Air Pollution Control: Carbon Adsorption for VOCs

Course No: C04-052
Credit: 4 PDH

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Chapter 1

Carbon Adsorbers

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October 2018

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1.1 Introduction

Adsorption is a non-destructive control technology employed to remove volatile organic compounds (VOCs) from low to medium concentration gas streams. Adsorption is generally used when a stringent outlet concentration must be met and/or recovery of the VOC is desired. Carbon adsorbers are used in a wide range of applications from controlling VOC and HAP emissions from storage tanks and process vents at refineries, chemical manufacturing, and pulp and paper plants, to control hydrogen sulfide and VOC emissions from municipal wastewater plants. Carbon adsorber canisters have been used in some smaller applications, such as in portable soil remediation systems. [28] In some industrial applications, adsorbers are used as concentrators to make other air pollution controls (e.g., condensers and thermal oxidizers) more cost-effective. Carbon adsorbers may also be used in conjunction with other less effective technologies (e.g., biofiltration or condensation) as a final polishing process to achieve VOC discharge limits. They are particularly useful for situations where there are relatively dilute VOC concentrations (less than 100 ppmv) and moderate flow rates, which can be difficult or uneconomical to remove using other types of pollution controls. [19] When properly designed, operated and maintained, carbon adsorbers can achieve high VOC removal efficiencies of 95 to 99 percent at input VOC concentrations of between 500 and 2,000 ppm in air. Removal efficiencies greater than 98 percent can be achieved for dilute waste streams. [20, 21]

Adsorption is the term used for the phenomenon where gas molecules passing through a bed of solid particles are selectively held on the surface of the solid by attractive forces which are weaker and less specific than those of chemical bonds. The term chemisorption is used where the gas molecules form actual chemical bonds with the adsorbent surface groups. Energy is released when a molecule from the gas stream adheres to the surface of the solid. This energy is known as the “heat of adsorption” and it typically equals or exceeds the heat of condensation. Adsorptive capacity of the solid for the gas tends to increase with the gas phase concentration, molecular weight, diffusivity, polarity, and boiling point. Most gases (“adsorbates”) can be removed (“desorbed”) from the adsorbent by increasing the temperature, decreasing the pressure or introducing a stronger adsorbed material to displace the VOCs. The methods used to regenerate adsorbent are described in more detail in Section 1.4.

Typical equipment life for carbon adsorbers is between 15 and 25 years. However, systems that handle waste gases that contain corrosive materials, such as hydrogen chloride or other acid gases, have shorter equipment life due to the impact of corrosion on the adsorber components. For example, waste streams that contain corrosive gases can corrode the adsorber vessel walls, carbon bed supports and outlet ducts. Corrosion of the bottom of the adsorber vessel is common where components of the gas stream condense to form corrosive liquids that collect on the bottom of the vessel. Corrosion of the carbon bed supports is also possible and may eventually cause the carbon beds to collapse. Carbon adsorbers used to control gas streams that contain corrosive materials should be constructed of materials that are designed for and resistant to corrosion.

While this chapter focuses primarily on VOC control, carbon adsorption is also used to control hazardous air pollutants (HAP), including inorganic HAPs such as hydrogen sulfide.

1.2 Types of Adsorbers

There are four types of adsorption equipment: (1) fixed regenerable beds; (2) disposable/ rechargeable canisters; (3) moving bed adsorbers; and (4) fluid-bed adsorbers. [2] Of these, the fixed-bed systems and canisters were the first developed and remain the most common systems in use today. [18]

1.2.1 Fixed-bed Units

Fixed-bed units can be sized for controlling continuous, VOC-containing streams over a wide range of flow rates, ranging from several hundred to several hundred thousand cubic feet per minute (cfm). The VOC concentration of streams that can be treated by fixed-bed adsorbers can be as low as several parts per billion by volume (ppbv) in the case of some toxic chemicals or as high as 25% of the VOCs' lower explosive limit (LEL). In some applications, the concentration of VOC in the waste gas stream may exceed the LEL. To reduce the risk of ignition, the waste gas stream is diluted to maintain the waste gas VOC concentration below 25% of LEL. (For most VOCs, the LEL ranges from 2,500 to 10,000 ppmv.[3])

Fixed-bed adsorbers may be operated in either intermittent or continuous modes. In intermittent operation, the adsorber removes VOC for a specified time (the "adsorption time"), which corresponds to the time during which the controlled source is emitting VOC. After the adsorber and the source are shut down (e.g., overnight), the unit begins the desorption cycle during which the captured VOC is removed from the carbon. This cycle, in turn, consists of three steps: (1) regeneration of the carbon by heating, generally by blowing steam through the bed in the direction opposite to the gas flow;[1] (2) drying of the bed, with compressed air or a fan; and (3) cooling the bed to its operating temperature via a fan. In most designs, the same fan is used both for bed drying and cooling. The length of the desorption cycle depends on several factors, including the characteristics of the contaminants in the waste stream, the type of adsorbent, and the regeneration method. The unit sits idle until the emission source starts operating again (for systems controlled by a single adsorber) or another adsorber is taken off-line for regeneration (for systems equipped with two or more adsorbers).

Fixed-bed adsorbers designed to operate continuously consist of two or more carbon beds where at least one regenerated carbon bed is available for adsorption at all times, thereby allowing the emissions source to operate continuously. In a system with two carbon beds, each bed must be large enough to handle the entire gas flow while adsorbing. Hence, twice as much carbon must be provided than an intermittent system handling the same flow. If the desorption cycle is significantly shorter than the adsorption cycle, it may be more economical to have three, four, or even more carbon adsorption beds operating at the same time. This approach reduces the amount of extra carbon capacity needed and can provide some additional benefits, such as maintaining a low VOC content in the effluent. (See Section 1.6 for information on designing a carbon adsorber control system, including how to determine the amount of carbon needed and how to estimate the adsorption and desorption times.)

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[1] Although steam is the most commonly used regenerant, there are situations where it should not be used. For example, steam regeneration should not be used for operations that emit halogenated VOCs, because the high temperature causes these VOCs to decompose.

A typical two-bed, continuously operated adsorber system is shown in Figure 1.1. One of the two beds is adsorbing at all times, while the other is desorbing or idled. As shown in Figure 1.1, the VOC-laden gas is collected by hoods or by direct pipeline connection and passes through a filter to remove particulates, before passing through the carbon bed in Bed #1 and exiting through the exhaust stack. Meanwhile, Bed #2 is in the desorption cycle in which steam flows through the bed and exits to a condenser. In the condenser, cooling water condenses the steam/VOC mixture. If part of the VOC is immiscible in water, the condensate can be sent to a decanter, where the VOC and water layers are separated and the VOC conveyed to storage. If the VOC collected contains a mixture of compounds, it may receive additional purification by distillation. The water layer is usually discharged to a wastewater treatment facility.

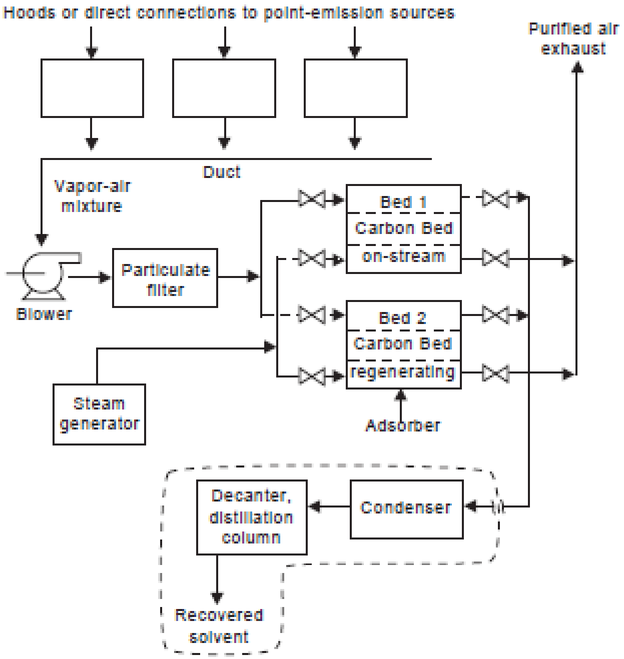

Figure 1.1: Typical-Two-Bed, Continuously Operated Fixed Bed Carbon Adsorber System

1.2.2 Canister Units

Canister-type adsorbers differ from fixed-bed units, in that they are normally limited to controlling lower-volume and intermittent gas streams, such as those emitted by storage tank vents, where process economics dictate that off-site regeneration is appropriate. The carbon canisters are not intended for in-situ desorption as in the fixed-bed units. Instead, they are either returned to the manufacturer or regenerated at a central desorption facility onsite.

Originally, canister adsorbers referred to relatively small returnable containers, such as 55-gallon drums. However, the term canister is becoming something of a misnomer as much of the growth in the industry is in larger vessels without regeneration capabilities. For example, one manufacturer supplies canister systems as large as 18,000 cfm and carbon capacities of 10,000 pounds. [25]

Once the carbon reaches a certain VOC content, the unit is shut down and either the carbon or the canister is replaced. The used carbon or the entire canister is then returned to a reclamation facility or regenerated at a central regeneration facility onsite. Several companies provide carbon replacement services for canisters. These companies analyze the spent carbon returned to them to determine whether it contains hazardous compounds. Fees for testing the spent carbon are typically between $800 and $1,000 based on 2018 prices, but fees may be higher depending on the application (e.g., wood treating and DBCP or PCB treatment).

Each canister unit consists of a vessel, activated carbon, inlet connection connected to distributer leading to the carbon bed, and an outlet connection to connect the unit to an exhaust stack. In one design (Calgon's Ventsorb®), 180 lbs of carbon are installed on an 8- inch gravel bed, in a 55-gallon drum with an internal collector. The type of carbon used depends on the composition of the VOC to be treated [6].

A single carbon canister may be used for emissions sources that operate intermittently or that can be shut down to allow replacement of a saturated carbon canister. However, most systems use two or more canisters, installed either in parallel or in series. Systems with canisters arranged in series are common. This design has two advantages: (1) any breakthrough that occurs in the first canister is controlled by the second canister; and (2) canisters can be replaced without disrupting the production process provided each canister is capable of controlling process emissions. When the first canister becomes saturated with VOC, the second canister becomes the primary carbon adsorber. When the carbon in the primary canister is saturated, the saturated canister is removed and a fresh canister added to the clean end. Periodic sampling for VOC breakthrough between the primary and secondary carbon canisters assures canister replacement occurs frequently enough to avoid VOC being emitted to the atmosphere. This approach also improves cost effectiveness of carbon replacement because the carbon canister is replaced at or near its saturation point. Although safer and more convenient, using two canisters in series is more expensive than systems using a single canister. In theory, a canister unit should remain in service longer than a fixed-bed regenerable unit operates in its adsorption cycle because new carbon adsorbent has a higher theoretical adsorption capacity than carbon regenerated in situ. The operating life of a carbon canister is expressed as the ratio of the theoretical capacity to the working capacity and is used to help ensure the allowable VOC concentration at the outlet is not exceeded. However, canisters sometimes remain connected until the carbon is near or at saturation because there is a cost incentive to operate each canister until the carbon is saturated. Also, unlike fixed-bed units whose outlet VOC concentrations are usually monitored continuously using flame ionization detectors, canisters are generally not monitored continuously. Adequate recordkeeping, periodic monitoring for VOC breakthrough, and bed life modeling provided by vendors are all worthwhile to ensure canister replacement occurs with sufficient frequency that VOC breakthrough does not occur. The primary indicator of the performance is the adsorber outlet VOC concentration. Other indicators of adsorber performance include inlet gas temperature, gas flow rate, inlet VOC concentration, pressure differential, inlet gas moisture content, and leak check monitoring.

