16 Baghouse (BAG)

BAG.1 DESCRIPTION OF FILTRATION

Describe the filtration process.

Solution

The basic filtration process may be conducted with many different types of fabric filters in which the physical arrangement of hardware and the method of removing collected material from the filter media will vary. The essential differences may be related, in general, to

1. Type of fabric
2. Cleaning mechanism
3. Equipment geometry
4. Mode of operation

Baghouse collectors are available for either intermittent or continuous operation. Intermittent operation is employed where the operational schedule of the dust-generating source permits halting the gas cleaning function at periodic intervals (regularly defined by time or by pressure differential) for removal of collected material from the filter media (cleaning). Collectors of this type are primarily utilized for the control of small-volume operations such as grinding, polishing, etc., and for aerosols of a very coarse nature. For most air pollution control installations and major dust control problems, however, it is desirable to use collectors that allow for continuous operation. This is accomplished by arranging several filter areas in a parallel-flow system and cleaning one area at a time according to some preset mode of operation.

BAG.2 CLEANING METHODS

Discuss the various cleaning methods employed in baghouses.
Solution

Baghouses may be characterized and identified according to the method used to remove collected material from the bags. Particle removal can be accomplished in a variety of ways, including shaking the bags, reversing the flow of gas through the bags, or rapidly expanding the bags by a pulse of compressed air. In general, the various types of bag cleaning methods can be divided into those involving fabric flexing and those involving a reverse flow of clean air.

Cleaning by mechanical shaking is accomplished by isolating one of several bag compartments from the air flow and vigorously shaking the bags for about a minute to dislodge the dust. For simplicity of operation, the bags are usually attached to a motor-driven oscillating carriage. Because of tensile stresses produced by this approach, strong fabric material must be used.

Reverse-flow baghouses employ an auxiliary fan that forces air through the bags in a direction opposite to that of filtration. This procedure collapses the bag and fractures the dust cake. The reverse-flow rate, which is ordinarily about the same as the face velocity during filtering, deflates the bag and helps to dislodge the dust cake from the fabric surface. It is common practice to combine shaking and reverse-flow cleaning in the same unit.

Reverse-pulse cleaning is the newest and perhaps the most efficient method to clean bags, given the proper working environment. A short pulse (about 0.1 s) of compressed air (approximately 100 psia) is injected into each bag through a venturi, causing the bag to expand while creating intense dust separating forces.

BAG.3 OVERALL COLLECTION EFFICIENCY

Calculate the overall efficiency of $N$ compartments in a baghouse operated in parallel, if the volumetric flowrates and inlet concentrations to each compartment are $q_1, q_2, \ldots, q_N$ and $c_1, c_2, \ldots, c_N$, respectively, and the corresponding efficiencies are $E_1, E_2, \ldots, E_N$. Also, express the result in terms of the $q$'s, the $E$'s, and the $c$'s.

Calculate the overall efficiency of a baghouse consisting of three compartments treating 9000 acfm of gas with an inlet loading of $4 \text{ gr/ft}^3$. The first and third compartments operate at a fractional efficiency of 0.995 while the second compartment operates at a fractional efficiency of 0.990. What is the overall efficiency of the baghouse if the flow and inlet concentration are evenly distributed? Also calculate the efficiency if the following flow distribution exists:

<table>
<thead>
<tr>
<th>Compartment</th>
<th>$q$ (acfm)</th>
<th>$c$ (gr/ft$^3$)</th>
<th>$E$</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>2500</td>
<td>3.8</td>
<td>0.995</td>
</tr>
<tr>
<td>2</td>
<td>4000</td>
<td>4.25</td>
<td>0.990</td>
</tr>
<tr>
<td>3</td>
<td>2500</td>
<td>3.8</td>
<td>0.995</td>
</tr>
</tbody>
</table>
Solution

Write the equation for the outlet concentration, \( c_{1o} \), from module 1 in terms of \( c_1 \) and \( E_1 \):

\[
E_1 = 1 - \left( \frac{c_{1o}}{c_1} \right) \\
c_{1o} = c_1 (1 - E_1)
\]

The equation for the inlet mass flowrate to module, \( \dot{m}_1 \), is

\[
\dot{m}_1 = c_1 q_1
\]

The equation for the outlet mass flowrate, \( \dot{m}_{1o} \), from module 1 is

\[
\dot{m}_{1o} = c_1 (1 - E_1) q_1
\]

For module \( i \),

\[
E_i = 1 - \left( \frac{c_{io}}{c_i} \right) \\
c_{io} = c_i (1 - E_i) \\
\dot{m}_i = c_i q_i \\
\dot{m}_{io} = c_i (1 - E_i) q_i
\]

The equation for the overall efficiency, \( E \), for modules 1, 2, \ldots, \( N \) is then

\[
E = 1 - \frac{\sum \dot{m}_{1o}}{\sum \dot{m}_i} = 1 - \frac{c_1 (1 - E_1) q_1 + c_2 (1 - E_2) q_2 + \cdots + c_N (1 - E_N) q_N}{c_1 q_1 + c_2 q_2 + \cdots + c_N q_N}
\]

\[
E = 1 - \frac{\sum c_i (1 - E_i) q_i}{\sum c_i q_i}
\]

The companion equation for the penetration \( P \) is

\[
P = \frac{c_1 P_1 q_1 + c_2 P_2 q_2 + \cdots + c_N P_{Nq_N}}{c_1 q_1 + c_2 q_2 + \cdots + c_N q_N}
\]

\[
= \frac{\sum c_i P_i q_i}{\sum c_i q_i}
\]
If the inlet concentrations $c$ to each module are equal, i.e.,

$$c_1 = c_2 = \cdots = c_N = c$$

the $c$ terms can be factored out from the above equation for efficiency to yield

$$E = 1 - \frac{(1 - E_1)q_1 + (1 - E_2)q_2 + \cdots + (1 - E_N)q_N}{q_1 + q_2 + \cdots + q_N}$$

Note that the total volumetric flowrate $q$ is given by

$$q = q_1 + q_2 + \cdots + q_N$$

Therefore,

$$E = 1 - \frac{(1 - E_1)q_1 + (1 - E_2)q_2 + \cdots + (1 - E_N)q_N}{q} = 1 - \frac{\sum (1 - E_i)q_i}{q}$$

Equivalently,

$$P = 1 - \frac{\sum q_iP_i}{q}$$

Using the data provided in the problem statement, calculate the efficiency for the situation where the flow is equally distributed with the same inlet loading. Since $q_i = 3000$ acfm for all modules,

$$E = 1 - \frac{(2)(3000)(1 - 0.995) + (1)(3000)(1 - 0.99)}{9000} = 0.9933 = 99.33\%$$

If neither the flow nor concentration are uniformly distributed, the general equation for the efficiency in a compartmentalized baghouse is used.

$$E = 1 - \frac{\sum c_i(1 - E_i)q_i}{\sum c_iq_i}$$

Substituting (see data),

$$E = 1 - \frac{(2)(3.8)(2500)(1 - 0.995) + (1)(4.25)(4000)(1 - 0.99)}{36,000} = 0.9926 = 99.26\%$$
It can be seen that poor flow distribution causes the efficiency to decrease. With much larger systems, this effect can end up being significant and may cause the baghouse to be out of compliance.

**BAG.4 BAGHOUSE COLLECTION**

The dimensions of a bag in a filter unit are 8 in. in diameter and 15 ft long. Calculate the filtering area of the bag. If the filtering unit consists of 40 such bags and is to treat 480,000 ft$^3$/h of gas from an open-hearth furnace, calculate the "effective" filtration velocity in feet per minute and acfm per square foot of filter area. Also calculate the mass of particles collected daily if the inlet loading is 3.1 gr/ft$^3$ and the unit operates at 99.99+\% collection efficiency.

**Solution**

Assume the bag to be cylindrical in shape with diameter $D$ and height $h$. The total area of the bag is

$$A = A_{\text{curved surface}} + A_{\text{flat top}}$$

$$= \pi Dh + \pi D^2/4$$

$$= \pi(\frac{8}{12})(15) + \pi(\frac{8}{12})^2/4$$

$$= 31.77 \text{ ft}^2$$

The total area for 40 bags is

$$A = (40)(31.77) = 1271 \text{ ft}^2$$

The filter velocity is then

$$v = \frac{q_G}{A} = \frac{(480,000/60)}{1271}$$

$$= 6.30 \text{ ft/min}$$

The calculation for acfm per square foot of filter area is the same.

Assuming 100\% collection efficiency, the mass collected daily is

$$\text{Mass collected} = q_Gc_i = (480,000)(24)(3.1)/7000$$

$$= 5102 \text{ lb/day}$$

Note that 7000 gr = 1 lb.
BAG.5 ADVANTAGES AND DISADVANTAGES OF BAGHOUSES

Discuss the advantages and disadvantages associated with a baghouse.

Solution

Some of the advantages and disadvantages associated with employing a baghouse for particulate control are listed below.

Advantages include:

1. Moderate capital cost
2. Moderate operating cost
3. Extremely high collection efficiencies
4. Dry collection
5. No resistivity problems

Disadvantages include:

1. Space requirements
2. Bag failure (to be discussed later)
3. Explosion hazards
4. Temperature dependence
5. Bag replacement

BAG.6 NUMBER OF BAGS, PRESSURE DROP, AND CLEANING FREQUENCY

A calcium hydroxide plant is required to treat the exhaust "fume" generated from the plant. The ash generated from the system is collected at the bottom of the baghouse while the exhaust gas flow of 350,000 acfm enters the baghouse with a loading of 6.0 gr/ft$^3$. The air-to-cloth ratio is 8.0 and the operating particulate collection efficiency is 99.3%. The maximum allowable pressure drop is 10 in. H$_2$O. The contractor's empirical equation for the pressure drop is given by

$$\Delta P = 0.3v + 4.0cv^2t$$

where $\Delta P =$ pressure drop in inches of water
$v =$ filtration velocity in ft/min
$c =$ dust concentration in lb/ft$^3$ of gas
$t =$ time in minutes since bags were cleaned
1. How many cylindrical bags, 12 in. in diameter and 30 ft high will be needed?

2. The system is designed to begin cleaning when the pressure drop reaches 10.0 in. H₂O, its maximum allowable value. How frequently should the bags be cleaned?

As particles are collected in the baghouse, the pressure drop across the fabric filtering media increases. Due in part to fan limitations, the filter must be cleaned at predetermined intervals. Dust is removed from the fabric by gravity and/or mechanical means. The fabric filters or bags are usually tubular or flat. As described earlier, the structure in which the bags hang is frequently referred to as a baghouse. The number of bags in a baghouse may vary from a few to several thousand. Quite often when great numbers of bags are involved, the baghouse is compartmentalized so that one compartment may be cleaned while others are still in service.

