed stream	$C_2 = 3.27\ \Delta P^{-0.28}$ ΔP – pressure drop, kPa	$C_3 = \left(\frac{1.65}{\rho_5 - \rho_f}\right)^{0.5}$

The final cut size, $D_{50} = D_{50b} \times C_1 \times C_2 \times C_3$
While designing the hydro cyclone, initially the dimensions of inlet, overflow and discharge openings can be assumed and pressure drop fir this configuration is evaluated and the procedure can be continued unless the minimum pressure drop value is achieved.
The permissible pressure drop value for in case of hydro cyclone is between 0.2-0.7 kg/cm^2
Capacity in case of 38^0 hydro cyclone can be estimated as:

$$q = K\ D\ d_u \left(\frac{\Delta P}{\rho_s}\right)^{0.5}$$

Where, q – slurry feed rate, lit/min; D- diameter of cyclone, cm; d_u-internal diameter of underflow, cm; ΔP- pressure drop between inlet and over flow outlet, kg/cm^2; ρ_s – density of slurry, kg/cm^3.

K – 0.51 for D = 80 to 100 mm

K – 0.524 for D = 125 to 600 mm

The ratio of $d_u/d_0 = 0.4$; d_i – diameter of inlet, d_0 – diameter of outlet

$(2d_i + d_0) = (D/2)$; $d_0/d_i = (1.0 - 2.0)$

Optimum diameter of hydro cyclone D is 5-8 times the internal diameter of inlet.

Length of conical section can be estimate as $= \left[\dfrac{D - d_u}{2 \tan(\alpha/2)} \right]$

The hydro cyclone performance totally depends upon solid contents of feed, feed flow rate and the pressure at which it enters the hydro cyclone. Any deviation in these parameters can lead to erratic performance.

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Exercise problem

1. Estimate the settling velocity for the hindered settling of glass spheres in water 68^oF when the suspension contains 1200 g of glass spheres in 1100 cu cm of total volume. The average diameter of the spheres, as determined from the photomicrographs, was 0.006 in., and the true density of the spheres was 154 lb mass/cu ft.

   (*Ans:* u_g = 0.061 fps; ϵ = 0.55; settling velocity of suspension, u = 0.0030 fps; R_e = 0.14)