1.2.3 Moving-Bed Adsorbers

Similar to fixed-bed adsorbers, moving-bed adsorbers consist of a permanent adsorber vessel in which the waste gases or vapors are brought into contact with the adsorbent. However, in the moving-bed adsorber, the spent adsorbent is continually regenerated. In a typical moving-bed adsorber, the adsorbent is held in two beds sited on coaxial rotating cylinders. VOC is adsorbed as the waste gas flows between the two rotating beds. As the cylinders are rotated, the portion of the adsorbent bed that is outside the waste gas stream is regenerated, thereby, increasing the potential for gas molecules to make contact with freshly regenerated adsorbent. [18]

1.2.4 Fluid-Bed Adsorbers

Fluid-bed adsorbers use a counter-flow design in which waste gas typically enters the adsorber vessel from the bottom and flows up through the vessel, while regenerated adsorbent enters from the top and slowly migrates to the bottom of the vessel. As adsorbent flows down through the vessel it gradually becomes saturated. The spent adsorbent is collected at the bottom of the vessel and transferred to a regeneration chamber, as illustrated in Figure 1.2. Because spent adsorbent is continually removed and fresh adsorbent added, fluid-bed adsorbers operate more efficiently and are less likely to experience the breakthrough issues associated with fixed-bed designs and canisters. [17, 18].

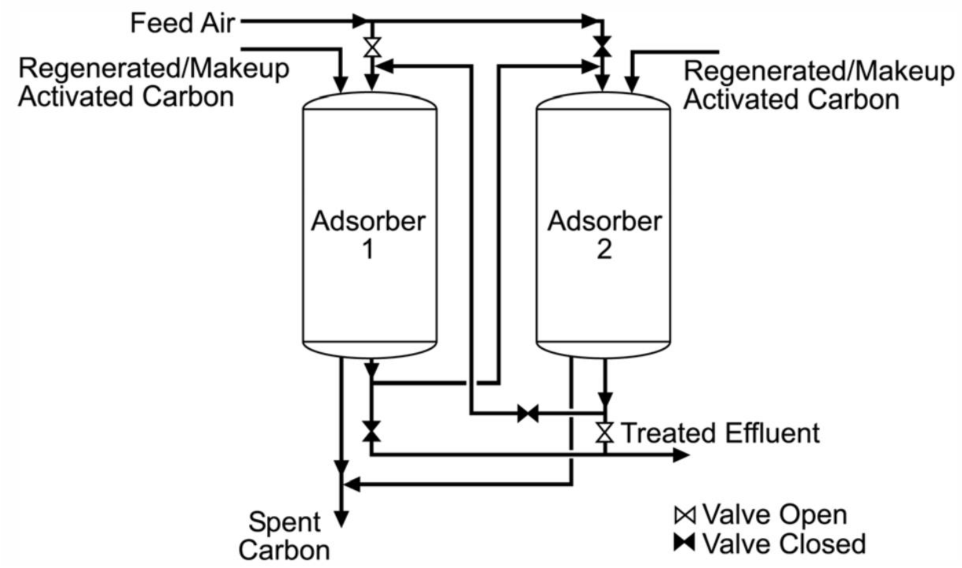

Figure 1.2: Schematic of a Fluid-Bed Adsorber [18]

1.3 Types of Adsorbent

There are three types of adsorbents in large scale use: activated carbon, synthetic zeolites and polymers. Each type of adsorbent has characteristics that make it effective for certain applications and impractical or uneconomic for others. A comparison showing the advantages and disadvantages of the three different types of adsorbents is provided in Table 1.1.

Activated Carbon:

Activated carbon was the first adsorbent used and remains the most commonly used adsorbent today. It is produced by high-temperature steam pyrolysis from a variety of materials, such as coal, wood and coconut husks. The resulting carbon has excellent adsorption characteristics due its high specific surface area (typically 800 to 1,400 square meters per gram) and the presence of small pores of various sizes (typically between 2 and 500 angstroms). The characteristics of the activated carbon produced vary depending on the type of material used. In general, activated carbon made from coconuts performs better at higher humidity than those made from coal. Activated carbon can also be modified to improve adsorption of other pollutants, such as hydrogen sulfide, mercury and ammonia. [18]

Activated carbon can adsorb a wide range of VOCs; however, there are some limitations. First, activated carbon is less effective for compounds that are highly polar, volatile or have small diameters. For example, vinyl chloride, methanol, and formaldehyde are not adsorbed well by activated carbon. Second, activated carbon is less effective in situations where the waste gas has high relative humidity as the water molecules readily adsorb to the activated carbon reducing the number of available absorption sites. [18] Some studies have shown a 30% reduction in removal efficiency when the relative humidity of the waste gas exceeds 75%. [20] Moisture in the bed can also promote biological growth on the carbon surface. [18] Third, carbon bed fires and explosions can occur with activated carbon adsorbers. Wastes with oxygen bearing compounds, such as peroxides, ketones, organic acids, aldehydes and organic sulfur compounds should be avoided as activated carbon can cause exothermic reactions with these compounds. The heat from exothermic reactions can ignite any flammable compounds present in the waste stream. In such cases, fire suppression measures can be used, such as increasing flow rate or using nitrogen or water to reduce the oxygen level. [18] Lastly, some VOCs may undergo chemical reactions to produce contaminants that are difficult to remove during desorption. For example, styrene monomers have been shown to polymerize to polystyrene making desorption possible only at very high temperatures. [18]

Activated carbon is generally regenerated by heating to temperatures of between 250 and 350°F. Steam is often used for this process, followed by a drying cycle to remove moisture. However, some molecules remain after regeneration. The adsorption capacity of regenerated carbon is typically about 50% of that for virgin material. The adsorption capacity declines during subsequent regeneration cycles, until eventually the activated carbon must be replaced. [21]

The typical cost for virgin activated carbon is between $1.90/lb and $6.50.[2] Costs for reactivated carbon are slightly lower and range from $0.95/lb to 1.55/lb.[3]

Zeolites:

Zeolites have a uniform crystalline structure with high specific surface areas and small, uniformly sized pores. Zeolites are sometimes called “molecular sieves” because their pores trap molecules of specific sizes, while allowing other molecules to pass through. For a given application, the zeolite must be carefully selected to match the molecular diameter of the VOCs to be removed.

Zeolites have specific surface areas that are similar to activated carbon. Naturally occurring zeolites include aluminosilicate minerals that are hydrophilic. However, synthetic zeolites are also available with characteristics tailored to a particular application. For example, some synthetic zeolites are hydrophobic with greater affinity for non-polar VOCs and low affinity for water molecules. Unlike activated carbon, zeolites are effective for removing VOC

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[2] Based on 2018 costs for virgin activated carbon provided by Carbtrol and Calgon Carbon Corporation.
[3] Based on 2018 costs for reactivated carbon provided by Calgon Carbon Corporation.

from waste streams with high relative humidity. As shown in the graph in Figure 1.3, the adsorption capacity of a hydrophobic zeolite is impacted only at very high relative humidity, while the adsorption capacity of activated carbon is significantly impacted at levels below 50%. [18]

Zeolites can be used to remove highly polar and volatile VOCs that are difficult to remove with activated carbon, including vinyl chloride, styrene, phenol, methane, acetone, methyl ethyl ketone, formaldehyde and sulfur compounds. Similar to activated carbon, reaction of the contaminants on the adsorbent can complicate removal during desorption. Zeolites are also not suitable for larger molecules as they are not adsorbed effectively. [18]

Zeolites can also be impregnated with other compounds to achieve targeted removal of certain compounds. For example, zeolites impregnated with potassium permanganate have been used to remove vinyl chloride from waste streams.

Zeolite is typically used to treat high-flow, low-concentration waste streams with VOC concentrations less than 150 ppmv and flow rates above 3,000 cfm. They have been used to control waste streams containing petroleum compounds and paint solvents. [18] Zeolites are regenerated by heating or vacuum. Since zeolites are able to withstand much higher temperatures than activated carbon, higher temperatures can be used thereby achieving higher levels of desorption. However, zeolites are more expensive than carbon with one report indicating zeolite costs can be twenty times that of activated carbon. [18]

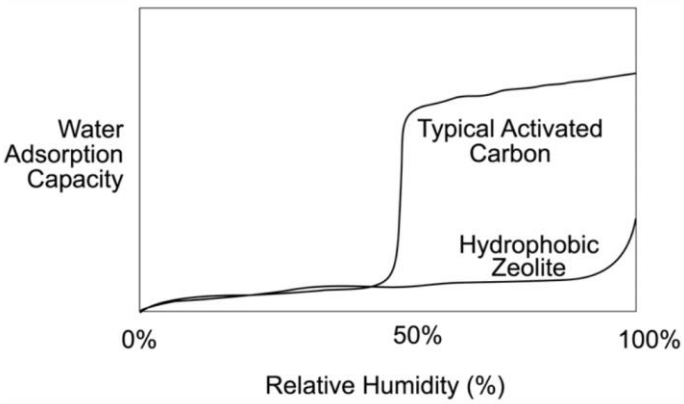

Figure 1.3: Effects of Humidity on Adsorption Capacity of Activated Carbon and Zeolite [18]

Polymers:

Polymer adsorbents are granules or beads made of synthetic polymers that are highly crosslinked to form a matrix of small pores and high surface areas. Polymer adsorbents can be used to control a wide range of VOCs, including styrene, toluene, xylenes, aldehydes, ketones, alcohols and chlorinated VOCs. However, polymers cannot be used with strong oxidizing agents. [18, 22, 23, 24]

The adsorption capacity for polymers is lower than that of zeolites but higher than for activated carbon. Unlike activated carbon adsorbers, they are less prone to fires and are capable of achieving removal efficiencies greater than 95%. [18] Polymers can be designed with a defined pore structure for selective removal of organic molecules. [23]

One advantage of polymeric adsorbents is their hydrophobic surfaces. This gives them a high tolerance for water vapor. Polymeric adsorbents can be used to efficiently treat VOC in waste streams with relative humidity greater than 90%. As a result, polymeric adsorbents have been used to control VOC emissions at soil remediation processes and industrial wastewater facilities, where the waste streams have high relative humidity. [18] Polymer adsorbents have been used to remove VOC/HAPs from soil vapor extraction (SVE) processes. One system demonstrated by Dow Chemical Company achieved 99% removal efficiency on a system with a flow rate of 271 ft^3/min, relative humidity of 98% and ambient temperature. The system consisted of two beds containing 191 lb of Dowex Optipore^{TM} polymeric adsorbent, which was regenerated onsite using conductive heating to 250°F and vacuum with low flow of carrier gas. Capital costs of the system were estimated to be less than $160,000 with annual costs of $2,732. [22] Figure 1.4 shows the water adsorption capacity of Dowex Optipore^{TM} compared to activated carbon.