There are several equations provided in the literature to describe pressure drop in a fabric filter system. The form most often used is

\[ \Delta P = \Delta P_{\text{fabric}} + \Delta P_{\text{cake}} = K_1 v + K_2 c^2 t \]

where \( \Delta P \) = total pressure drop across both fabric and cake (in. H₂O)

\( v \) = superficial velocity through the bag/cake (ft/min)

\( c \) = inlet particulate loading in gas (lb/ft³)

\( t \) = elapsed time in filtering cycle (min)

\( K_1 \) = resistance coefficient for the bag (fabric) (in. H₂O/[ft/min])

\( K_2 \) = resistance coefficient for the deposited dust (cake) (in. H₂O/[(lb/ft³)² min])

The size of a baghouse is primarily determined by the area of filter cloth required to filter the gases. The choice of a filtration velocity (or its equivalent, the air-to-cloth ratio (ACR or A/C) in actual cubic feet per minute of gas filtered per square foot of filter area) must take certain factors into consideration. Although the higher velocities are usually associated with the greater pressure drops, they also reduce the filter area required. Practical experience has led to the use of a series of ACRs for various materials collected and types of equipment. Ratios in current use range from < 1:1 to > 15:1. The choice depends on cleaning method, fabric, and characteristics of the particles.

For the same cleaning efficiency, felted fabrics in pulse-jet baghouses are often capable of higher air-to-cloth ratios than woven fabrics in reverse-air baghouses, thereby requiring less filter cloth area and, consequently, less space for a given air or gas volume. Woven fabrics in reverse-air baghouses usually have ACRs of 1:1 to 5:1; felted fabrics in pulse-jet baghouses usually have ratios of 3:1 to 15:1, or ratios several times those of woven fabrics. This is balanced, though, by the higher cost of the felt fabrics.
Solution

Calculate the total required surface area $A$ of the bags if the air-to-cloth ratio is 8.0:

$$A = \frac{\text{Volumetric gas flowrate}}{\text{Filtration velocity}}$$

$$= \frac{350,000}{8}$$

$$= 43,750 \text{ ft}^2$$

Calculate the surface area of each bag, $a$, and the number of the bags required, $N$:

$$a = \pi DL + \pi D^2/4$$

$$= \pi(12/12)(30) + \pi(12/12)^2/4$$

$$= 95 \text{ ft}^2$$

$$N = \frac{A}{a}$$

$$= \frac{43,750}{50}$$

$$= 461 \text{ bags}$$

Solve the pressure drop equation explicitly for the time:

$$\Delta P = 0.3v + 4cv^2 t$$

$$t = \frac{\Delta P - 0.3v}{4cv^2}$$

The concentration $c$ is given by

$$c = \frac{6.0}{7000}$$

$$= 8.57 \times 10^{-4} \text{ lb}/\text{ft}^3$$

Solving for the time yields

$$t = \frac{10 - 0.3(8)}{(4)(8.57 \times 10^{-4})(8)}$$

$$= 34.6 \text{ min}$$

BAG.7 BAG FAILURE

A baghouse has been used to clean a particulate gas stream for nearly 30 years. There are 600 8-in. diameter bags in the unit; 50,000 acfm of dirty gas at 250°F
enters the baghouse with a loading of 5.0 gr/ft$^3$. The outlet loading is 0.03 gr/ft$^3$. Local Environmental Protection Agency (EPA) regulations state that the outlet loading should not exceed 0.4 gr/ft$^3$. If the system operates at a pressure drop of 6 in. of water, how many bags can fail before the unit is out of compliance? The Theodore–Reynolds equation applies and all the contaminated gas emitted through the broken bags may be assumed the same as that passing through the tube sheet thimble.

The effect of bag failure on baghouse efficiency can be described by the following equations:

\[
P_t^* = P_t + P_{tc}
\]

\[
P_{tc} = \frac{0.528(\Delta P)^{0.5}}{\phi}
\]

\[
\phi = \frac{q}{LD^2(T + 460)^{0.5}}
\]

where \(P_t^*\) = penetration after bag failure

\(P_t\) = penetration before bag failure

\(P_{tc}\) = penetration correction term; contribution of broken bags to \(P_t^*\)

\(\Delta P\) = pressure drop, in. H$_2$O

\(\phi\) = dimensional parameter

\(q\) = volumetric flowrate of contaminated gas, acfm

\(L\) = number of broken bags

\(D\) = bag diameter, in.

\(T\) = temperature, °F

For a detailed development of the above equation, refer to “Effect of Bag Failure on Baghouse Outlet Loading,” Theodore and Reynolds, JAPCA, August 1979, 870–872.

**Solution**

Calculate the efficiency \(E\) and penetration \(P_t\) before the bag failure(s):

\[
E = \frac{\text{Inlet loading} - \text{Outlet loading}}{\text{Inlet loading}}
\]

\[
= \frac{(5.0 - 0.03)}{5.0}
\]

\[
= 0.9940 = 99.40\%
\]

\[
P_t = 1 - 0.9940
\]

\[
= 0.0060 = 0.60\%
\]
The efficiency and penetration, \( P_t^* \), based on regulatory conditions are

\[
E = \frac{5.0 - 0.4}{5.0} = 0.9200 = 92.00\%
\]

\[
P_t^* = 1 - 0.9200 = 0.0800 = 8.00\%
\]

The penetration term, \( P_{tc} \), associated with the failed bags is then

\[
P_{tc} = 0.0800 - 0.0060 = 0.0740
\]

Write the equation(s) for \( P_{tc} \) in terms of the failed number of bags, \( L \). Since

\[
P_{tc} = \frac{0.528(\Delta P)^{0.5}}{\phi}
\]

and

\[
\phi = \frac{q}{LD^2(T + 460)^{0.5}}
\]

then,

\[
L = \frac{qP_{tc}}{(0.582)\Delta P^{0.5}D^2(T + 460)^{0.5}}
\]

The number of bag failures that the system can tolerate and still remain in compliance is now calculated:

\[
L = \frac{(50,000)(0.074)}{(0.582)(6)^{0.5}(8)^2(250 + 460)^{0.5}} = 1.52
\]

Thus, if two bags fail, the baghouse is out of compliance.

It is important to note that each bag in a set may have a different life as a result of fabric quality, bag manufacturing tolerances, location in the collector, and variation in the bag cleaning mechanism. Any one or a combination of these factors can cause bags to fail. This means that a baghouse will experience a series of intermittent bag failures until the failure rate requires total bag replacement. Typically, a few bags will fail initially or after a short period of operation due to installation damage or manufacturing defects. The failure rate should then remain very low until the
operating life of the bags is approached, unless a unique failure mode is present within the system. The failure then increases, normally at a near exponential rate. Industry often describes this type of failure rate behavior as a “bathtub” curve. The reader is referred to the Theodore tutorial titled Accident and Emergency Management, ETS International, Roanoke, VA, 1998 for detailed information on system accidents/failures.

The proper time to replace a broken bag depends on the type of collector and the resultant effect on outlet emissions. In “inside bag collection” types of collectors, it is very important that dust leaks be stopped as quickly as possible to prevent adjacent bags from being abraded by jet streams of dust emitted from the broken bag. This is called the “domino effect” of bag failure. “Outside bag collection” systems do not have this problem, and the speed of repair is determined by whether the outlet opacity has exceeded its limits. Often, it will take several broken bags to create an opacity problem, and a convenient maintenance schedule can be employed instead of emergency maintenance.

In either type of collector, the location of the broken bag or bags has to be determined and corrective action taken. In a noncompartmentalized unit, this requires system shutdown and visual inspection. In inside collectors, bags often fail close to the bottom, near the tube sheet. Accumulation of dust on the tube sheets, the holes themselves, or unusual dust patterns on the outside of the bags often occurs. Other probable bag failure locations in reverse-air bags are near anticollapse rings or below the top cuff. In shaker bags, one should inspect the area below the top attachment. Improper tensioning can also cause early failure.

In outside collectors, which are normally top-access systems, inspection of the bag itself is difficult; however, location of the broken bag or bags can normally be found by looking for dust accumulation on top of the tube sheet, on the underside of the top-access door, or on a blow pipe.

**BAG.8 FREQUENCY OF BAG FAILURE**

As a recently assigned plant engineer, you are asked to troubleshoot the plant’s baghouse. The baghouse is used to collect the dust created in the manufacture of an extremely expensive drug. The dust is collected and recycled into the main process. Over the past 6 months (since the baghouse was installed) the amount of dust collected has dropped off significantly without any change in the inlet loading. Since the baghouse is operated on a round-the-clock basis, i.e., 24 h per day, 7 days per week, the bags (for this unit) cannot be inspected to find the problem. The following data have been collected:

- Flowrate = 60,000 acfm (60°F)
- Dust loading = 6.00 gr/ft³
- Number of bags = 500
- Diameter of bags = 5.0 in.
Pressure drop = 9.1 in. H$_2$O

<table>
<thead>
<tr>
<th>Months of Operation</th>
<th>Amount Collected (lb/h)</th>
</tr>
</thead>
<tbody>
<tr>
<td>New</td>
<td>3054.9</td>
</tr>
<tr>
<td>1</td>
<td>2900.4</td>
</tr>
<tr>
<td>2</td>
<td>2808</td>
</tr>
<tr>
<td>3</td>
<td>2653.8</td>
</tr>
<tr>
<td>4</td>
<td>2530.2</td>
</tr>
<tr>
<td>5</td>
<td>2376</td>
</tr>
<tr>
<td>6</td>
<td>1789.8</td>
</tr>
</tbody>
</table>

Having recently attended a class on the effects of bag failures given by the foremost authority in this field, you are asked to determine if the loss has been caused by broken bags, and, if so, how many have broken every month.

**Solution**

First calculate the inlet loading (IL):

$$IL = (60,000)(6.0)(60)/(7000) = 3085.7 \text{ lb/h}$$

The calculations of the initial efficiency $E$ and penetration $P$ follow:

$$E = 3054.9/3085.7 = 0.99 = 99\%$$

$$P - 1 - 0.99 = 0.01 = 1.0\%$$

After the first month,

$$E^* = 2900/3085.7 = 0.94 = 94\%$$

$$P^* = 1 - 0.94 = 0.06 = 6\%$$

(* indicates that time has passed since baghouse installation).

The amount of increase in the penetration $P_c$ is

$$P_c = 0.06 - 0.01 = 0.05$$

Using the equation (See Problem BAG.7),

$$\phi = 35.1 = \frac{q}{Ld^2(t + 460)^{1/2}}$$

$$L = \frac{q}{\phi D^2(t + 460)^{1/2}} = \frac{60,000}{(35.1)(5)^2(520)^{1/2}}$$

$$= 2.998 \approx 3$$
Similarly,

<table>
<thead>
<tr>
<th>Month</th>
<th>$P^*$ (%)</th>
<th>$P_c$ (%)</th>
<th>$\phi$</th>
<th>$L$</th>
<th>$\Delta L$</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>6</td>
<td>5</td>
<td>35.1</td>
<td>3</td>
<td>3</td>
</tr>
<tr>
<td>2</td>
<td>9</td>
<td>8</td>
<td>21</td>
<td>5</td>
<td>2</td>
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<td>7</td>
<td>2</td>
</tr>
<tr>
<td>4</td>
<td>18</td>
<td>17</td>
<td>10.5</td>
<td>10</td>
<td>3</td>
</tr>
<tr>
<td>5</td>
<td>23</td>
<td>22</td>
<td>7.5</td>
<td>14</td>
<td>4</td>
</tr>
<tr>
<td>6</td>
<td>42</td>
<td>41</td>
<td>4.2</td>
<td>25</td>
<td>11</td>
</tr>
</tbody>
</table>

The baghouse problem appears to be bag failure. Note that the efficiency drops below 90% after just 2 months. The reader should consider whether the bag failure distribution with time is reasonable.