Another advantage of polymers is that they can be used to adsorb reactive solvents without catalyzing their decomposition. Reactive solvents such as acetone, methyl ethyl ketone, cyclohexanone and styrene have been adsorbed and desorbed without measurable decomposition. Uncontrollable exothermic reactions that result in bed fires are also said to be less likely with polymer-based adsorbers than the carbon-based systems. [18, 23, 24]

Polymers are regenerated using heat, pressure and nitrogen purge gas. Desorption temperatures of up to about 250°F may be used with most polymers. Typically, a small quantity of cooling gas (e.g., air or nitrogen) is passed through the adsorber during regeneration. Regenerated polymers typically have lower adsorption capacity than the virgin polymer, but the decrease is typically less than observed with activated carbon. However, the reduction in the working capacity must be considered when designing polymer-based adsorbers as more polymer adsorbent will be needed to offset the decreased adsorption capacity. [18, 23, 24]

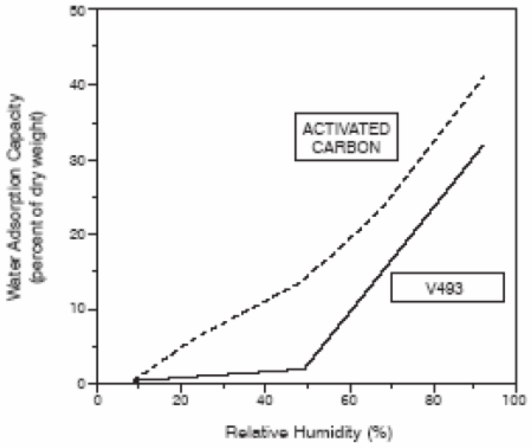

Figure 1.4: Effects of Humidity on Adsorption Capacity of Activated Carbon and Dowex Optipore^{TM} V493 Polymer [24]

Table 1.1: Comparison of Activated Carbon, Zeolite and Polymer Adsorbents

Absorbent Type	Characteristics	Advantages	Disadvantages
Activated Carbon	• Small pores of various sizes (typically between 2 and 500 angstroms). • High specific surface areas of between 800 and 1,400 square meters/gram.	• Lower cost. • Suitable for waste streams containing wide range of VOCs.	• Not effective for VOCs with high polarity (e.g., alcohols, organic acids). • Not effective for highly volatile compounds (e.g., vinyl chloride, MTBE) • Reduced capacity in high moisture applications. • High annual costs for carbon replacement/regeneration when used for concentrated waste streams. • Fire hazard if used with oxygen bearing compounds or VOCs having high heat of adsorption. • Degrades during desorption cycles.
Zeolites	• Uniform Crystalline structures with uniform pores. Pores are typically smaller than those in activated carbon. • Specific surface areas comparable to activated carbon.	• Highly selective VOC removal. • Good at removing small, highly polar and very volatile compounds. • Can be used in high humidity applications. • Longer life. • Less susceptible to fire. • Less susceptible to degradation	• Higher initial costs. • Not suitable for waste streams containing wide ranges of VOC.
Polymers	• Crosslinked polymers with high specific surface areas. Pores are generally larger than found in activated carbon.	• Can be used with reactive solvents without risk of catalyzing their decomposition. • Desorb more quickly than activated carbon. • Less susceptible to fire. • Longer life than activated carbon. • Lower affinity for water than activated carbon.	• Higher initial cost. • Regeneration temperatures cannot exceed 125ºC. • Not suitable for waste streams containing strong oxidizing agents.

The remainder of this chapter focuses primarily on the use of activated carbon, since this is the most commonly used adsorbent for VOCs.

1.4 Adsorbent Regeneration

Regeneration involves removing or desorbing the adsorbed molecules from the adsorbent and can be achieved by changing either the temperature or pressure. If the waste stream flow rate and composition are predictable, regeneration can occur at a preset time. However, if flow rate or the VOC concentration vary unpredictably, a sensor can be used to control when the regeneration cycle occurs. [18, 19]

Three types of regeneration systems are available: thermal swing regeneration, vacuum regeneration and pressure swing adsorption.

Thermal Swing Regeneration:

In systems using thermal swing regeneration, the temperature is increased (swung) from the ambient conditions under which VOC is adsorbed to higher temperatures for the regeneration cycle. Increasing the temperature, increases the kinetic energy of the VOC molecules. The molecules are desorbed when the kinetic energy of the VOC molecules increases sufficiently to overcome the forces holding molecules to the adsorbent. The temperature required depends on the VOC present, but is typically between 250 to 350^{o}F. Once the VOC is desorbed, cooled air is passed through the adsorber vessel to cool and dry the adsorbent before returning the vessel back to the adsorption cycle. Thermal regeneration has historically used steam, which provides the energy to desorb the adsorbate and also carries the desorbed compounds out of the adsorber vessel. The steam is then condensed and the VOC can either be recovered from the wastewater or the wastewater can be sent to wastewater treatment. Hence, steam-based regeneration systems will have additional equipment and operating costs associated with recovering the VOC and/or disposing of the wastewater generated. [19]

In addition to steam, other heat sources have been developed, including using microwaves, embedded heaters, and heated nitrogen. These alternative heat sources make VOC recovery easier and avoid the wastewater treatment costs associated with the steam-based systems. [19]

The physically adsorbed species in the smallest pores of the solid and the chemisorbed species may require rather high temperatures to be removed, and for all practical purposes cannot be desorbed during regeneration. Approximately 3 to 5 percent of organics adsorbed on virgin activated carbon is either chemisorbed or very strongly physically adsorbed and is difficult to desorb during regeneration. [1]

Vacuum Regeneration:

In a vacuum regeneration system, a vacuum pump is used to lower the pressure in the adsorber to below the vapor pressure of the adsorbed VOC. This causes the VOC to boil off from the adsorbent. Some systems use a small flow of purge gas (such as dry nitrogen) to transport of the VOC vapor to the recovery vessel. [19]

Pressure Swing Adsorption:

In a pressure swing adsorption system, the pressure is used to selectively adsorb and desorb individual VOCs based on their molecular characteristics and affinity for the adsorbent used. The method relies on the fact that more VOC is adsorbed at higher pressure. In these systems, the pressure is raised during the adsorption cycle and reduced during the regeneration cycle. When the adsorbent is saturated, the waste stream is stopped and the pressure of the adsorber vessel is reduced. The reduction in pressure results in desorption of the VOC, thereby regenerating the adsorbent. Pressure swing adsorption systems have been used to control gasoline vapors from storage tank vents. [19]

1.5 Adsorption Theory

At equilibrium, the quantity of gas that is adsorbed on activated carbon is a function of the adsorption temperature and pressure, the chemical species being adsorbed, and the carbon characteristics, such as carbon particle size and pore structure. For a given adsorbent-VOC combination at a given temperature, an adsorption isotherm can be constructed that shows the relationship between the mass of adsorbate per unit weight of adsorbent (i.e., the "equilibrium adsorptivity") to the partial pressure of the VOC in the gas stream.

Figure 1.5 shows a set of typical adsorption isotherms for VOC adsorption on activated carbon. As shown in the graph, the adsorptivity increases with increasing VOC partial pressure and decreases with increasing temperature.

Isotherms with this convex shape are designated as "Type I" isotherms. The Freundlich isotherm in equation 1.1 can be fit to a portion of a Type I curve and is commonly used in the industrial design of adsorbers.[2]

$$w_e = kP^m \tag{1.1}$$

where

w_e = equilibrium adsorptivity (lb adsorbate/lb adsorbent)
P = partial pressure of VOC in gas stream (psia)
k,m = empirical parameters

The treatment of adsorption from gas mixtures is complex and beyond the scope of this chapter. However, where two VOCs in a gas mixture have nearly identical adsorption isotherms, the VOC with the lower vapor pressure will displace the VOC with the higher vapor pressure even when the latter VOC has previously adsorbed to the carbon surface. Thus, during the course of the adsorption cycle the carbon's capacity for a higher vapor pressure constituent decreases. This phenomenon should be considered when sizing the adsorber. To be conservative, the adsorption cycle requirements should be based on the least adsorbable component in a mixture and the desorption cycle should be based on the most adsorbable component.[1]

The equilibrium adsorptivity is the maximum amount of adsorbate the carbon can hold at a given temperature and VOC partial pressure. In actual control systems where there are not two beds operating in series, however, the entire carbon bed is never allowed to reach equilibrium. Instead, once the outlet concentration reaches a preset limit (the "breakthrough concentration"), the adsorber is shut down for desorption or (in the case of canister units) replacement and disposal. At the point where the carbon adsorber is shut down, the average bed VOC concentration may only be 50% or less of the equilibrium concentration. That is, the carbon bed may be at equilibrium (“saturated”) at the gas inlet, but contain only a small quantity of VOC near the outlet.

As Equation 1.1 indicates, the Freundlich isotherm is a power function that plots as a straight line on log-log paper. Conveniently, for the concentrations/partial pressures normally encountered in carbon adsorber operation, most VOC-activated carbon adsorption conforms to Equation 1.1. At the very low concentrations typical of VOC breakthrough, a linear approximation (on arithmetic coordinates) to the Freundlich isotherm is adequate. However, the Freundlich isotherm does not accurately represent the isotherm at high gas concentrations and thus should be used with care as such concentrations are approached.

Adsorptivity data for selected VOCs were obtained from Calgon Corporation, a vendor of activated carbon. The vendor presents adsorptivity data in two forms: a set of graphs displaying equilibrium isotherms [7] and as a modification of the Dubinin-Radushkevich (D-R) equation, a semi-empirical equation that predicts the adsorptivity of a compound based on its adsorption potential and polarizability. [8] In this Manual, the modified D-R equation is referred to as the Calgon fifth-order polynomial. The data displayed in the Calgon graphs [7] has been fit to the Freundlich equation. The resulting Freundlich parameters are shown in Table 1.2 for a limited number of chemicals. The adsorbates listed include aromatics (e.g., benzene, toluene), chlorinated aliphatics (dichloroethane), and one ketone (acetone). However, the list is far from all-inclusive.

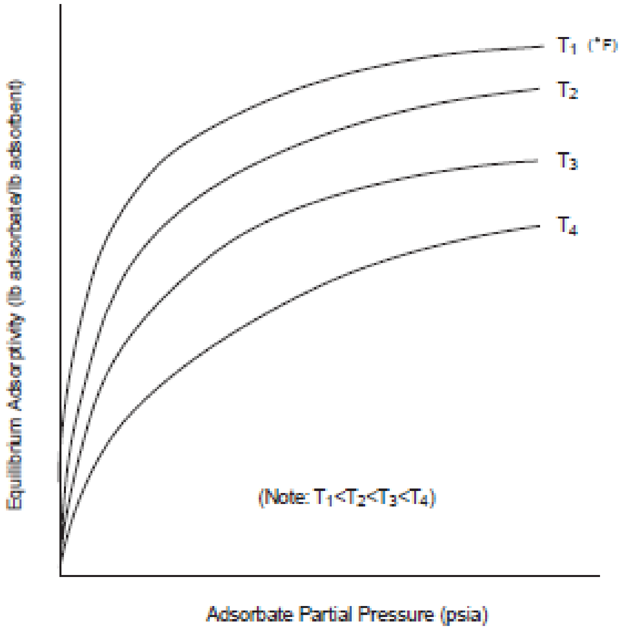

Figure 1.5: Type 1 Adsorption Isotherms for Hypothetical Adsorbate

Notice that a range of partial pressures is listed with each set of parameters, k and m. (Note: In one case (m-xylene) the isotherm was so curvilinear that it had to be split into two parts, each with a different set of parameters.) This is the range to which the parameters apply. Extrapolation beyond this range—especially at the high end—can introduce inaccuracy to the calculated adsorptivity.

But high-end extrapolation may not be necessary, as the following will show. In most air pollution control applications, the system pressure is approximately one atmosphere (14.696 psia). The upper end of the partial pressure ranges in Table 1.2 goes from 0.04 to 0.05 psia. According to Dalton's Law, at a total system pressure of one atmosphere this corresponds to an adsorbate concentration in the waste gas of 2,720 to 3,400 ppmv. As discussed in Section 1.2.1, the adsorbate concentration is usually kept at 25% of the lower explosive limit (LEL).[4] For many VOCs, the LEL ranges from 1 to 1.5 volume %, so that 25% of the LEL would be 0.25 to 0.375% or 2,500 to 3,750 ppmv, which approximates the high end of the partial pressure ranges in Table 1.2.

Finally, each set of parameters applies to a fixed adsorption temperature, ranging from 77°F to 104°F. These temperatures reflect typical operating conditions, although adsorption can take place as low as 32°F and even higher than 104°F. As the adsorption temperature increases to much higher levels, however, the equilibrium adsorptivity decreases to such an extent that VOC recovery by carbon adsorption may become economically impractical.

Table 1.2: Parameters for Selected Adsorption Isotherms [7][a]

Adsorbate	Adsorption Temp (°F)	Isotherm Parameters k	m	Range of Isotherm[b] (psia)
Benzene	77	0.597	0.176	0.0001-0.05
Chlorobenzene	77	1.05	0.188	0.0001-0.01
Cyclohexane	100	0.505	0.210	0.0001-0.05
Dichloroethane	77	0.976	0.281	0.0001-0.04
Phenol	104	0.855	0.153	0.0001-0.03
Trichloroethane	77	1.06	0.161	0.0001-0.04
Vinyl Chloride	100	0.200	0.477	0.0001-0.05
m-Xylene	77	0.708	0.113	0.0001-0.001
	77	0.527	0.0703	0.001-0.05
Acrylonitrile	100	0.935	0.424	0.0001-0.015
Acetone	100	0.412	0.389	0.0001-0.05
Toluene	77	0.551	0.110	0.001-0.05

[a] Each isotherm is of the form $w = kP^m$. (See text for definition of terms.) Data are for adsorption of Calgon type "BPL" carbon.

[b] Equation should not be extrapolated outside these ranges.

The Calgon fifth-order polynomial is somewhat more accurate than the Freundlich parameters from Table 1.2. The polynomial contains a temperature parameter, and it allows one to estimate adsorption isotherms for compounds not shown in Table 1.2 if pure component data are available. The pure component data required are the saturation pressure, liquid molar volume,

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[4] Although, Factory Mutual Insurance will reportedly permit operation at up to 50% of the LEL, if proper VOC monitoring is used.

and the refractive index. It is, however, somewhat more complex to use than the Freundlich equation. The Calgon fifth-order polynomial is as follows:

The mass loading, w_e, is calculated from

$$w_e = \frac{0.01G}{V_m} MW_{ads} \tag{1.2}$$

where

w_e = mass loading, i.e., equilibrium adsorptivity (g adsorbate per g carbon)[5]
G = carbon loading at equilibrium (cm^3 liquid adsorbate per 100 g carbon)
V_m = liquid molar volume of adsorbate (cm^3 per gmole)
MW_{ads} = molecular weight of adsorbate

Note that the terms in Equation 1.2 are given in metric units, not English. This has been done because the carbon loading, G, is calculated from a regression equation in which all the terms are expressed in metric units. This equation for G is the Calgon fifth-order polynomial:

$$\log_{10}(G) = A_0 + A_1Y + A_2Y^2 + A_3Y^3 + A_4Y^4 + A_5Y^5 \tag{1.3}$$

where

A_0 = 1.71
A_1 = -1.46×10^{-2}
A_2 = -1.65×10^{-3}
A_3 = -4.11×10^{-4}
A_4 = $+3.14 \times 10^{-5}$
A_5 = -6.75×10^{-7}

and Y is calculated from several equations which follow.

The first step in calculating Y is to calculate the parameter X using Equation 1.4:

$$\chi = \frac{\varepsilon}{(2.303\ RV_m)} \tag{1.4}$$

The adsorption potential, $\mathcal{E}$, is calculated using Equation 1.5:

$$\varepsilon = RT\ \ln\left(\frac{P_s}{P_i}\right) \tag{1.5}$$

where

R = 1.987 (calories per g-mole-K)
T = absolute temperature (K)
P_s = vapor pressure of adsorbate at the temperature T (kPa)
P_i = partial pressure of adsorbate (kPa)

---

[5] This, of course, is equal to *lb* adsorbate per *lb* carbon.

By substituting for $\mathcal{E}$ in the Equation 1.4, X can alternatively be calculated from[6]:

$$\chi = \left(\frac{T}{V_m}\right)\log 10\left(\frac{P_s}{P_i}\right) \tag{1.6}$$

The next step in calculating Y is to calculate the relative polarizability, Γ.

$$\Gamma = \frac{\Theta_i}{\Theta_o} \tag{1.7}$$

where

Θ_i = polarizability of component i per unit volume, where component i is the adsorbate
Θ_o = polarizability of component o per unit volume, where component o is the reference component, n-heptane.

$$\Theta = \frac{n^2 - 1}{n^2 + 2} \tag{1.8}$$

Once X and Γ are known, Y can be calculated from:

$$Y = \frac{\chi}{\Gamma} \tag{1.9}$$

Calgon also has a proprietary, seventh-order form in which two additional coefficients are added to the Calgon fifth-order polynomial, but the degree of fit reportedly is improved only modestly. [8] Additional sources of isotherm data include activated carbon vendors, handbooks (such as Perry's Chemical Engineer's Handbook), and the literature.

1.6 Design Procedure

1.6.1 Sizing Parameters

Data received from adsorber vendors indicate that the size and purchase cost of a fixed-bed or canister carbon adsorber system primarily depend on five parameters:

1. The volumetric flow of the VOC laden gas passing through the carbon bed(s);
2. The inlet and outlet *VOC* mass loadings of the gas stream;
3. The adsorption time (i.e., the time a carbon bed remains on-line to adsorb VOC before being taken off-line for desorption of the bed);

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[6] Alternatively, if the available values for T, P_i, P_s, and V_m are in English units, they may be substituted into this equation without conversion. However, to make the result dimensionally consistent with Equation 1.3, it be multiplied by the conversion factor, *34.7*.

4. The working capacity of the activated carbon in regenerative systems or the *equilibrium capacity* in the case of non-regenerative systems,

5. The moisture content of the gas stream.

In addition, the cost could also be affected by other stream conditions, such as the presence/absence of excessive amounts of particulate, moisture, or other substances that require the use of extensive pretreatment and/or corrosive-resistant construction materials. If the inlet concentrations are above 1,000 ppm, the moisture content does not significantly affect the working capacity of activated carbon. However, if the VOC inlet concentration is below 1,000 ppm or the relative humidity is above 50%, then the moisture begins to compete with the adsorbate for the available adsorption sites. If the VOC concentration is below 1,000 ppm or the waste gas has a relative humidity above 50%, then dehumidification equipment can be used to reduce the moisture content of the waste stream before it enters the adsorber vessel. Alternatively, a zeolite or synthetic polymer adsorbent may be used that has lower affinity for water molecules than activated carbon. Zeolite and synthetic polymer adsorbents are effective for waste streams with high moisture contents. Many zeolites and polymers are effective up to a relative humidity of 90%. [18, 21, 22, 23]

The purchased cost depends to a large extent on the volumetric flow (usually measured in actual ft^3/min). The flow, in turn, determines the size of the vessels housing the carbon, the capacities of the fan and motor needed to convey the waste gas through the system, and the diameter of the ducting.

Also important are the VOC inlet and outlet gas stream loadings, the adsorption time, and the working or equilibrium capacity of the carbon. These variables determine the amount and cost of carbon charged to the system initially and, in turn, the cost of replacing that carbon after it is exhausted (typically, five years after startup). Moreover, the amount of the carbon charge affects the size and cost of the auxiliary equipment (condenser, decanter, bed drying/cooling fan), because the sizes of these items are tied to the amount of VOC removed by the bed. The amount of carbon also has a bearing on the size and cost of the vessels.

A carbon adsorber vendor [9] supplied data that illustrate the dependency of the equipment cost on the amount of the carbon charge. Equipment costs were obtained for fixed-bed adsorbers sized to handle three gas flow rates ranging from 4,000 to 100,000 scfm and to treat inlet VOC (toluene) concentrations of 500 and 5,000 ppm. Each adsorber was assumed to have an eight-hour adsorption time. As one might expect, the equipment costs for units handling higher gas flow rates were higher than those handling lower gas flow rates. Likewise, at each of the gas flow rates, the units sized to treat the 5,000 ppm VOC streams had higher equipment costs than those sized to treat the 500 ppm concentration. These cost differences ranged from 23 to 29% and averaged 27%. These higher costs were partly needed to pay for the additional carbon required to treat the higher concentration streams. But some of these higher costs were also needed for enlarging the adsorber vessels to accommodate the additional carbon and for the added structural steel to support the larger vessels. Also, larger condensers, decanters, cooling water pumps, etc., were necessary to treat the more concentrated streams. (See Section 1.7 for procedures to estimate capital costs.)

The VOC inlet loading is set by the source parameters, while the outlet loading is set by the VOC emission limit. (For example, in many states, the average VOC outlet concentration from adsorbers may not exceed 25 ppm.)

1.6.2 Determining Adsorption and Desorption Times

The relative times for adsorption and desorption and the adsorber bed configuration (i.e., whether single or multiple and series or parallel adsorption beds are used) establish the adsorption/ desorption cycle profile. The cycle profile is important in determining carbon and vessel requirements and in establishing desorption auxiliary equipment and utility requirements. An example will illustrate. In the simplest case, an adsorber would be controlling a process which emits a relatively small amount of VOC intermittently—say, during one 8-hour shift per day. During the remaining 16 hours the system would either be desorbing or on standby. Properly sized, such a system would only require a single bed, which would contain enough carbon to treat eight hoursworth of gas flow at the specified inlet concentration, temperature, and pressure. Multiple beds, operating in parallel, would be needed to treat large gas flows (>100,000 actual ft^3/min, generally) [9], as there are practical limits to the sizes to which adsorber vessels can be built. But, regardless of whether a single bed or multiple beds were used, the system would only be on-line for part of the day.