**BAG.9 COLLECTION EFFICIENCY MODEL**

Consider the situation where 50,000 acfm of gas with a dust loading of 5.0 grain/ft$^3$ flows through a baghouse with an average filtration velocity of 10 ft/min. The pressure drop is given by

$$\Delta P = 0.20v + 5.0 c_i v^2 t$$

where $\Delta P =$ pressure drop, in. of H$_2$O  
$v =$ filtration velocity, ft/min  
$c_i =$ dust concentration, lb/ft$^3$ of gas  
$t =$ time after bags were cleaned, min

The fan can maintain the volumetric flowrate up to a pressure drop of 5.0 in. of water. Show that the baghouse can be operated for 8.40 min between cleanings.

In an attempt to determine the efficiency of this unit at "terminal" conditions, both the fabric and deposited cake (individually) were subjected to laboratory experimentation. The following data were recorded:

Fabric alone: thickness = 0.1 in.

1 cloth

Concentration at 1 = 1.0 g/cm$^3$

at 2 = 0.1245 g/cm$^3$
Using the Theodore and Reynolds collection efficiency model, determine the overall efficiency at the start and end of a cleaning cycle.

The number of variables necessary to design a fabric filter is very large. Since fundamentals cannot treat all of these factors in the design and/or prediction of performance of a filter, this determination is basically left up to the experience and judgment of the design engineer. In addition, there is no one formula that can determine whether or not a fabric filter application is feasible. A qualitative description of the filtration process is possible, although quantitatively the theories are far less successful. Theory, coupled with some experimental data, can help predict the performance and design of the unit.

As discussed previously, the state of the art of engineering process design for baghouses is the selection of filter medium, superficial velocity, and cleaning method that will yield the best economic compromise. Industry relies on certain simple guidelines and calculations, which are usually considered proprietary information, to achieve this. Despite the progress in developing pure filtration theory, and in view of the complexity of the phenomena, the most common methods of correlation are based on predicting a form of a final equation that can be verified by experiment. For example, an equation that can be used for determining the collection efficiency of a baghouse is (Introduction to Hazardous Waste Incineration by Santoleri, Reynolds, and Theodore, Wiley-Interscience, 2000):

$$E = 1 - e^{-(\psi L + \phi t)}$$

where
- $\psi$ = constant (determined by experiment) based on the fabric ($\text{ft}^{-1}$, or units consistent with $L$)
- $\phi$ = constant (determined by experiment) based on the cake ($s^{-1}$, or units consistent with $t$)
- $t$ = time of operation to develop the cake thickness ($s$)
- $L$ = fabric thickness ($\text{ft}$)
- $E$ = collection efficiency (dimensionless)

### Solution

Solve the equation for the pressure drop explicitly for $t$:

$$t = \frac{\Delta P - 0.2v}{5c_{ij}v^2} = \frac{5.0 - 0.2(10)}{5(5/7000)(10)^2}$$

$$= 8.4 \text{ min}$$

Cake alone: time $= 3.4043 \text{ min}$ (end of cleaning cycle)
Evaluate the parameter $\psi$ with units of $\text{ft}^{-1}$ and the parameter $\phi$ with units of $\text{min}^{-1}$:

\[
\ln\left(\frac{c_o}{c_i}\right) = -\psi L \\
\ln(0.1245/1.0) = -\psi(0.1/12) \\
\psi = 250 \text{ ft}^{-1} \\
\ln\left(\frac{c_o}{c_i}\right) = -\phi t \\
\ln(0.0778/1.0) = -\phi(8.4) \\
\phi = 0.304 \text{ min}^{-1}
\]

Calculate the efficiency at the end of the filtering (cleaning) cycle.

\[
E = 1 - e^{-\psi L} e^{-\phi t} \\
= 1 - e^{-250(0.1/12)} e^{-0.304(8.4)} \\
= 0.9903 \\
= 99.03\% \text{ (at end of cleaning cycle)}
\]

Also calculate the efficiency at the start of the filtering (cleaning) cycle:

\[
E = 1 - (0.1245)(1) \quad t = 0 \\
= 0.8755 \\
= 87.55\% \text{ (at start of cycle)}
\]

The reader should note that $\phi$ and $\psi$ are modified inertial impaction numbers based on the cake and fabric, respectively. Unfortunately, this key equation has been totally ignored by responsible EPA individuals in this field. Taxpayers’ dollars continue (for over a dozen years) to be provided to contractors whose research efforts have produced little, if any, usable results in developing quantitative equations to describe collection efficiency.

Using the above model, one can show that the exit concentration ($w_e$) for the combined resistance system (the fiber and the cake) is

\[
w_e = w_i e^{-(\psi L + \phi t)}
\]

where $w_e =$ exit concentration; units consistent with $w_i$  
$w_i =$ inlet concentration
BAG.10 FILTER BAG FABRIC SELECTION

It is proposed to install a pulse-jet fabric filter system to clean an airstream containing particulate pollutants. You are asked to select the most appropriate filter bag fabric considering performance and cost. Pertinent design and operating data, as well as fabric information, are given below.

Volumetric flowrate of polluted airstream = 10,000 scfm (60°F, 1 atm)
Operating temperature = 250°F
Concentration of pollutants = 4.00 gr/ft^3
Average ACR = 2.5 cfm/ft^2 cloth
Collection efficiency requirement = 99%

<table>
<thead>
<tr>
<th>Filter Bag</th>
<th>A</th>
<th>B</th>
<th>C</th>
<th>D</th>
</tr>
</thead>
<tbody>
<tr>
<td>Tensile strength</td>
<td>Excellent</td>
<td>Above average</td>
<td>Fair</td>
<td>Excellent</td>
</tr>
<tr>
<td>Recommended maximum</td>
<td>260</td>
<td>275</td>
<td>260</td>
<td>220</td>
</tr>
<tr>
<td>temperature (°F)</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Resistance factor</td>
<td>0.9</td>
<td>1.0</td>
<td>0.5</td>
<td>0.9</td>
</tr>
<tr>
<td>Cost per bag, ($)</td>
<td>26</td>
<td>38</td>
<td>10</td>
<td>20</td>
</tr>
<tr>
<td>Standard size</td>
<td>8 in. x 16 ft</td>
<td>10 in. x 16 ft</td>
<td>1 ft x 16 ft</td>
<td>1 ft x 20 ft</td>
</tr>
</tbody>
</table>

Note: No bag has an advantage from the standpoint of durability under the operating conditions for which the bag was designed.

Solution

A wide variety of woven and felted fabrics are used in fabric filters. Clean felted fabrics are more efficient dust collectors than woven fabrics, but woven materials are capable of giving equal filtration efficiency after a dust layer accumulates on the surface. When a new woven fabric is placed in service, visible penetration of dust may occur until buildup of the cake or dust layer. This normally takes from a few hours to a few days for industrial applications, depending on dust loadings and the nature of the particles.

When using woven fabrics, care must be exercised to prevent overcleaning so as not to completely dislodge the filter cake; otherwise, efficiency will drop. Overcleaning of felted fabrics is generally impossible because they always retain substantial dust deposits within the fabric. Felted fabrics require more thorough cleaning methods than woven materials. If felted fabrics are used, filter cleaning is limited to the reverse-pulse method. When woven fabrics are employed, any cleaning technique may be used. Woven fabrics are available in a greater range of temperature and corrosion-resistant materials than felts and, therefore, cover a wider range of applications.
Bag D is eliminated since its recommended maximum temperature (220°F) is below the operating temperature of 250°F. Bag C is also eliminated since a pulse-jet fabric filter system requires the tensile strength of the bag to be at least above average.

Consider the economics for the two remaining choices. The cost per bag is $26.00 for A and $38.00 for B. The gas flowrate and filtration velocity are

\[ q_a = 10,000 \left( \frac{250 + 460}{60 + 460} \right) \]

\[ = 13,654 \text{ acfm} \]

\[ v_f = 2.5 \text{ cfm/ft}^2 \text{ cloth} \]

\[ = 2.5 \text{ ft/min} \]

The filtering (bag) area is then

\[ A_c = \frac{q}{v_f} \]

\[ = \frac{13,654}{2.5} \]

\[ = 5,462 \text{ ft}^2 \]

For bag A, the area and number, \( N \), of bags are

\[ A = \pi Dh \]

\[ = \pi \left( \frac{8}{12} \right)(16) \]

\[ = 33.5 \text{ ft}^2 \]

\[ N = \frac{A_c}{A} \]

\[ = \frac{5,462}{33.5} \]

\[ = 163 \]

For bag B:

\[ A = \pi \left( \frac{10}{12} \right)(16) \]

\[ = 41.9 \text{ ft}^2 \]

\[ N = \frac{5462}{41.9} \]

\[ = 130 \]
The total cost (TC) for each bag is as follows:

For bag A:

\[ TC = N \text{ (cost per bag)} \]
\[ = (163)(26.00) \]
\[ = 4238 \]

For bag B:

\[ TC = (130)(38.00) \]
\[ = 4940 \]

Since the total cost for bag A is less than bag B, select bag A.
VENTURI SCRUBBERS (VEN)

VEN.1 POWER REQUIREMENTS FOR A VENTURI SCRUBBER

Calculate the power requirement of a venturi scrubber treating 380,000 acfm of gas and operating at a pressure drop of 60 in. H₂O.