However, if the process operates continuously (24 hours), an extra carbon bed would have to be installed to provide adsorptive capacity during the time the first bed is being regenerated. The amount of this extra capacity depends on the number of carbon beds that would be adsorbing at any one time, the length of the adsorption period relative to the desorption period, and whether the beds were operating in parallel or in series. If only one bed is adsorbing, a second would be needed to come on-line when the first is shut down for desorption. In this case, 100% extra capacity would be needed. Similarly, if five beds in parallel operate in a staggered adsorption cycle, only one extra bed would be needed and the extra capacity would be 20% (i.e., 1/5)—provided, of course, that the adsorption time is at least five times as long as the desorption time. The relationship between adsorption time, desorption time, and the required extra capacity can be generalized.

$$M_c = M_{cI}\, f \tag{1.10}$$

where

M_c, M_{cI} = amounts of carbon required for continuous or intermittent control of a given source, respectively (lbs)
f = extra capacity factor (dimensionless)

This equation shows the relationship between M_c, and M_{cI}. Section 1.6.3 shows how to calculate these quantities.

The factor, f, is related to the number of beds adsorbing (N_A) and desorbing (N_D) in a continuous system as follows:

$$f = 1 + \frac{N_D}{N_A} \tag{1.11}$$

(Note: N_A is also the number of beds in an intermittent system that would be adsorbing at any given time. The total number of beds in the system would be $N_A + N_D$)

It can be shown that the number of desorbing beds required in a continuous system (N_D) is related to the desorption time (θ_D), adsorption time (θ_A), and the number of adsorbing beds, as follows:

$$\theta_D \leq \theta_A \left(\frac{N_D}{N_A} \right) \tag{1.12}$$

θ_D is the total time needed for bed regeneration, drying, and cooling. For instance, for an eight-hour adsorption time, in a continuously operated system of seven beds (six adsorbing, one desorbing), θ_D would be equal to or less than 8 hours/6 beds or one hour and twenty minutes. If desorption takes longer than one hour and twenty minutes, then additional beds would be needed to provide sufficient extra capacity during desorption.

1.6.3 Estimating Carbon Requirement

1.6.3.1 Overview of Carbon Estimation Procedures

Obtaining the carbon requirement (M_c or M_{cI}) is not as straightforward as determining the other adsorber design parameters. When estimating the carbon charge, the sophistication of the approach used depends on the data and tools available.

For this Manual, we have adopted a rule-of-thumb procedure for estimating the carbon requirement. This procedure, while approximate in nature, appears to have the acceptance of vendors and field personnel. It is sometimes employed by adsorber vendors to make rough estimates of carbon requirement and is relatively simple and easy to use. It normally yields results incorporating a safety margin, the size of which depends on the bed depth (short beds would have less of a safety margin than deep beds), the effectiveness of regeneration, the particular adsorbate and the presence or absence of impurities in the stream being treated.

1.6.3.2 Carbon Estimation Procedure

The rule-of-thumb carbon estimation procedure is based on the “working capacity” (w_c, lb VOC/lb carbon). This is the difference per unit mass of carbon between the amount of VOC on the carbon at the end of the adsorption cycle and the amount remaining on the carbon at the end of the desorption cycle. It should not be confused with the “equilibrium capacity” (w_e,) defined above in Section 1.5. Recall that the equilibrium capacity measures the capacity of virgin activated carbon when the VOC has been in contact with it (at a constant temperature and partial pressure) long enough to reach equilibrium. In adsorber design, it would not be feasible to allow the bed to reach equilibrium. If it were, the outlet concentration would rapidly increase beyond the allowable outlet (or “breakthrough”) concentration until the outlet concentration reached the inlet concentration. During this period the adsorber would be violating the emission limit. With non-regenerable (canister) type systems, placing multiple vessels in a series can substantially decrease concerns of breakthrough.

The working capacity is some fraction of the equilibrium capacity. Like the equilibrium adsorptivity, the working capacity depends upon the temperature, the VOC partial pressure, and the VOC composition. The working capacity also depends on the flow rate and the carbon bed parameters.

The working capacity, along with the adsorption time and VOC inlet loading, is used to compute the carbon requirement for a canister adsorber or for an intermittently operated fixed-bed adsorber as follows:

$$M_{cI} = \frac{m_{voc}}{w_c}\theta_A \tag{1.13}$$

where

m_{VOC} = maximum VOC inlet loading (lb/h)

For waste streams where the VOC concentration is variable, the maximum VOC inlet loading should be used in order to ensure compliance with VOC emission limits at all inlet conditions.

Combining this with Equations 1.10 and 1.11 yields the general equation for estimating the system total carbon charge for a continuously operated system:

$$M_c = \frac{m_{voc}}{w_c}\theta_A\left(1 + \frac{N_D}{N_A}\right) \tag{1.14}$$

Values for w_c may be obtained from knowledge of operating units. If no value for w_c is available for the VOC (or VOC mixture) in question, the working capacity may be estimated at 50% of the equilibrium capacity, as follows:

$$w_c \approx 0.5 w_{e(\max)} \tag{1.15}$$

where

$w_{e(max)}$ = the equilibrium capacity (lb VOC/lb carbon) taken at the adsorber inlet (i.e., the point of maximum VOC concentration).

(Note: To be conservative (that is, to avoid understating working capacity), this 50% figure should be lowered if short desorption cycles, very high vapor pressure constituents, high moisture contents significant amounts of impurities, or difficult- to-desorb VOCs are involved. Furthermore, the presence of strongly adsorbed impurities in the inlet VOC stream may significantly shorten carbon life.)

As Equation 1.14 shows, the carbon requirement is directly proportional to the adsorption time. This would tend to indicate that a system could be designed with a shorter adsorption time to minimize the carbon requirement (and equipment cost). There is a trade-off here not readily apparent from Equation 1.14, however. Certainly, a shorter adsorption time would require less carbon. But, it would also mean that a carbon bed would have to be desorbed more frequently. This would mean that the regeneration steam would have to be supplied to the bed(s) more frequently to remove (in the long run) the same amount of VOC. Further, each time the bed is regenerated the steam supplied must heat the vessel and carbon, as well as drive off the adsorbed

VOC. And the bed must be dried and cooled after each desorption, regardless of the amount of VOC removed. Thus, if the bed is regenerated too frequently, the bed drying/cooling fan must operate more often, increasing its power consumption. Also, more frequent regeneration tends to shorten the carbon life. As a rule-of-thumb, the optimum regeneration frequency for fixed-bed adsorbers treating streams with moderate to high VOC inlet loadings is once every 8 to 12 hours.[1]

1.7 Estimating Total Capital Investment

This section describes the procedures that can be used to estimate the purchased costs for fixed-bed and canister-type carbon adsorbers. Since entirely different procedures should be used to estimate the purchased costs of fixed-bed and canister-type adsorbers, the two systems are discussed separately.

1.7.1 Fixed-Bed Systems

As indicated in the previous section, the purchased cost for fixed-bed carbon adsorbers using steam regeneration is a function of the volumetric flow rate, VOC inlet and outlet loadings, the adsorption time, and the working capacity of the activated carbon. As Figure 1.1 shows, the adsorber system is made up of several different items. Of these, the adsorber vessels and the carbon comprise from one-half to nearly 90% of the total equipment cost. (See Section 1.7.1.3.) There is also auxiliary equipment, such as fans, pumps, condensers, decanters, and internal piping, but because these usually comprise a small part of the total purchased cost, they may be "factored" from the costs of the carbon and vessels without introducing significant error into a cost estimate at a study-level of accuracy, as this one is. The costs of these major items are considered separately.

1.7.1.1 Carbon Cost

Carbon Cost, C_c, in dollars ($) is simply the product of the initial carbon requirement (M_c) and the current price of carbon.

$$C_c = CC \times M_c \tag{1.16}$$

where

C_c = Cost of carbon for the adsorber ($)
CC = Unit cost of carbon ($/lb)
M_c = Amount of carbon required (lbs)

As adsorber vendors buy carbon in very large quantities (million-pound lots or larger), their cost is somewhat lower than the list price. Vendors report typical carbons cost between $1.90 to $6.50 per pound for virgin carbon and $0.95 to $1.55 per pound for reactivated carbon Carbtrol reported costs of $1.90/lb for virgin carbon purchased in 1,100 lb sacks. [25, 29]

1.7.1.2 Vessel Cost

The cost of an adsorber vessel is primarily determined by its dimensions which, in turn, depend upon the amount of carbon it must hold and the superficial gas velocity through the bed that must be maintained for optimum adsorption. The desired superficial velocity is used to calculate the cross-sectional area of the bed perpendicular to the gas flow. An acceptable superficial velocity is established empirically, considering desired removal efficiency, the carbon particle size and bed porosity, and other factors. For example, one adsorber vendor recommends a superficial bed velocity of 85 ft/min[9], while an activated carbon manufacturer cautions against exceeding 60 ft/ min in systems operating at one atmosphere.[7] Another vendor uses a 65 ft/min superficial face velocity in sizing its adsorber vessels.[10] Lastly, there are practical limits to vessel dimensions which also influence their sizing. That is, due to shipping restrictions, vessel diameters rarely exceed 12 feet, while their length is generally limited to 50 feet.[10]

The cost of a vessel is usually correlated with its weight. However, as the weight is often difficult to obtain or calculate, the cost may be estimated from the external surface area. This is true because the vessel material cost—and the cost of fabricating that material—-is directly proportional to its surface area. The surface area (S, ft^2) of a vessel is a function of its length (L, ft) and diameter (D, ft), which in turn, depend upon the superficial bed face velocity, the L/D ratio, and other factors.

Most adsorber vessels are cylindrical in shape and may be erected either vertically or horizontally (as in Figure 1.1). For horizontal vessels, the carbon volume occupies no more than 1/3 of the vessel volume [9, 10]. It can be shown that this constraint limits the bed depth to no more than:

$$Maximum\ \ bed\ \ depth \approx \frac{\pi\, D}{12} \tag{1.17}$$

The vessel length, L, and diameter, D, can be estimated by solving two relationships, namely, (1) the equation relating carbon volume, and thus vessel volume, to L and D, and (2) the equation relating volumetric flow rate, superficial velocity, and cross-section normal to flow. If one assumes that the carbon bulk density is 30 lb/ft^3, then one can show that:

$$D = \frac{0.127\, M_c^{'}\, v_b}{Q'} \tag{1.18}$$

$$L = \frac{7.87}{M_c^{'}} \left(\frac{Q'}{v_b} \right)^2 \tag{1.19}$$

where

D = vessel diameter (ft)
L = vessel length (ft)
V_b = bed superficial velocity (ft/min)
M_c' = carbon requirement *per vessel* (lbs)
Q' = volumetric flow rate *per adsorbing vessel* (acfm)

Because the constants in Equations 1.18 and 1.19 are not dimensionless, one must be careful to use the units specified in these equations.