Solution

The gas horsepower may be calculated from

\[ HP = \frac{q \Delta P}{6356} \]

\[ q = \text{acfm} \quad \Delta P = \text{in. H₂O} \]

\[ = \frac{(380,000)(60)}{6356} \]

\[ = 3587 \text{hp} \]

Alternately, the following equation may be used:

\[ P_G = 0.157 \Delta P' \]

\[ = 0.157 (60) \]

\[ = 9.42 \text{ hp/1000 acfm} \]

Thus, the power would be

\[ (9.42 \text{ hp/1000 acfm})(380,000 \text{ acfm}) = 3580 \text{ hp} \]

To determine the brake horsepower, the gas horsepower must be divided by the fan efficiency. Assuming a fan efficiency of 60\%, the operating brake horsepower would be

\[ \text{Brake hp} = \frac{3580}{0.60} \]

\[ = 5970 \text{ hp} \]
VEN.2 COLLECTION EFFICIENCY

A venturi scrubber is employed to reduce the discharge of fly ash to the atmosphere. The unit is presently treating 215,000 acfm of gas, with a concentration of 4.25 gr/ft$^3$, and operating at a pressure drop of 32 in. H$_2$O. Experimental studies have yielded the following particle size collection efficiency data:

<table>
<thead>
<tr>
<th>Particle Diameter (µm)</th>
<th>Weight Fraction $w_i$</th>
<th>Collection Efficiency (%)</th>
</tr>
</thead>
<tbody>
<tr>
<td>5</td>
<td>0.00</td>
<td>30</td>
</tr>
<tr>
<td>10</td>
<td>0.00</td>
<td>42</td>
</tr>
<tr>
<td>20</td>
<td>0.02</td>
<td>86</td>
</tr>
<tr>
<td>30</td>
<td>0.05</td>
<td>93</td>
</tr>
<tr>
<td>50</td>
<td>0.08</td>
<td>97</td>
</tr>
<tr>
<td>75</td>
<td>0.10</td>
<td>98.7</td>
</tr>
<tr>
<td>100</td>
<td>0.75</td>
<td>99.9+</td>
</tr>
</tbody>
</table>

Estimate the overall collection efficiency of the unit.

Solution

The overall efficiency of the unit can be calculated from

$$E_T = \sum w_i E_i$$

The solution is presented in tabular form below:

<table>
<thead>
<tr>
<th>Particle Diameter (µm)</th>
<th>Weight Fraction $w_i$</th>
<th>Collection Efficiency $E_i$ (%)</th>
<th>$w_i E_i$ (%)</th>
</tr>
</thead>
<tbody>
<tr>
<td>5</td>
<td>0.00</td>
<td>30</td>
<td>0.00</td>
</tr>
<tr>
<td>10</td>
<td>0.00</td>
<td>42</td>
<td>0.00</td>
</tr>
<tr>
<td>20</td>
<td>0.02</td>
<td>86</td>
<td>1.72</td>
</tr>
<tr>
<td>30</td>
<td>0.05</td>
<td>93</td>
<td>4.65</td>
</tr>
<tr>
<td>50</td>
<td>0.08</td>
<td>97</td>
<td>7.76</td>
</tr>
<tr>
<td>75</td>
<td>0.10</td>
<td>98.7</td>
<td>9.87</td>
</tr>
<tr>
<td>100</td>
<td>0.75</td>
<td>99.9+</td>
<td>75.00</td>
</tr>
</tbody>
</table>

$$E_T = 99.00$$

VEN.3 DISCHARGE FROM A VENTURI SCRUBBER

With reference to Problem VEN.2, calculate the daily mass (in tons) of fly ash collected by the scrubbing liquid and discharged to the atmosphere. Also obtain the
particle size distribution of the fly ash collected and discharged to the atmosphere. Comment on the results.

**Solution**

The total mass entering is

\[
\text{Mass}_{\text{in}} = \left( \frac{4.25 \text{ gr}}{\text{ft}^3} \right) \left( \frac{215,000 \text{ ft}^3}{\text{min}} \right) \left( \frac{60 \text{ min}}{\text{h}} \right) \left( \frac{24 \text{ h}}{\text{day}} \right) \left( \frac{1 \text{ lb}}{7000 \text{ gr}} \right) \left( \frac{1 \text{ ton}}{2000 \text{ lb}} \right)
\]

\[
= 93.968 \approx 94 \text{ tons/day}
\]

\[
\text{Mass}_{\text{collected}} = (0.99)(94) = 93 \text{ tons/day}
\]

\[
\text{Mass}_{\text{discharged}} = 94 - 93 = 1 \text{ ton/day}
\]

With regard to the particle size distribution calculations, the results are presented in tabular form.

<table>
<thead>
<tr>
<th>Particle Diameter ((\mu\text{m}))</th>
<th>Weight Fraction (W_i)</th>
<th>Mass Entering (tons/day)</th>
<th>Mass Collected (tons/day)</th>
<th>Mass Discharged (tons/day)</th>
</tr>
</thead>
<tbody>
<tr>
<td>5</td>
<td>0.00</td>
<td>0.00</td>
<td>0.00</td>
<td>0.00</td>
</tr>
<tr>
<td>10</td>
<td>0.00</td>
<td>0.00</td>
<td>0.00</td>
<td>0.00</td>
</tr>
<tr>
<td>20</td>
<td>0.02</td>
<td>1.88</td>
<td>1.62</td>
<td>0.26</td>
</tr>
<tr>
<td>30</td>
<td>0.05</td>
<td>4.70</td>
<td>4.37</td>
<td>0.33</td>
</tr>
<tr>
<td>50</td>
<td>0.08</td>
<td>7.52</td>
<td>7.30</td>
<td>0.22</td>
</tr>
<tr>
<td>75</td>
<td>0.10</td>
<td>9.40</td>
<td>9.28</td>
<td>0.12</td>
</tr>
<tr>
<td>100</td>
<td>0.75</td>
<td>70.50</td>
<td>70.43</td>
<td>0.07</td>
</tr>
<tr>
<td>(\sum)</td>
<td></td>
<td>94.00</td>
<td>93.00</td>
<td>1.00</td>
</tr>
</tbody>
</table>

The weight fractions, \(w_i\), for the collected and discharged streams are given below.

<table>
<thead>
<tr>
<th>Particle Diameter ((\mu\text{m}))</th>
<th>(W_i) Collected Mass</th>
<th>(W_i) Discharged Mass</th>
</tr>
</thead>
<tbody>
<tr>
<td>5</td>
<td>0.000</td>
<td>0.00</td>
</tr>
<tr>
<td>10</td>
<td>0.000</td>
<td>0.00</td>
</tr>
<tr>
<td>20</td>
<td>0.017</td>
<td>0.26</td>
</tr>
<tr>
<td>30</td>
<td>0.047</td>
<td>0.33</td>
</tr>
<tr>
<td>50</td>
<td>0.078</td>
<td>0.22</td>
</tr>
<tr>
<td>75</td>
<td>0.100</td>
<td>0.12</td>
</tr>
<tr>
<td>100</td>
<td>0.757</td>
<td>0.07</td>
</tr>
<tr>
<td>(\sum)</td>
<td>0.999</td>
<td>1.00</td>
</tr>
</tbody>
</table>
VEN.4 LIQUID DROPLET SIZE

The following data were collected using a bench-scale venturi scrubber:

\[
\begin{align*}
\text{Gas rate} &= 1.56 \text{ ft}^3/\text{s} \\
\text{Liquid rate} &= 0.078 \text{ gal/min} \\
\text{Throat area} &= 1.04 \text{ in.}^2
\end{align*}
\]

Estimate the average liquid droplet size in the scrubber. Repeat the calculation using a simplified equation.

**Solution**

The size of the droplets generated in scrubber units affects both the collection efficiency and pressure drop, i.e., small droplet sizes requiring high-pressure atomization give greater collection efficiencies. Various correlations are available in the literature to estimate the mean liquid drop diameter from different types of atomizers under different operating conditions. These correlations are applicable to fluids within a certain range of operating conditions and properties such as the volume ratio of liquid to gas, the relative velocity of gas to liquid, the type of nozzle, the surface tension of the liquid, etc. In using one of these correlations to estimate droplet diameter, it is important to select a correlation that takes these factors into consideration.

The empirical relationship of Nukiyama and Tanasawa (NT) is probably the best known and the most widely used to predict the average droplet size in pneumatic (gas-atomized) sprays. In this type of spray the stream of liquid is broken up or atomized by contact with a high-velocity gas stream. The original NT relationship is given by

\[
d_0 = \left( \frac{1920}{v_r} \right) \left( \frac{\sigma}{\rho'_L} \right)^{1/2} + (5.97) \left( \frac{\mu'_L}{\sigma \rho'_L^{1/2}} \right)^{0.45} \left( \frac{L'}{G'} \right)^{1.5}
\]

where \(d_0\) = average surface volume mean droplet diameter, \(\mu\)m
\(v_r\) = relative velocity of gas to liquid, ft/s
\(\sigma\) = liquid surface tension, dyn/cm
\(\rho'_L\) = liquid density, g/cm\(^3\)
\(\mu'_L\) = liquid viscosity, P
\(L'/G'\) = ratio of liquid-to-gas volumetric flowrates at the venturi throat

For water scrubbing systems, one may use

\[
\begin{align*}
\sigma &= 72 \text{ dyn/cm} \\
\rho'_L &= 1.0 \text{ g/cm}^3 \\
\mu'_L &= 0.00982 \text{ Pa}
\end{align*}
\]
This equation reduces to the following expression for standard air and water in a venturi scrubber:

\[ d_0 = \left( \frac{16,400}{v} \right) + 1.45 \, R^{1.5} \]

where \( v \) = gas velocity at venturi throat, ft/s

\( R \) = ratio of liquid-to-gas flowrates, gal/1000 actual ft\(^3\)

From the given data,

\[ G' = 1.56 \, \text{ft}^3/\text{s} \]

\[ L' = (0.078 \, \text{gal/min})(1 \, \text{ft}^3/7.48 \, \text{gal})(1 \, \text{min}/60 \, \text{s}) \]

\[ = 1.74 \times 10^{-4} \, \text{ft}^3/\text{s} \]

\[ A = 1.04 \, \text{in.}^2 = 7.22 \times 10^{-3} \, \text{ft}^2 \]

The gas velocity is

\[ v_G = G'/A = 1.56/7.22 \times 10^{-3} \]
\[ = 216 \, \text{ft/s} \]

The liquid velocity is

\[ v_L = L'/A = 1.74 \times 10^{-4}/7.22 \times 10^{-3} \]
\[ = 0.0241 \, \text{ft/s} \]

The relative velocity is

\[ v_r = v_G - v_L \approx v_G = 216 \, \text{ft/s} \]

Using the Nukiyama and Tanasawa correlation for droplet diameter,

\[ d_0 = \left( \frac{1920}{v_r} \right) \left( \frac{\sigma}{\rho_L} \right)^{1/2} + (5.97)\left( \frac{\mu_L}{(\sigma \rho_L)^{1/2}} \right)^{0.45} \left( \frac{1000 \, L'}{G'} \right)^{1.5} \]

\[ = \left( \frac{1920}{216} \right) \left( \frac{72}{1'} \right)^{1/2} + (5.97)\left( \frac{0.00982}{[(72)(1)]^{1/2}} \right)^{0.45} \left( \frac{1000 \, 1.74 \times 10^{-4}}{1.56} \right)^{1.5} \]

\[ = 75.4 \, \mu\text{m} \]
The ratio of liquid-to-gas flow rates is

\[ R = \frac{(0.078 \text{ gal/min})(1 \text{ min}/60 \text{ s})(1000)}{(1.56 \text{ ft}^3/\text{s})} = 0.833 \text{ gal/1000 acf} \]

Using the simplified equation gives

\[ d_0 = \frac{(16,400/v)}{216} + 1.45 \cdot 0.833 \]
\[ = \left(\frac{16,400}{216}\right) + 1.45 \cdot 0.833 \]
\[ = 77.03 \mu\text{m} \]

VEN.5 PRESSURE DROP

Using the data provided in Problem VEN.4, estimate the pressure drop across the bench-scale unit. Use both the Theodore and Calvert equations.