Although other design considerations can result in different values of L and D, these equations result in L and D which are acceptable from the standpoint of “study” cost estimation for horizontal, cylindrical vessels which are larger than 2-3 feet in diameter.

The carbon requirement and flow rate for each adsorber vessel can be calculated as follows.

$$M_c' = \frac{M_c}{(N_A + N_D)} \qquad Q' = \frac{Q}{N_A} \tag{1.20}$$

At gas flow rates (Q') of less than 9,000 scfm, it is usually more feasible to erect the adsorber vessels vertically instead of horizontally.[10] If so, the vessel diameter can be calculated from the volumetric flow rate per adsorbing vessel and the bed superficial velocity as follows:

$$D = \left(\frac{4Q'}{\pi v_b}\right)^{1/2} \tag{1.21}$$

The vertical vessel length will depend principally on the carbon bed thickness. Additional space must be included below the carbon bed for bed support and above and below the bed for distribution and disengaging of the gas stream and for physical access to the carbon bed. In smaller diameter vessels, access to both sides of the bed is usually not required. However, 1 to 1½ feet must be provided on each side for gas distribution and disengagement, or 2 to 3 feet overall. For longer vessels, 2 to 3 feet at each end of the vessel is typically provided for access space.

Given the mass of carbon in the bed, the carbon bulk density, and the bed diameter (i.e., the cross-sectional area normal to flow), determining the carbon bed thickness is straight forward using the following equation:

$$t_b = \frac{\text{volume of carbon}}{\text{cross}-\text{sectional area normal to flow}} = \frac{\left(\dfrac{M_c'}{\rho_b}\right)}{\left(\dfrac{Q'}{v_b}\right)} \tag{1.22}$$

where

ρ_b = carbon bulk density (lb/ft^3, assume 30 lb/ft^3)

The vessel length is, therefore,

$$L = t_b + t_{a,g} \tag{1.23}$$

where

$t_{a,g}$ = access/gas distribution allowance
= 2 to 6 feet (depending on vertical vessel diameter)

Finally, use the following equation to calculate the surface area of either a horizontal or vertical vessel:

$$S = \pi D(L + D/2) \tag{1.24}$$

Similar equations can be developed for other vessel shapes, configurations, etc.

Based on vendor data, we developed a correlation between adsorber vessel cost and surface area: [10]

$$C_v = 271 \times F_m \times S^{0.778} \tag{1.25}$$

where

C_v = vessel cost (fall 1999 \$), F.O.B. vender[7]
F_m = adjustment factor for fabrication material (from Table 1.3)
S = surface area of the vessels (ft^2)

Table 1.3: Adjustment Factors to Obtain Costs for Fabricated Material

Material	F_m Factor	Reference(s)
Stainless steel, 304	1.0	[10]
Stainless steel, 316	1.3	[9,10,11]
Carpenter 20 CB-3	1.9	[11]
Monel-400	2.3	[9,11]
Nickel-200	3.2	[11]
Titanium	4.5	[11]

Equation 1.25 is valid for carbon adsorption vessels with surface areas (S) within the following range:

$$97 \le S \le 2110 \; ft^2$$

---

[7] Two vendors provided information for the 1999 updates, neither felt that modifications to the capital costs of adsorber system between 1989 and 1999 were appropriate. The major change for 1999 was a decrease in the price of carbon.[4, 5]

304 stainless steel is the most common material used in fabricating adsorber vessels. [9, 10]

1.7.1.3 Total Purchased Cost

As stated earlier, the costs of such items as the fans, pumps, condenser, decanter, instrumentation, and internal piping can be factored from the sum of the costs for the carbon and vessels. Based on four data points derived from costs supplied by an equipment vendor [10], we found that, depending on the total gas flow rate (*Q*), the ratio (R_c) of the total adsorber equipment cost to the cost of the vessels and carbon ranged from 1.14 to 2.24. These data points spanned a gas flow rate range of approximately 4,000 to 500,000 acfm. The following regression formula fit these four points:

$$R_c = 5.82\,Q^{-0.133} \tag{1.26}$$

where

Q is in the range of 4,000 to 500,000 acfm
Correlation coefficient (r) = 0.872

The total adsorber equipment cost (C_A) is the product of R_c and the sum of the carbon and vessel costs, as follows:

$$C_A = R_c\left[C_c + C_v\left(N_A + N_D\right)\right] \tag{1.27}$$

1.7.1.4 Total Capital Investment

As discussed in Section 1, in the methodology used in this Manual, the total capital investment (TCI) is estimated from the total purchased cost via an overall direct/indirect installation cost factor. A breakdown of that factor for carbon adsorbers is shown in Table 1.4. As Section 1.2 indicates, the TCI also includes costs for land, working capital, and off-site facilities, which are not included in the direct/indirect installation factor. However, as these items are rarely required with adsorber systems, they will not be considered here. Further, no factors have been provided for site preparation (SP) and buildings (Bldg.), as these site-specific costs depend very little on the purchased equipment cost.

The installation factor is applied to the total purchased equipment cost, which includes the costs of auxiliary equipment (i.e., the stack, external ductwork, etc) and the freight costs and sales taxes (if applicable). The external ductwork consists of the ducting needed to convey the exhaust gas from the emission source to the adsorber system, and then from the adsorber to the stack. (Costs for ductwork and stacks are shown elsewhere in this Manual.) Normally, the adjustment would also cover the instrumentation cost, but this cost is usually included with the adsorber equipment cost. Finally, note that these factors reflect “average” installation conditions and could vary considerably, depending upon the installation circumstances.

Table 1.4: Capital Cost Factors for Carbon Adsorbers [12]

Direct Costs	
Purchased equipment costs	
Adsorber + auxiliary equipment[a]	As estimated, A
Instrumentation[b]	0.10 A
Sales taxes	0.03 A
Freight	0.05 A
Purchased equipment cost, PEC	B = 1.18 A
Direct installation costs	
Foundations & supports	0.08 B
Handling & erection	0.14 B
Electrical	0.04 B
Piping	0.02 B
Insulation	0.01 B
Painting	0.01 B
Direct installation costs	0.30 B
Site preparation	As required, SP
Buildings	As required, Bldg
Total Direct Costs, DC	1.30 B + SP + Bldg.
Indirect Costs (installation)	
Engineering	0.10 B
Construction and field expenses	0.05 B
Start-up	0.02 B
Performance test	0.01 B
Total Indirect Costs, IC	0.18 B
Contractor Fees	0.10(DC + IC)
Contingencies, C	CF(DC + IC)
Total Capital Investment	1.48 B + SP + Bldg + C + Contractor Fees

[a] Ductwork and any other equipment normally not included with unit furnished by adsorber vendor.

[b] Instrumentation and controls often furnished with the adsorber, and thus included in the EC.

[c] Where “CF” is the contingency factor. Typical values for CF for mature technologies such as carbon adsorbers range from 5 to 15 percent.

1.7.2 Canister Systems

Once the carbon requirement is estimated using the procedure in Section 1.6.3, the number of canisters is determined. This is done simply by dividing the total carbon requirement (M_c) by the amount of carbon contained in each canister. This quotient, rounded to the next highest digit, yields the required number of canisters to control the vent in question. Costs for typical canisters of various sizes are listed in Table 1.5. These costs include the vessel, carbon, and connections, but do not include taxes, freight, or installation charges. The cost per canister generally decreases as the quantity of canisters purchased increases. Costs are for canisters containing granular or pelletized activated carbon, commonly used in industrial adsorption applications (e.g., Calgon’s AP4-60). However, to treat certain VOCs, more expensive specialty adsorbents are needed, such as Calgon’s HGR-P - an impregnated pellet activated carbon, which is designed for mercury removal in gas phase applications. These adsorbents can increase the equipment cost by 60% or more.[6] Costs for activated carbon range from \$0.95 to \$6.50 per pound based on 2018 data. Prices vary depending on mesh, activity and type.

The current trend is toward the use of larger non-regenerable fixed-bed canisters with capacities of 1,000 to 10,000 pounds, where the carbon is typically exchanged in the field. These adsorbers are usually atmospheric designs made of thin steel with an internal coating to inhibit corrosion. Typical prices for these large canisters are provided in Table 1.6. Annual maintenance costs are reported to range from 3% to 10% of the installed capital costs.

Table 1.5: Equipment Costs for Typical Canister Adsorbers [25, 29]

Canister Size (in lb of Activated Carbon)	Maximum Flow Rate (cfm)	Canister Type	Cost with Virgin Carbon (\$)*	Cost with Reactivated Carbon (\$)*
140	500	Polyethylene	720	-
140	500	Epoxy-Lined Steel	1,105	-
170	300	Epoxy-Lined Steel	1,090	-
180	100	Epoxy-Lined Steel	1,600	980
200	100	Epoxy-Lined Steel	785	-

* Costs are in 2018 dollars and exclude taxes and freight charges.

Table 1.6: Equipment Costs for Typical Large Canister Adsorbers [25, 29]

Adsorber Size (in lb of Activated Carbon)	Maximum Flow Rate (cfm)	Canister Type	Cost with Virgin Carbon (\$)*	Cost with Reactivated Carbon (\$)*
1,000	600	Epoxy-Lined Steel	6,600	-
1,000	1,000	Epoxy-Lined Steel	11,500	7,000
2,000	2,000	Epoxy-Lined Steel	19,000	10,000
2,000	750	Carbon Steel	22,000	13,200
3,000	2,000	Epoxy-Lined Steel	13,900	-
4,100	8,000	Polypropylene	45,000	-
5,000	2,500	Carbon Steel	42,600	20,100
8,000	4,500	Carbon Steel	66,000	30,000
10,000	18,000	Polypropylene	94,500	-

* Costs are in 2018 dollars and exclude installation, taxes and freight charges.

As fewer installation materials and labor are required to install a canister unit than a fixed-bed system, the composite installation factor is consequently lower. The only costs required are those needed to place the canisters at, and connect them to, the source. This involves a small amount of piping only; little or no electrical work, painting, foundations, or the like would be needed. For typical sites, twenty percent of the sum of the canister(s) cost, freight charges, and applicable sales taxes covers the installation cost. However, installation costs may be higher depending on site-specific conditions. For example, canister units retrofitted at congested sites may require additional duct work, blowers, and valves.

The number and size of canisters can be adjusted to accommodate certain design specifications. For example, the number of canisters calculated using the approach outlined above must be doubled for systems where two canisters are used in series. Similarly, for a system designed to minimize the frequency of canister replacement, canisters with larger capacity than estimated using the approach outlined above should be selected.

1.8 Estimating Total Annual Cost

The total annual cost for a carbon adsorption system is comprised of three components: direct costs, indirect costs, and recovery credits. Each component is considered separately in this section.

1.8.1 Direct Annual Costs

Direct annual costs include the following expenditures: steam, cooling water, electricity, carbon replacement, operating and supervisor labor, and maintenance labor and materials. Of these, only operator/supervisor labor, maintenance and materials labor, electricity and solid waste disposal or carbon replacement/regeneration apply to the canister-type adsorbers.

1.8.1.1 Steam

As explained in Section 1.1, steam is used during the desorption cycle. The quantity of steam required will depend on the amount of carbon in the vessel, the vessel dimensions, the type and amount of VOC adsorbed, and other variables. Experience has shown that the steam requirement ranges from approximately 3 to 4 lbs of steam/lb of adsorbed VOC. [9, 10] Using the midpoint of this range, the annual costs for steam can be estimated by the equation:

$$C_s = 3.50\ m_{voc}\ \theta_s\ p_s \tag{1.28}$$

where

C_s = steam cost (\$/yr)
θ_s = system operating hours (h/yr)
m_{voc} = VOC inlet loading (lbs/hr)
p_s = steam price (\$/thousand lbs)

If steam price data are unavailable, one can estimate its cost at 130% of the fuel cost for oil and natural gas-fired boilers. For example, if the natural gas price for an industrial facility is \$4.00/million BTU [8], then the estimated steam price would be \$5.20/million BTU or approximately \$5.00/thousand lbs of steam. The 1.3 factor covers the costs of producing the steam, including costs for water, water treatment (e.g., clarification, softening, demineralization), feedwater pumping, combustion air fan power, sewer charges for boiler blowdown, air pollution controls for the boiler, maintenance materials and maintenance labor. [26]

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[8] Average price of natural gas in 2017 for industrial facilities, based on data published by the U.S. Department of Energy, Energy Information Administration (see https://www.eia.gov/dnav/ng/ng_pri_sum_dcu_nus_a.htm).

1.8.1.2 Cooling Water

Cooling water is consumed by the condenser in which the steam-VOC mixture leaving the desorbed carbon bed is condensed. Most of the condenser duty is comprised of the latent heat of vaporization (ΔH_v) of the steam and VOC. As the VOC ΔH_v are usually small compared to the steam ΔH_v, (about 1000 BTU/lb), the VOC ΔH_v may be ignored. So may the sensible heat of cooling the water-VOC condensate from the condenser inlet temperature (about 212°F) to the outlet temperature. Therefore, the cooling water requirement is essentially a function of the steam usage and the allowable temperature rise in the coolant, which is typically 30 to 40°F.[9] Using the average temperature rise (35°F), we can write:

$$C_{cw} = 3.43 \frac{C_s}{P_s} p_{cw} \tag{1.29}$$

Where

C_{cw} = cooling water cost (\$/yr)
p_{cw} = cooling water price (\$/thousand gal.)

If the cooling water price is unavailable, use \$1.00 to \$8.25/thousand gallons.[9]

1.8.1.3 Electricity

In fixed-bed adsorbers, electricity is consumed by the system fan, bed drying/cooling fan, cooling water pump, and solvent pump(s). Both the system and bed fans must be sized to overcome the pressure drop through the carbon beds. But, while the system fan must continuously convey the total gas flow through the system, the bed cooling fan is only used during a part of the desorption cycle (one-half hour or less).

For both fans, the horsepower needed depends both on the gas flow and the pressure drop through the carbon bed. The pressure drop through the bed (ΔP_b) depends on several variables, such as the adsorption temperature, bed velocity, bed characteristics (e.g., void fraction), and thickness. But, for a given temperature and carbon, the pressure drop per unit thickness depends solely on the gas velocity. For instance, for Calgon's "PCB" carbon (4 x 10 mesh), the following relationship holds: [7]

$$\frac{\Delta P_b}{t_b} = 0.03679 \ v_b + 1.107 \times 10^{-4} \ v_b{}^2 \tag{1.30}$$

where

$\Delta P_b/t_b$ = pressure drop through bed (inches of water/foot of carbon)
v_b = superficial bed velocity (ft/min)

---

[9] Based on rates for industrial facilities report in 2013 compiled by Black & Veatch. (see 2012/2013 "50 Largest Cities Water/Wastewater Rate Survey." Available at http://www.saws.org/who_we_are/community/RAC/docs/2014/50-largest-cities-brochure-water-wastewater-rate-survey.pdf.

As Equation 1.22 shows, the bed thickness (t_b, ft) is the quotient of the bed volume (V_b) and the bed cross-sectional area (A_b). For a 30 lb/ft^3 carbon bed density, this becomes

$$t_b = \frac{V_b}{A_b} = \frac{0.0333\, M_c'}{A_b} \tag{1.31}$$

(For vertically erected vessels, $A_b = Q/V_b$, while for horizontally erected cylindrical vessels, $A = LD$.) Once ΔP_b is known, the system fan horsepower requirement (hp_{sf}) can be calculated:

$$hp_{sf} = 2.50 \times 10^{-4} Q\ \Delta P_s \tag{1.32}$$

where

Q = gas volumetric flow through system (acfm)
ΔP_s = total system pressure drop = $\Delta P_b + 1$

(The extra inch accounts for miscellaneous pressure losses through the external ductwork and other parts of the system.[9][10] However, if extra long duct runs and/or preconditioning equipment are needed, the miscellaneous losses could be much higher.)

This equation incorporates a fan efficiency of 70% and a motor efficiency of 90%, or 63% overall.

The horsepower requirement for the bed drying/cooling fan (hp_{cf}) is computed similarly. While the bed fan pressure drop would still be ΔP_b, the gas flow and operating times would be different. For typical adsorber operating conditions, the drying/cooling air requirement would be 50 to 150 ft^3/lb carbon, depending on the bed moisture content, required temperature drop, and other factors. The operating time (θ_{cf}) would be the product of the drying/cooling time per desorption cycle and the number of cycles per year. It can be shown that:

$$\theta_{cf} = 0.4\, \theta_D \left(\frac{N_A \theta_s}{\theta_A} \right) \tag{1.33}$$

(The "0.4" allows for the fact that as a rule-of-thumb, approximately 40% of the desorption cycle is used for bed drying/cooling.)

The cooling water pump horsepower requirement (hp_{cwp}) would be computed as follows:

$$hp_{cwp} = \frac{2.52 \times 10^{-4} q_{cw}\ H\ s}{\eta} \tag{1.34}$$

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[10] To obtain a more precise estimate of ductwork pressure drop, refer to Section 2 of this Manual.

where

q_{cw} = cooling water flow (gal/min)
H = required head (nominally 100 feet of water)
s = specific gravity of fluid relative to water at 60°F
η = combined pump-motor efficiency.

The annual operating hours for the cooling water pump (θ_{cwp}) would be computed using Equation 1.33, after substituting "0.6" for 0.4. The 0.6 factor accounts for the fact that the cooling water pump is only used during the steaming portion of the regeneration, while the condenser is in operation.

Equation 1.34 may also be used to compute the solvent pump horsepower requirement. In the latter case, the flow (q_s) would be different of course, although the same head—100 ft. of water—could be used. The specific gravity would depend on the composition and temperature of the condensed solvent. For example, the specific gravity of toluene at 100°F would be approximately 0.86 at 70°F. (However, the solvent pump horsepower is usually very small—usually < 0.1 hp.— so its electricity consumption can usually be neglected.)

Once the various horsepowers are calculated, the electricity usage (in kWh) is calculated, by multiplying each horsepower value by 0.746 (the factor for converting hp to kilowatts) and the number of hours each fan or pump operates annually. For the system fan, the hours would be the annual operating hours for the system (θ_s). But, as discussed above, the operating times for the bed drying/cooling fan and cooling water pump would be different.

To obtain the annual electricity cost, simply multiply kWh by the electricity price (in \$/kWh) that applies to the facility being controlled.

For canister units, use Equation 1.32 to calculate the fan horsepower requirement. However, instead of P_b use the following to compute the total canister pressure drop P_c inches of water:[6]

$$\Delta P_c = 0.0471\ Q_c + 9.29 \times 10^{-4}\ Q_c^2 \qquad (1.35)$$

where

Q_c = flow through the canister (acfm).

1.8.1.4 Carbon Replacement

As discussed above, the carbon has a different economic life than the rest of the adsorber system. Therefore, its replacement cost must be calculated separately. Employing the procedure detailed in Section 1, Chapter 2 of this Control Cost Manual, we have:

$$CRC_c = CRF_c \left(1.08\, C_c + C_{cl}\right) \qquad (1.36)$$

where

CFR_c = capital recovery factor for the carbon
1.08 = taxes and freight factor
C_c = cost of carbon, \$
C_{cl} = labor cost for carbon replacement, \$

The replacement labor cost covers the labor cost for removing spent carbon from vessels and replacing it with virgin or regenerated carbon. The cost would vary with the amount of carbon being replaced, the labor rates, and other factors, but can be estimated using equation 1.37. For example, to remove and replace a 50,000 pound carbon charge would require about 16 person-days. [13] Hence, a typical carbon replacement rate is 379 lbs of carbon per hour.

$$C_{cl} = \frac{LR \times M_c}{CRR} \tag{1.37}$$

where

C_{cl} = labor cost for carbon replacement, \$
CRR = carbon replacement rate, lbs of carbon/hour (default value, 379 lbs/hour)
LR = labor rate, \$/hour
M_c = quantity of carbon replaced, lbs

A typical life for the carbon is five years. However, if the inlet contains VOCs that are very difficult to desorb, tend to polymerize, or react with other constituents, a shorter carbon lifetime of one or two years is likely. [1]

The capital recovery factor (CFR_c) is calculated using the following equation:

$$CRF_c = \frac{i\ (1+i)^n}{(1+i)^n - 1} \tag{1.38}$$

Where *i* is the assumed interest rate and *n* is the expected life of the carbon. For a five-year life and 5% interest rate, CRF_c is 0.2310.

1.8.1.5 Solid Waste Disposal

Disposal costs are rarely incurred with fixed-bed adsorbers, because the carbon is almost always regenerated in place, not discarded. The carbon in canister units should also be regenerated in most cases. For larger vessels, common practice is for a carbon vendor to pick up the spent carbon and replace it with fresh carbon. The spent carbon is then returned to a central facility for regeneration. The EPA encourages both solvent recovery and reuse of spent carbon as pollution prevention and waste minimization techniques.

In some cases, the solvent characteristics, such as their toxicity or the difficulty in desorbing them from the carbon, may make landfill disposal the preferred option. In these cases, an entire canister—carbon, drum, connections, etc.— may be shipped to a secure landfill. The cost of landfill disposal could vary considerably, depending on the number of canisters disposed of, the location of the landfill, etc. Based on data obtained from two large landfills, disposal costs range from approximately \$35 to \$65 per canister, excluding transportation costs.[14, 15]

1.8.1.6 Operating and Supervisory Labor

The operating labor for adsorbers is relatively low, as most systems are automated and require little attention. One-half operator hour per shift is typical.[12] The annual labor cost is the product of this labor requirement and the operating labor wage rate ($/h) which, naturally, varies based on the facility location, type of industry, etc. Add to this 15% to cover supervisory labor, as suggested in Section 1, Chapter 2 of the Control Cost Manual.

1.8.1.7 Maintenance Labor and Materials

Use 0.5 hours/shift for maintenance labor [12] and the applicable maintenance wage rate. If the latter data are unavailable, estimate the maintenance wage rate at 110% of the operating labor rate, as Section 1 suggests. Finally, for maintenance materials, add an amount equal to the maintenance labor, also per Section 1, Chapter 2 of the Manual.