Solution

The pressure drop for gas flowing through a venturi scrubber can be estimated from knowledge of liquid acceleration and frictional effects along the wall of the equipment. Frictional losses depend largely on the scrubber geometry and usually are determined experimentally. The effect of liquid acceleration is, however, predictable. An equation (developed by Calvert) for estimating pressure drop through venturi scrubbers (given as a function of throat gas velocity and liquid-to-gas ratio) assuming that all the energy is used to accelerate the liquid droplets to the throat velocity of the gas is

\[ \Delta P = 5 \times 10^{-5} \cdot v^2 R \]

where \( \Delta P \) is the pressure drop, in. H\(_2\)O, \( v \) is the gas velocity, ft/s, and \( R \) is the liquid-to-gas ratio, gal/1000 acf. Another somewhat simpler equation that applies over a fairly wide range of values for \( R \) is given below (L. Theodore, personal notes):

\[ \Delta P' = 0.8 + 0.12R \]

where \( \Delta P' \) is a dimensionless pressure drop equal to the pressure drop divided by the density and velocity head \((v^2/2g_c)\).
Using the Calvert equation to estimate the pressure drop,

\[ \Delta P = (5 \times 10^{-5})(216)^2(0.833) \]
\[ = 1.943 \text{ in. H}_2\text{O} \]

Using Theodore's equation,

\[ \Delta P' = 0.8 + (0.12)(0.833) \]
\[ = 0.90 \]

Thus,

\[ \Delta P = (0.90)(v^2/2g_c)\rho \]
\[ = (0.90)\left(\frac{216^2}{(2)(32.2)}\right)(0.0775) \]
\[ = 50.5 \text{ psf} \]

Since 1 psf = 0.1922 in. H\(_2\)O,

\[ \Delta P = 9.71 \text{ in. H}_2\text{O} \]

The Calvert equation significantly underpredicts the pressure drop at low values of \( R \). Note that this equation fails when \( R \) is zero.

VEN.6 THROAT AREA

A consulting firm has been requested to calculate the throat area of a venturi scrubber to operate at a specified collection efficiency.

To achieve high collection efficiency of particulates by impaction, a small droplet diameter and high relative velocity between the particle and droplet are required. In a venturi scrubber this is often accomplished by introducing the scrubbing liquid at right angles to a high-velocity gas flow in the venturi throat (vena contracta). Very small water droplets are formed, and high relative velocities are maintained until the droplets are accelerated to their terminal velocity. Gas velocities through the venturi throat typically range from 12,000 to 24,000 ft/min. The velocity of the gases alone causes the atomization of the liquid.

Perhaps the most popular and widely used venturi scrubber collection efficiency equation is that originally suggested by Johnstone:

\[ E = 1 - e^{-kR\psi^{1/2}} \]
where $E$ = fractional collection efficiency

\[ k = \text{correlation coefficient whose value depends on the system geometry and operating conditions, typically 0.1–0.2, 1000 acf/gal} \]

\[ R = \frac{q_L}{q_G} = \text{liquid-to-gas ratio, gal/1000 acf} \]

\[ \psi = C \rho_p v d_p^2 / 18 d_o \mu, \text{the inertial impaction parameter} \]

\[ \rho_p = \text{particle density, lb/ft}^3 \]

\[ v = \text{gas velocity at venturi throat, ft/s} \]

\[ d_p = \text{particle diameter, ft} \]

\[ d_o = \text{droplet diameter, ft} \]

\[ \mu = \text{gas viscosity, (lb/ft} \cdot \text{s)} \]

\[ C = \text{Cunningham correction factor} \]

*Note:* Some engineers define $\psi$ as

\[ \psi = \frac{C \rho_p v d_p^2}{9 d_o \mu} \]

This change is reflected in the correlation coefficient, $k$.

Pertinent data are given below.

- Volumetric flowrate of process gas stream = 11,040 acfm (at 68°F)
- Density of dust = 187 lb/ft$^3$
- Liquid-to-gas ratio = 2 gal/1000 ft$^3$
- Average particle size = 3.2 μm (1.05 × 10$^{-5}$ ft)
- Water droplet size = 48 μm (1.575 × 10$^{-4}$ ft)
- Johnstone scrubber coefficient, $k$ = 0.14
- Required collection efficiency = 98%
- Viscosity of gas = 1.23 × 10$^{-5}$ lb/(ft · s)
- Cunningham correction factor = 1.0

**Solution**

Calculate the inertial impaction parameter, $\psi$, from Johnstone's equation:

\[ E = 1 - e^{-kR\psi^{1/2}} \]

\[ 0.98 = 1 - e^{-(0.14)(2)\psi^{1/2}} \]

Solving for $\psi$,

\[ \psi = 195.2 \]
From the calculated value of $\psi$ above, back calculate the gas velocity at the venturi throat, $v$:

$$
\psi = \frac{\rho_p v d_p^2}{18 d_0 \mu}
$$

$$
v = \frac{18 \psi d_0 \mu}{\rho_p d_p^2} = \frac{(18)(195.2)(1.575 \times 10^{-4})(1.23 \times 10^{-5})}{(187)(1.05 \times 10^{-5})^2}
$$

$$
= 330.2 \text{ ft/s}
$$

Calculate the throat area, $S$, using gas velocity at the venturi throat, $v$:

$$
S = \frac{q}{v} = \frac{(11,040)}{(60)(330.2)}
$$

$$
= 0.557 \text{ ft}^2
$$

**VEN.7 THREE VENTURI SCRUBBERS IN SERIES**

Three identical venturi scrubbers are connected in series. If each operates at the same efficiency and liquid-to-gas ratio, $q_L/q_G$, calculate the liquid-to-gas ratio, assuming the Johnstone equation to apply. Data are provided below.

- $E_0$ (overall) = 99%
- Inlet loading = 200 gr/ft$^3$
- Johnstone scrubber coefficient, $k = 0.14$
- Inertial impaction parameter, $\psi = 105$

**Solution**

First calculate the outlet loading (OL) from the last unit:

$$
OL = IL (1 - E_o)
$$

$$
= 200 (1 - 0.99)
$$

$$
= 2.0 \text{ gr/ft}^3
$$

Express the individual and overall efficiencies in terms of the penetration $P$:

$$
P_0 = 1 - E_o = 1 - 0.99 = 0.01
$$

$$
P_1 = 1 - E_1
$$

$$
P_2 = 1 - E_2
$$

$$
P_3 = 1 - E_3
$$
Calculate the individual efficiency for each venturi scrubber, noting that the efficiencies (or penetrations) are equal:

\[ P_3 = P_1 P_2 P_3 = P^3 \]
\[ P^3 = 0.01 \]
\[ P = 0.215 \]
\[ E = 1 - P \]
\[ = 1 - 0.215 \]
\[ = 0.785 = 78.5\% \]

Using the Johnstone equation, solve for the liquid-to-gas ratio, \( q_L/q_G \):

\[ \ln(1 - E) = -k \left( \frac{q_L}{q_G} \right) \phi^{0.5} \]
\[ \left( \frac{q_L}{q_G} \right) = - \frac{\ln(1 - E)}{k \phi^{0.5}} = - \frac{\ln(1 - 0.785)}{(0.14)(105)^{0.5}} \]
\[ = 1.07 \text{ gal/1000 acf} \]
\[ = 1.07 \text{ gpm/1000 acfm} \]

VEN.8 COMPLIANCE CALCULATIONS ON A SPRAY TOWER

Contact power theory is an empirical approach relating particulate collection efficiency and pressure drop in wet scrubber systems. The concept is an outgrowth of the observation that particulate collection efficiency in spray-type scrubbers is mainly determined by pressure drop for the gas plus any power expended in atomizing the liquid. Contact power theory assumes that the particulate collection efficiency in a scrubber is solely a function of the total power loss for the unit. The total power loss, \( P_T \), is assumed to be composed of two parts: the power loss of the gas passing through the scrubber, \( P_G \), and the power loss of the spray liquid during atomization, \( P_L \). The gas term can be estimated by

\[ P_G = 0.157 \Delta P \]

where \( P_G \) is the contacting power based on gas stream energy input in hp/1000 acfm and \( \Delta P \) is the pressure drop across the scrubber in inches of water. In addition,

\[ P_L = 0.583 P_L (q_L/q_G) \]

where \( P_L \) is the contacting power based on liquid stream energy input in hp/1000 acfm, \( P_L \) is the liquid inlet pressure in psi, \( q_L \) is the liquid feed rate in gal/min, and \( q_G \) is the gas flowrate in ft\(^3\)/min. Then

\[ P_T = P_G + P_L \]
To correlate contacting power with scrubber collecting efficiency, the latter is best expressed as the number of transfer units. The number of transfer units is defined by analogy to mass transfer and given by

$$N_t = \ln\left(\frac{1.0}{1 - E}\right)$$

where $N_t$ is the number of transfer units, dimensionless, and $E$ is the fractional collection efficiency, dimensionless. The relationship between the number of transfer units and collection efficiency is by no means unique. The number of transfer units for a given value of contacting power (hp/1000 acfm) or vice versa varies over nearly an order of magnitude. For example, at 2.5 transfer units ($E = 0.918$), the contacting power ranges from approximately 0.8 to 10.0 hp/1000 acfm, depending on the scrubber and the particulate.

For a given scrubber and particulate properties, there will usually be a very distinct relationship between the number of transfer units and the contacting power. The number of transfer units for a series of scrubbers and particulates is plotted against total power consumption; a linear relation, independent of the type of scrubber, is obtained on a log-log plot. The relationship could be expressed by

$$N_t = \alpha \rho^\beta_T$$

where $\alpha$ and $\beta$ are the parameters for the type of particulates being collected and the scrubber unit.

A vendor proposes to use a spray tower on a lime kiln operation to reduce the discharge of solids to the atmosphere. The inlet loading is to be reduced to meet state regulations. The vendor's design calls for a certain water pressure drop and gas pressure drop across the tower. You are requested to determine whether this spray tower will meet state regulations. If the spray tower does not meet state regulations, propose a set of operating conditions that will meet the regulations. The state regulations require a maximum outlet loading of 0.05 gr/ft$^3$. Assume that contact power theory applies. Operating and design data are provided:

- Gas flowrate = 10,000 acfm
- Water rate = 50 gal/min
- Inlet loading = 5.0 gr/ft$^3$
- Maximum gas pressure drop across the unit = 15 in. H$_2$O
- Maximum water pressure drop across the unit = 100 psi

The vendor's design and operating data are also available:

- $\alpha = 1.47$
- $\beta = 1.05$
- Water pressure drop = 80 psi
- Gas pressure drop across the tower = 5.0 in. H$_2$O
Solution

Calculate the contacting power based on the gas stream energy input, $P_G$, in hp/1000 acfm:

$$P_G = (0.157) \Delta P$$
$$= (0.157)(5.0)$$
$$= 0.785 \text{ hp/1000 acfm}$$

Calculate the contacting power based on the liquid stream energy input, $P_L$, in hp/1000 acfm:

$$P_L = 0.583 P_L \left(\frac{q_L}{q_G}\right)$$
$$= (0.583)(80)(50/10,000)$$
$$= 0.233 \text{ hp/1000 acfm}$$

The total power loss, $P_T$, in hp/1000 acfm is then

$$P_T = P_G + P_L$$
$$= 0.785 + 0.233$$
$$= 1.018 \text{ hp/1000 acfm}$$

The number of transfer units, $N_t$, is

$$N_t = \alpha \beta_T$$
$$= (1.47)(1.018)^{1.05}$$
$$= 1.50$$

The collection efficiency can be calculated based on the design data given by the vendor:

$$N_t = \ln \left(\frac{1.0}{1 - E}\right)$$

or

$$E = 1 - e^{-N_t} = 1 - e^{-1.50}$$
$$= 77.7\%$$
The collection efficiency required by state regulations, \( E_s \), is

\[
E_s = \frac{\text{Inlet loading} - \text{Outlet loading}}{\text{Inlet loading}} \quad (100)
\]

\[
= \frac{5.0 - 0.05}{5.0} \quad (100)
\]

\[= 99.0\%\]

Since \( E_s > E \), the spray tower does not meet the regulations.

One may now propose a set of operating conditions that will meet the regulations:

\[
N_t = \ln\left(\frac{1.0}{1 - E}\right) = \ln\left(\frac{1.0}{1 - 0.99}\right)
\]

\[= 4.605\]

The total power loss, \( P_T \), in hp/1000 acfm is

\[
N_t = \alpha P_T^{0.7}
\]

\[4.605 = (1.47)(P_T)^{1.05}\]

Solving for \( P_T \)

\[P_T = 2.96 \text{ hp/1000 acfm}\]

Calculate the contacting power based on the gas stream energy input, \( P_G \), using a \( \Delta P \) of 15 in. H\(_2\)O:

\[
P_G = 0.157 \Delta P
\]

\[= (0.157)(15)
\]

\[= 2.355 \text{ hp/1000 acfm}\]

The liquid stream energy input, \( P_L \), is then

\[
P_L = P_T - P_G
\]

\[= 2.96 - 2.355
\]

\[= 0.605 \text{ hp/1000 acfm}\]
Calculate $q_L/q_G$, in gal/acf, using $P_L$ in psi:

$$
q_L/q_G = \frac{P_L}{(0.583)(P_L)}
= \frac{0.605}{(0.583)(100)}
= 0.0104
$$

The new water flow rate, $q'_L$ in gal/min, is therefore

$$
q'_L = (q_L/q_G)(10,000\text{ acfm})
= (0.0104)(10,000\text{ acfm})
= 104\text{ gal/min}
$$

The new set of operating conditions that will meet the regulations are

$$
\Delta P = 15\text{ in. H}_2\text{O} \\
P_L = 100\text{ psi} \\
q'_L = 104\text{ gal/min} \\
\mathcal{P}_T = 2.96\text{ hp/1000 acfm}
$$

Unlike the Johnstone equation approach, this method requires specifying two coefficients. The validity and accuracy of the coefficients available from the literature for the contact power theory equations have been questioned. Some numerical values of $\alpha$ and $\beta$ for specific particulates and scrubber devices are provided below.

<table>
<thead>
<tr>
<th>Aerosol</th>
<th>Scrubber Type</th>
<th>$\alpha$</th>
<th>$\beta$</th>
</tr>
</thead>
<tbody>
<tr>
<td>Raw gas (lime dust and soda fume)</td>
<td>Venturi and cyclonic spray</td>
<td>1.47</td>
<td>1.05</td>
</tr>
<tr>
<td>Prewashed gas (soda fume)</td>
<td>Venturi, pipe line, and cyclonic spray</td>
<td>0.915</td>
<td>1.05</td>
</tr>
<tr>
<td>Talc dust</td>
<td>Venturi</td>
<td>2.97</td>
<td>0.362</td>
</tr>
<tr>
<td>Black liquor recovery furnace fume</td>
<td>Venturi and cyclonic spray</td>
<td>1.75</td>
<td>0.620</td>
</tr>
<tr>
<td>Phosphoric acid mist</td>
<td>Venturi</td>
<td>1.33</td>
<td>0.647</td>
</tr>
<tr>
<td>Foundry cupola dust</td>
<td>Venturi</td>
<td>1.35</td>
<td>0.621</td>
</tr>
<tr>
<td>Open-hearth steel furnace fume</td>
<td>Venturi</td>
<td>1.26</td>
<td>0.569</td>
</tr>
<tr>
<td>Talc dust</td>
<td>Cyclone</td>
<td>1.16</td>
<td>0.655</td>
</tr>
<tr>
<td>Ferrosilicon furnace fume</td>
<td>Venturi and cyclonic spray</td>
<td>0.870</td>
<td>0.459</td>
</tr>
<tr>
<td>Odorous mist</td>
<td>Venturi</td>
<td>0.363</td>
<td>1.41</td>
</tr>
</tbody>
</table>
A venturi scrubber is being designed to remove particulates from a gas stream. The maximum gas flowrate of 30,000 acfm has a loading of 4.8 gr/ft³. The average particle size is 1.2 μm and the particle density is 200 lb/ft³. Neglect the Cunningham correction factor. The Johnstone coefficient, $k$, for this system is 0.15. The proposed water flowrate is 180 gal/min and the gas velocity is 250 ft/s.

1. What is the efficiency of the proposed system?
2. What would the efficiency be if the gas velocity were increased to 300 ft/s?
3. Determine the pressure drop for both gas velocities. Assume Calvert's equation to apply.
4. Determine the daily mass of dust collected and discharged for each gas velocity.
5. What is the discharge loading in each case?

**Solution**

1. The ratio of liquid-to-gas flowrates is given by
   \[ R = \frac{(180)(1000)}{(30,000)} = 6.0 \text{ gal/1000 acf} = 6.0 \text{ gpm/acfm} \]
   and
   \[ v_G = 250 \text{ ft/s} \]
   \[ d_p = 1.2 \text{ μm} = 3.937 \times 10^{-6} \text{ ft} \]
   \[ \rho_p = 200 \text{ lb/ft}^3 \]
   \[ \mu = 1.23 \times 10^{-5} \text{ lb/(ft} \cdot \text{s}) \]

   Assume the Nukiyama–Tanasawa (NT) equation to apply:
   \[ d_0 = \frac{(16,400/v)}{1.45 R^{1.5}} \]
   \[ = \frac{(16,400/250)}{1.45 (6.0)^{1.5}} \]
   \[ = 86.91 \mu m \]

   \[ N_1 = \frac{d_p^2 \rho_p v}{9 \mu d_0} \]
   \[ = \frac{(3.937 \times 10^{-6})^2 (200)(250)}{(9)(1.23 \times 10^{-5})(2.85 \times 10^{-4})} \]
   \[ = 24.56 \]

   \[ E = 1 - e^{-kR} \sqrt{N_1} \]
   \[ = 1 - e^{-(0.15)(6)} \sqrt{24.56} \]
   \[ = 0.9884 = 98.84\% \]
2. If \( v \) were increased to 300 ft/s,

\[
d_0 = \left(\frac{16,400}{300}\right) + 1.45(6.0)^{1.5}
= 75.98 \mu m
\]

\[
N_1 = \frac{(3.937 \times 10^{-6})^2(200)(300)}{(9)(1.23 \times 10^{-5})(2.85 \times 10^{-4})}
= 29.48
\]

\[
E = 1 - e^{-0.15(6)\sqrt{29.48}}
= 99.24\%
\]

3. The pressure drops are given by (see Problem VEN.5)

\[
\Delta P_a = (5 \times 10^{-5})(250)^2(6) = 18.75 \text{ in. H}_2\text{O}
\]

\[
\Delta P_b = (5 \times 10^{-5})(300)^2(6) = 27 \text{ in. H}_2\text{O}
\]

4. The total mean loading, TML, is

\[
TML = (4.8 \text{ gr/ft}^3)(30,000)(60)(24)/7000
= 29,600 \text{ lb/day} = 14.81 \text{ tons/day}
\]

For \( v = 250 \text{ ft/s} \),

Dust collected = (0.9884)(29,600)
= 29,300 \text{ lb/day}

Dust discharged = 344 \text{ lb/day}

For \( v = 300 \text{ ft/s} \),

Dust collected = (0.9924)(29,600)
= 29,400 \text{ lb/day}

Dust discharged = 225 \text{ lb/day}
5. The discharge loading (DL) for $v = 250 \text{ ft/s}$ is

$$DL = (4.8)(1 - E) = (4.8)(1 - 0.9884)$$

$$= 0.056 \text{ gr/ft}^3$$

and for $v = 300 \text{ ft/s}$ is

$$DL = (4.8)(1 - E) = (4.8)(1 - 0.9924)$$

$$= 0.036 \text{ gr/ft}^3$$

VEN.10 OPEN-HEARTH FURNACE APPLICATION

The installation of a venturi scrubber is proposed to reduce the discharge of particulates from an open-hearth steel furnace operation. Preliminary design information suggests water and gas pressure drops across the rubber of $5.0 \text{ psia}$ and $36.0 \text{ in. of H}_2\text{O}$, respectively. A liquid-to-gas ratio of $6.0 \text{ gpm/1000 acfm}$ is usually employed with this industry. Estimate the collection efficiency of the proposed venturi scrubber. Assume contact power theory to apply with $\alpha$ and $\beta$ given by 1.26 and 0.57, respectively. Recalculate the collection efficiency if the power requirement on the liquid side is neglected.

Solution

Due to the low water pressure drop, it can be assumed that

$$\mathcal{P}_G \gg \mathcal{P}_L \quad \mathcal{P}_T \approx \mathcal{P}_G$$

with

$$\mathcal{P}_G = 0.157 (\Delta P)$$

Solving for $\mathcal{P}_G$ gives

$$\mathcal{P}_G = (0.157)(36)$$

$$= 5.65 \text{ hp/1000 acfm}$$

The number of transfer units is calculated from

$$N_t = \alpha \mathcal{P}_T^\beta$$

$$= (1.26)(5.65)^{0.57}$$

$$= 3.38$$
The collection efficiency can now be calculated:

\[ N_t = 3.38 = \ln\left(\frac{1}{1 - E}\right) \]

\[ E = 0.966 = 96.6\% \]

Since the power requirement on the liquid side is neglected, the efficiency remains the same.
HYB.1 DESCRIPTION OF HYBRID SYSTEMS

Briefly describe hybrid systems.

Solution

Hybrid systems are defined as those types of control devices that involve combinations of control mechanisms, for example, fabric filtration combined with electrostatic precipitation. Unfortunately, the term hybrid system has come to mean different things to different people. The two most prevalent definitions employed today for hybrid systems are:

1. Two or more pieces of different air pollution control equipment connected in series, e.g., a baghouse followed by an absorber.
2. An air pollution control system that utilizes two or more collection mechanisms simultaneously to enhance pollution capture, e.g., an ionizing wet scrubber (IWS), which will be discussed shortly.

HYB.2 DRY SO₂ SCRUBBER

Estimate the water requirement of a spray dryer (dry SO₂ scrubber) at a coal-fired incineration facility that treats 150,000 lb/h of a flue gas at 2180°F. Assume an approach temperature to the adiabatic saturation temperature (AST) of 40°F. The AST can be assumed to be 180°F.

Solution

A spray dryer flue gas desulfurization (FGD) operation consists of four major steps:

1. Absorbent preparation
2. Absorption and drying
3. Solids collection
4. Solids disposal
Flue gas exiting the process (usually the combustion air preheater) comes in contact with an alkaline-water solution in a spray dryer. The flue gas passes through a contact chamber, and the solution or slurry is sprayed into the chamber with a rotary or nozzle atomizer. The heat of the flue gas evaporates the water in the atomized droplets while the droplets absorb SO\textsubscript{2} from the flue gas. The SO\textsubscript{2} reacts with the alkaline reagent to form solid-phase sulfite and sulfate salts. Most of the solids (and any fly ash present) are carried out of the dryer in the exiting flue gas. The rest fall to a hopper at the bottom of the dryer. With spray drying, in contrast to wet FGD, the flue gas is not saturated with moisture after the absorption step. However, the gas approaches within 20–50°F (11–28°C) of the adiabatic saturation temperature (AST). The water requirement to “cool” the gases to a temperature approaching the AST is an important design and operational requirement.

*Note:* Normally, cooling by liquid quenching is essentially accomplished by introducing a liquid (usually water) directly to the hot gases. When the water evaporates, the heat of vaporizing the water is obtained at the expense of the hot combustion gas, resulting in a reduction in the gas temperature. The temperature of the combustion gases discharged from the unit is at the adiabatic saturation temperature of the combustion gas if the operation is adiabatic and the gas leaves the unit saturated with water vapor. (A saturated gas contains the maximum water vapor possible at the temperature; any increase in water content will result in condensation.) Simple calculational and graphical procedures are available for estimating the adiabatic saturation temperature of a gas (see *Introduction to Hazardous Waste Incineration*, Santoleri, Reynolds, and Theodore, Wiley-Interscience, 2000).

For hot combustion gases being cooled approximately 2000°F, the quench water requirement may be estimated by (personal notes, L. Theodore)

\[ w_{\text{water}} = \frac{1}{2} w_{\text{flue}} \]

where \( w_{\text{water}} \), \( w_{\text{flue}} \) are the water and flue gas flow rates, respectively, in consistent units. The AST of most combustion gases is approximately 175°F.

Calculate the discharge temperature \( T \) from the spray dryer section of the dry scrubber:

\[ T = 40 + \text{AST} \]
\[ = 40 + 180 = 220°F \]

The dry scrubber water requirement rate is approximately

\[ w_{\text{water}} = \frac{1}{2} w_{\text{flue}} \]
\[ = (0.5)(150,000) \]
\[ = 75,000 \text{ lb/h} \]
HYB.3 LIME REQUIREMENT FOR A SPRAY DRYER

Combustion of a hazardous waste produces 15,000 acfm of flue gas at 700°F and 1 atm. You have been asked to calculate lime requirements for this process. HCl and SO₂ concentrations are 10,000 and 250 ppm, respectively. HCl must be controlled to 99% collection efficiency or 4 lb/h. SO₂ emissions are to be controlled at 70% collection efficiency. A spray dryer is used to control the HCl and SO₂ emissions. Ca(OH)₂ is the sorbent that will react with HCl and SO₂ to form CaCl₂ and CaSO₄, respectively. Assume that it is necessary to provide 10% excess lime feed for the required HCl removal and 30% excess lime feed for total SO₂ removal.

What is the required feed rate of Ca(OH)₂? What is the total mass production rate of solids from the spray dryer? Assume that the excess solids in the spray dryer are Ca(OH)₂. Also size the spray dryer if the residence time (based on actual inlet conditions) is 10 s and the length-to-diameter ratio of the unit is 1.75 (neglecting hopper volume).

Solution

Determine the mass flowrates of HCl and SO₂ in the flue gas:

\[ PV = nRT \]
\[ Pq = \dot{n}RT \]
\[ \dot{m} = \dot{n}(MW) = Pq(MW)/RT \]
\[ \dot{m}_{HCl} = \frac{(1 \text{ atm})(0.01)(15,000 \text{ acfm})(36.5 \text{ lb/lbmol})}{[0.7302 \text{ atm} \cdot \text{ft}^3/\text{lrbmol} \cdot \circ R]} \frac{(460 + 700\circ R)(1 \text{ h}/60 \text{ min})}{(460 + 700\circ R)(1 \text{ h}/60 \text{ min})} = 387.8 \text{ lb HCl/h} \]
\[ \dot{m}_{SO₂} = \frac{(1 \text{ atm})(0.00025)(15,000 \text{ acfm})(64 \text{ lb/lbmol})}{[0.7302 \text{ atm} \cdot \text{ft}^3/\text{lrbmol} \cdot \circ R]} \frac{(460 + 700\circ R)(1 \text{ h}/60 \text{ min})}{(460 + 700\circ R)(1 \text{ h}/60 \text{ min})} = 17.0 \text{ lb SO₂/h} \]

Determine which appropriate regulation applies for HCl control; 0.01(387.8 lb/h) = 3.88 lb/h which is less than 4 lb/h. Therefore, the 4 lb/h rule applies.

Write the two balanced chemical reaction equations (one for HCl and one for SO₂):

\[ 1.3 \text{Ca(OH)₂} + \text{SO₂} + \frac{1}{2} \text{O₂} \rightarrow \text{CaSO₄} + \text{H₂O} + 0.3 \text{Ca(OH)₂} \]
\[ 1.1 \text{Ca(OH)₂} + 2 \text{HCl} \rightarrow \text{CaCl₂} + 2 \text{H₂O} + 0.1 \text{Ca(OH)₂} \]
Determine the molar amount of Ca(OH)\(_2\) needed for neutralization. SO\(_2\) removal requires 1.3 mol lime/mol SO\(_2\), while HCl removal requires 1.1/2 or 0.55 mol lime/mol HCl.

Determine the feed rate of Ca(OH)\(_2\) required in lbmol/h:

\[
\text{Lime feed rate} = \left(\frac{17 \text{ lb SO}_2/\text{ h}}{64 \text{ lb SO}_2/\text{lbmol}}\right) \left(\frac{1.3 \text{ lbmol lime}}{\text{lbmol SO}_2}\right) + \left(\frac{(387.3 - 4.0) \text{ lb HCl/ h}}{36.45 \text{ lb HCl/lbmol}}\right) \left(\frac{0.55 \text{ lbmol lime}}{\text{lbmol HCl}}\right)
\]

\[
= 6.13 \text{ lbmol lime/ h}
\]

Calculate the Ca(OH)\(_2\) feed rate in lb/h:

\[
\text{Lime feed rate} = (6.13)(74)
\]

\[
= 453 \text{ lb/ h}
\]

Determine the production rate of CaSO\(_4\) solids in lb/h. There is one mol of CaSO\(_4\) produced per mol of SO\(_2\) reacted. Since 70% of the SO\(_2\) reacts,

\[
\text{Production rate of CaSO}_4 = (17)(0.7)/64 = 0.186 \text{ lbmol/ h}
\]

\[
= (0.186)(136) = 25.3 \text{ lb/h}
\]

Determine the CaCl\(_2\) produced in lb/h. One mole of CaCl\(_2\) is produced per 2 mol of HCl reacted; 4 lb HCl are not reacted:

\[
\text{Production rate} = (387.3 - 4.0)(0.5)/36.45 = 5.258 \text{ lbmol/h}
\]

\[
= (5.258)(110.9) = 583.1 \text{ lb/h}
\]

Determine the unreacted Ca(OH)\(_2\) remaining in lb/h. One mol of Ca(OH)\(_2\) is required to react in the production of either 1 mol of CaSO\(_4\) or 1 mol of CaCl\(_2\). Using previous results,

\[
\text{Ca (OH)}_2 \text{unreacted} = 6.12 - 5.258 - 0.186 = 0.676 \text{ lbmol/h}
\]

\[
= (0.676)(74) = 50.0 \text{ lb/h}
\]

Calculate the total solids produced in lb/h:

\[
\text{Total solids} = \text{CaCl}_2 + \text{CaSO}_4 + \text{Ca(OH)}_2 \text{ (unreacted)}
\]

\[
= 583.1 + 25.3 + 50.0
\]

\[
= 658.4 \text{ lb/h}
\]
Calculate the volume of the spray dryer in ft$^3$:

\[ V = (15,000)(10)/(60) \]
\[ = 2500 \text{ ft}^3 \]

Size the spray dryer. Assume a cylindrical shape with $L =$ length and $D =$ diameter.

\[ V = \pi D^2 L/4 \]

Noting that $L$ has been specified to be $1.75D$,

\[ V = \pi D^2 (1.75D)/4 \]
\[ D^3 = (2500)(4)/(\pi)(1.75) \]
\[ = 1818.9 \text{ ft}^3 \]
\[ D = 12.2 \text{ ft} \]
\[ L = (1.75)(12.2) \]
\[ = 21.4 \text{ ft} \]

**HYB.4 SPRAY DRYER VS. WET SCRUBBER**

List the advantages the spray dryer has over traditional wet scrubbers.

**Solution**

Among the inherent advantages that the spray dryer enjoys over wet scrubbers are

1. Lower capital costs
2. Lower draft losses
3. Reduced auxiliary power
4. Reduced water consumption, with liquid-to-gas (L/G) ratios significantly lower than those of wet scrubbers
5. Continuous, two-stage operation

**HYB.5 WET ELECTROSTATIC PRECIPITATOR (ESP) DESIGN**

An exhaust stream from an industrial operation with a particulate loading of 5.0 gr/ft$^3$ and a flowrate of 660 acfm is to be treated with a wet tubular ESP. Based on the data provided below, design the unit.
Outlet loading (required) = 0.08 gr/ft$^3$
Migration (drift) velocity = 0.52 ft/s
Cylindrical tube diameter = 6 in.
Suggested average throughput velocity = 3 ft/s
Water requirement = 10 gal/(min • tube)

Assume the Deutsch–Anderson equation to apply.

Tubular precipitators, generally used for collecting mists or fogs, consist of cylindrical collection electrodes with discharge electrodes located in the center of the cylinders. Dirty gas flows into the cylinder where precipitation occurs. The negatively charged particles migrate to and are collected on grounded collecting tubes. The collected dust or liquid is removed by washing the tubes with water sprays located directly above the tubes. (These precipitators have also been referred to as water-walled ESPs.) Tube diameters typically vary from 0.5 to 1 ft (0.15 to 0.31 m), with length usually ranging from 6 to 15 ft (1.85 to 4.6 m).

Solution

The required collection efficiency $E$ is

$$E = \frac{(5.0 - 0.08)}{5.0}$$

$$= 0.984 = 98.4\%$$

Apply the DA equation and calculate the required collection area:

$$E = 1 - e^{-\left(\frac{q}{Aw}\right)}$$

$$A = -\left(\frac{q}{W}\right) \ln(1 - E)$$

$$= -\left(\frac{660}{(0.52)(60)}\right) \ln(1 - 0.984)$$

$$= 87.5 \text{ ft}^2$$

The cross-sectional area of each tube, $A_c$, is

$$A_c = \frac{\pi(D)^2}{4}$$

$$= \frac{\pi(6/12)^2}{4}$$

$$= 0.196 \text{ ft}^2$$
Calculate the volumetric flowrate of exhaust gas, $q_1$, passing through one tube:

$$q_1 = vA$$

$$= (3)(0.196) = 0.588 \text{ acfs}$$

$$= 35.3 \text{ acfm}$$

Determine the required number of tubes, $n$:

$$n = \frac{q}{q_1}$$

$$= \frac{660}{35.3}$$

$$= 18.7 \approx 19 \text{ tubes}$$

Calculate the length of each tube:

$$n\pi DL = 87.5 \text{ ft}^2$$

$$L = \frac{87.5}{[(\pi)(0.5)(19)]}$$

$$= 2.93 \approx 3 \text{ ft}$$

Calculate the water requirement, $W$, in gal/day:

$$W = (10)(19)(60)(24)$$

$$= 273,600 \text{ gal/day}$$

The design appears adequate although the tube length is a bit short. This can be compensated for by operating with a higher throughput velocity. This in turn will correspondingly decrease the required number of tubes.

**HYB.6 ADVANTAGES AND DISTADVANTAGES OF WET ELECTROSTATIC PRECIPITATORS**

List the advantages and disadvantages of wet electrostatic precipitator (WEP) usage.

**Solution**

Some of the advantages of a WEP include:

1. Simultaneous gas adsorption and dust removal.
2. Low energy consumption.
3. No dust resistivity problems.
4. Efficient removal of fine particles.
Disadvantages of the WEP are the following:

1. Low gas absorption efficiency.
2. Sensitivity to changes in flowrate.
3. Dust collection is wet.

HYB.7 IONIZING WET SCRUBBER CALCULATIONS

A three-stage Ceilcote ionizing wet scrubber (IWS) is currently treating an 8.0-ft/s discharge stream from a hospital waste incinerator. The inlet (from the incinerator) and outlet particulate loadings to/from the IWS are 1.23 and 0.017 gr/dscf, respectively, and it operates at a pressure drop of 4.75 in. H$_2$O. Ceilcote provided the following design data on this system:

- Pressure drop = 1.55 in. H$_2$O/stage
- Number of stages = 3
- Collection efficiency of each stage = 74%
- Average bulk throughput velocity = 8.0 ft/s
- Length of packing per stage = 5 ft

Compare the operating conditions of pressure drop and collection efficiency with the design specifications provided by the vendor.

Solution

Note: Much of the following writeup has been drawn (with permission) from the Ceilcote literature (Ceilcote Air Pollution Control, Strangeville, OH).

The term ionizing wet scrubber was first used by the Ceilcote Co., located in Berea, Ohio, and has found wide application in the air pollution control field. This system is a proven means for the removal of pollutants from industrial process gas streams. The IWS combines the established principles of electrostatic particle charging, image force attraction, inertial impaction, and gas absorption to collect submicron solid particles, liquid particles, and noxious and malodorous gases simultaneously. The IWS system requires little energy and its collection efficiency is high for both submicron and micron size particles.

The ionizing wet scrubber utilizes high-voltage ionization to electrostatically charge particulate matter in the gas stream before the particles enter a Tellerette packed scrubber section where they are removed by attraction of the charged particles to neutral surfaces. Larger particles equal to or greater than 3–5 μm are collected through inertial impaction. As small particles flow through the scrubber, they pass close to the surfaces of the Tellerettes and scrubbing liquid droplets. The electrostatic charges on the particles cause them to be attracted to these neutral surfaces by image force attraction. All particles are eventually washed out of the
scrubber with the exit liquor. Noxious and malodorous gases are also absorbed and reacted in the same scrubbing liquor.

The IWS system utilizes Tellerette packing as its collection surface to achieve particulate removal. Scrubbing liquid droplets also act as collection surfaces. Particles of any size or composition are collected by the IWS. Fine particles (0.05–2 μm) are collected with high efficiency as well as coarse particles (2 μm and larger), regardless of their composition (organic or inorganic with either high or low resistivity).

Particle collection efficiency over long-term service remains consistently high. Interestingly, the percent of particulate removed varies little with load and particle size distribution over a wide range. As particulate load increases, the percent removed remains nearly constant. In addition, the collection efficiency for fine particles is nearly as great as for coarse particles. The IWS system also simultaneously absorbs gases. Noxious gases are removed through physical absorption and/or adsorption that is accompanied by chemical reaction. Pressure drop through a single-stage IWS is only 0.5–1.5 in. of H₂O. Energy for particle charging is low—approximately 0.2–0.4 kVA per 1000 cfm. The shell and most internal parts of the IWS are commonly fabricated of Duracor (fiberglass-reinforced plastic) and thermoplastic materials. This predominance of plastic construction assures corrosion-free operation in the presence of acid gases such as HCl, HF, Cl₂, NH₃, SO₂, and SO₃. For noncorrosive applications, metallic construction is also available. Factory-assembled modules, available in standard capacities from 900 to 54,000 acfm, can also be grouped together to handle virtually any gas volume.

With reference to the problem statement, calculate the overall pressure drop, ΔP, across the unit based on vendor data:

\[ \Delta P = (3)(1.55) \]
\[ = 4.65 \text{ in. H}_2\text{O} \]

Compare the above result with the actual pressure drop:

\[ \text{DIFF} = 4.75 - 4.65 = 0.10 \text{ in. H}_2\text{O} \]
\[ \%\text{DIFF} = (1.10/4.75)100 \]
\[ = 2.1\% = 0.021 \]

The operating particulate collection efficiency \( E \) of the unit is

\[ E = (1.23 - 0.017)/1.23 \]
\[ = 0.9862 = 98.62\% \]
Calculate the design collection efficiency based on design data:

\[ P = (1 - 0.74)^3 \]
\[ = (0.26)^3 \]
\[ = 0.0176 \]

\[ E = 1 - P \]
\[ = 1 - 0.0176 \]
\[ = 0.9824 = 98.24\% \]

The unit appears to be operating at an efficiency slightly above that indicated by the vendor.

**HYB.8 TWO-STAGE IWS SYSTEM**

Ceilcote has submitted the design for a two-stage IWS system for particulate control of a proposed hazardous waste incineration (HWI) facility that is early in the permit review process. The following data (estimated) have been provided for the facility:

- **Inlet loading** = 1.17 gr/dscf corrected to 7% oxygen
- **Outlet loading** = 0.015 gr/dscf corrected to 7% oxygen
- **Flue gas flowrate** = 20,000 acfm

*Note:* The federal particulate regulation (at the time of submission of the design) for HWI facilities is 0.08 gr/dscf corrected to 7% oxygen (or corrected to 50% excess air). State and/or local regulations can be lower; most state regulations are 0.03, but there are some states requiring that a 0.015 level be met. The reader is referred to *Introduction to Hazardous Waste Incineration* by Santoleri, Reynolds and Theodore (Wiley-Interscience, 2000) for additional details on these regulations and HWIs in general.

The following performance data are available on IWS systems from Ceilcote:

<table>
<thead>
<tr>
<th>Average Throughput Velocity (ft/s)</th>
<th>Collection Efficiency (%)</th>
</tr>
</thead>
<tbody>
<tr>
<td>8.0</td>
<td>75</td>
</tr>
<tr>
<td>7.0</td>
<td>80</td>
</tr>
<tr>
<td>6.0</td>
<td>85</td>
</tr>
<tr>
<td>5.0</td>
<td>90</td>
</tr>
<tr>
<td>4.0</td>
<td>94</td>
</tr>
</tbody>
</table>
The following equipment data are also provided:

<table>
<thead>
<tr>
<th>Unit</th>
<th>Frontal (face) Area (ft²)</th>
</tr>
</thead>
<tbody>
<tr>
<td>IWS-400</td>
<td>40</td>
</tr>
<tr>
<td>IWS-500</td>
<td>50</td>
</tr>
<tr>
<td>IWS-600</td>
<td>60</td>
</tr>
<tr>
<td>IWS-700</td>
<td>70</td>
</tr>
<tr>
<td>IWS-800</td>
<td>80</td>
</tr>
<tr>
<td>IWS-900</td>
<td>90</td>
</tr>
<tr>
<td>IWS-1000</td>
<td>100</td>
</tr>
</tbody>
</table>

Design (size) the IWS unit based on the information given above.

Solution

First, calculate a velocity that will provide a collection efficiency that will meet regulation specifications. Use the velocity to then determine the frontal area requirement, which will in turn effectively size the unit.

The required overall particulate collection efficiency $E_0$ is

$$E_0 = (1.17 - 0.015)/1.17 = 0.9872 = 98.72\%$$

For a two-stage unit, assuming equal penetrations,

$$1 - E_0 = P_1 P_2 = P^2$$

Thus,

$$P^2 = 1 - 0.9872 = 0.0128$$

$$P = 0.1131$$

and, the single-stage efficiency is

$$E = 1 - P$$

$$= 1 - 0.1131$$

$$= 0.8869 = 88.69\%$$
Using the vendor's velocity–efficiency data and linearly interpolating yields the following velocity:

\[ v = 5.262 \text{ ft/s} \]

The face area requirement of the unit, \( A \), in \( \text{ft}^2 \) is

\[ A = \frac{20,000}{(5.262)(60)} = 63.3 \text{ ft}^2 \]

Since the unit must meet or exceed the regulatory requirement, select the IWS-700. Because the size of the unit selected exceeds the required collection efficiency, calculate a revised efficiency and the corresponding throughput velocity:

\[ v = \frac{20,000}{(70)(60)} = 4.76 \text{ ft/s} \]

The corresponding single-stage efficiency (linearly interpolating) is

\[ E = 90 + (4.76 - 5.0)(94 - 90)/(4.0 - 5.0) \]

\[ = 90.96\% \]

The overall efficiency then becomes

\[ E_0 = 1 - (1 - 0.9096)^2 \]

\[ = 0.9918 = 99.18\% \]