1.8.2 Indirect Annual Costs

These include such costs as capital recovery, property taxes, insurance, overhead, and administrative costs ("G&A"). The capital recovery cost is based on the equipment lifetime and the annual interest rate. (See Section 1.2 for a thorough discussion of the capital recovery cost and the variables that determine it.) For adsorbers, the equipment lifetime is typically 15 to 25 years, except for the carbon, which, as stated above, typically needs to be replaced after five years. Therefore, when figuring the system capital recovery cost, one should base it on the installed capital cost less the cost of replacing the carbon (i.e., the carbon cost plus the cost of labor necessary to replace it). Substituting the initial carbon and replacement labor costs from Equation 1.36, we obtain:

$$CRC_s = \left[TCI - \left(1.08C_c + C_{cl}\right)\right]CRF_s \tag{1.39}$$

where

CRC_s = capital recovery cost for adsorber system ($/yr)
TCI = total capital investment ($)
1.08 = taxes and freight factor
C_c, C_{cl} = initial carbon cost (F.O.B. vendor) and carbon replacement labor cost, respectively ($)
CRF_s = capital recovery factor for adsorber system (defined in Section 1.2).

For a 15-year life and a 5% annual interest rate, the CRF_s is 0.0963.

As Section 1.2 indicates, the suggested factor to use for property taxes, insurance, and administrative charges is 4% of the TCI. Finally, the overhead is calculated as 60% of the sum of operating, supervisory, and maintenance labor, and maintenance materials.

The above procedure applies to canister units as well, except in those cases where the entire unit and not just the carbon is replaced. The piping and ducting cost can usually be considered a capital investment with a useful life of ten years. However, whether the canister itself would be treated as a capital or an operating expense would depend on the particular application and would need to be evaluated on a case-by-case basis.

1.8.3 Recovery Credits and Disposal Costs

During the desorption cycle, VOC is desorbed, condensed, and separated from the steam condensate. The recovered VOC can be re-used onsite (e.g., as a solvent or burned as a fuel), sold to recyclers, or sent to a disposal site. If the recovered VOC is sufficiently pure, it can be reused onsite or sold. As the example problem in Section 1.9 illustrates, if the quantity of recovered VOC is large enough, its value can offset the annual costs of the control device, resulting in a net annual credit. However, the current market price of the VOC and its purity impact the size of the credit. The greater the purity, the higher the value. However, if the recovered VOC contains impurities or is a mixture of compounds, further treatment, such as distillation, will be required. Purification and separation costs are beyond the scope of this chapter. Needless to say, the costs of these operations can be significant and should be included in estimates of the capital and operating costs. They may offset the revenues generated by the sale of the VOC. Where the facility is unable to re-use or sell the waste VOC collected, the facility may incur costs associated with its off-site disposal. Costs of disposal for wastes containing hazardous pollutants, such as benzene, can be high and alternative control systems, such as incinerators, may be more cost effective than a carbon adsorber.

Recovery Credits:

The following equation can be used to calculate the recovery credits:

$$RC = m_{voc}\ \theta_s\ p_{voc}\ E \tag{1.40a}$$

where

RC = recovery credit ($/yr)
m_{voc} = VOC inlet loading (lbs/h)
θ_s = system operating hours (h/yr)
p_{voc} = resale value of the recovered VOC ($/lb)
E = adsorber VOC control efficiency

By definition, the efficiency (E) is the difference between the inlet and outlet VOC mass loading, divided by the inlet loading. However, during an adsorption cycle the outlet VOC loading will increase from essentially zero at the start of the cycle to the breakthrough concentration at the end of the cycle. Because the efficiency is a function of time, it should be calculated via integration over the length of the adsorption cycle. However, this approach requires knowledge of the temporal variation of the outlet loading during the adsorption cycle. If this knowledge is not available, a conservative approximation of the efficiency may be made by setting the outlet loading equal to the breakthrough concentration.

Disposal Costs:

In situations where the collected VOC must be sent off-site for disposal, the following equation should be used to calculate the disposal costs:

$$Disposal_{Cost} = m_{voc}\ \theta_s\ D_{voc}\ E \tag{1.40b}$$

where

$Disposal_{Cost}$ = Disposal Cost ($/yr)
m_{voc} = VOC inlet loading (lbs/h)
θ_s = System operating hours (h/yr)
D_{voc} = Disposal costs for the recovered VOC ($/lb)
E = Adsorber VOC control efficiency

1.8.4 Total Annual Cost

Finally, as explained in Section 1, the total annual cost (TAC) is the sum of the direct and indirect annual costs and disposal cost, less any recovery credits, or:

$$\text{TAC} = \text{DAC} + \text{IAC} + \text{Disposal}_{\text{cost}} - \text{RC} \qquad (1.41)$$

where

DAC = Direct annual costs,
IAC = Indirect annual costs,
$\text{Disposal}_{\text{cost}}$ = Costs for disposing or otherwise treating recovered VOC, and
RC = Recovery credit.

1.8.5 Cost Effectiveness

The cost in dollars per ton of VOC removed per year, is calculated using equation 1.42:

$$Cost\ Effectiveness = \frac{TAC}{VOC\ Removed/yr} \qquad (1.42)$$

where

Cost Effectiveness = the cost effectiveness, $/ton
VOC Removed/year = annual mass of VOC removed by the carbon adsorber, tons/year

1.9 Example Problem

An example problem that calculates both the design parameters and capital and annual costs for a carbon adsorber applied to a printing plant is presented below. The printing plant will operate continuously for 360 days/year and emit 100 lb/hour of toluene. The waste gas is expected to contain negligible quantities of particulate matter and moisture. The plant proposes to use a carbon adsorber that consists of three beds, with two carbon beds adsorbing VOC, while the third bed is desorbing or on standby. Each carbon bed will be housed in a metal vessel made with 304 stainless steel. No additional site preparation or building modifications are necessary for the installation of the new adsorber. The toluene will be recovered and sold to a solvent recycling company.

The following assumptions are made to perform the calculations:

VOC emitted = Toluene
Required control efficiency = 98%
Emission rate (M_{voc}) = 100 lbs/hour
Toluene concentration at inlet = 710 ppm
Partial pressure of toluene = 0.0104 psia (at atmospheric pressure and 77^{o}F)
Number of operating hours = 8,640 hours/year
Total time for adsorption (Θ_A) = 12 hours
Time available for desorption = 5 hours
Total waste gas flow rate (Q) = 10,000 acfm (at atmospheric pressure and 77^{o}F)

Assume the recovered toluene is recycled at the source.

Superficial bed velocity (v_b) = 75 ft/min
Expected life of the absorber vessels and auxiliary equipment = 15 years
Expected life of carbon = 5 years

In addition to these assumptions, other important inputs to the cost calculations are:

Carbon cost (CC) = $4.20 per lb[11]
Electricity = $0.0676 per kWh[12]
Steam = $5.00 per 1,000 lbs of steam[13]
Cooling water = $3.55 per 1,000 gallons of water[14]
Operator labor rate = $27.48 per hour[15]
Re-sale price for recovered toluene = $0.33 per lb[16]

1.9.1 Design Parameter Example

Carbon Working Capacity: At the stated flow and pollutant loading, the toluene inlet concentration is 710 ppm. This corresponds to a partial pressure of 0.0104 psia. Substituting the partial pressure and the toluene isotherm parameters from Table 1.2 into Equation 1.1, we calculate an equilibrium capacity of 0.333 lb VOC/lb carbon.

$$w_e = k\,P^m = (0.551)(0.0104)^{0.11} = 0.333\,lbVOC\ /\,lbCarbon$$

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[11] Median cost for virgin carbon based on 2018 prices provided by Calgon Carbon Corporation (Moon Township, PA) and Carbtrol Corporation (Bridgeport, CT).
[12] Average electricity price for the industrial sector is based data compiled by the Department of Energy, Energy Information Administration, November 6, 2017. Available at https://www.eia.gov/electricity/sales_revenue_price/.
[13] Price of steam was estimated by multiplying the average price of natural gas for industrial facilities ($4.00/MMBtu) by 1.3 and converting to $/1,000 lb of steam. Natural gas price is the 2017 average price for industrial facilities published by the U.S. Department of Energy, Energy Information Administration (see https://www.eia.gov/dnav/ng/ng_pri_sum_dcu_nus_a.htm).
[14] Average water rates for industrial facilities in 2013, compiled by Black & Veatch. (see 2012/2013 "50 Largest Cities Water/Wastewater Rate Survey." Available at http://www.saws.org/who_we_are/community/RAC/docs/2014/50-largest-cities-brochure-water-wastewater-rate-survey.pdf.
[15] Bureau of Labor Statistics, May 2017 National Occupational Employment and Wage Estimates – United States, May 2017 (https://www.bls.gov/oes/current/oes_nat.htm). Hourly rates for operators based on data for plant and System Operators – other (51-8099).
[16] For this example, recovered toluene is valued at one-half the December 2017 market price of $717.64/metric ton.[16]

By applying the rule-of-thumb discussed in Section 1.6.3.2, the working capacity may be estimated at 50% of the equilibrium capacity, as follows:

$$w_c \approx 0.5 w_{c(\max)} \approx 0.167\, lbVOC\ / lbCarbon$$

where

$w_{c(max)}$ = Equilibrium capacity at the adsorber inlet (lb VOC/lb carbon)

Carbon Requirement: When operating, the adsorber will have two beds adsorbing and one bed regenerating/in stand-by at all times. Using equation 1.12 we can estimate the amount of the desorption time (Θ_D):

$$\theta_D \le \theta_A \left(\frac{N_D}{N_A} \right) = 12\, hours \times \left(\frac{1}{2} \right) = 6\, hours$$

where

N_D = Number of desorbing beds
N_A = Number of adsorbing beds
Θ_A = Total time for adsorption (hours)
Θ_D = Total time needed for bed regeneration, drying, and cooling (hours)

Because the stated design desorption time of 5 hours is less than the 6 hours needed for bed regeneration, the proposed bed configuration is feasible.

Equation 1.14 is used to calculate the amount of carbon required (M_c):

$$M_c = \frac{m_{voc}}{w_e} \theta_A \left(1 + \frac{N_D}{N_A} \right) = \left(\frac{100\, lbVOC\ / hour \times 12\, hours}{0.167\, lbVOC\ / lbCarbon} \right) \left(1 + \frac{1}{2} \right) = 10{,}800\, lbs$$

Adsorber Vessel Dimensions: Assuming the vessels will be erected horizontally, the vessel diameter (D), length (L), and surface area (S) are calculated using Equations 1.18, 1.19, and 1.24, respectively.

$$D = \frac{0.127\, M_c' v_b}{Q'} = \frac{0.127\,(3{,}600)(75)}{5{,}000} = 6.86\ ft$$

$$L = \frac{7.87}{M_c'} \left(\frac{Q'}{v_b} \right)^2 = \frac{7.87}{3{,}600} \left(\frac{5{,}000}{75} \right)^2 = 9.72\ ft$$

$$S = \pi D \left(L + \frac{D}{2} \right) = \pi (6.86\ ft) \left(9.72\ ft + \frac{6.86\ ft}{2} \right) = 283\ ft$$

Where the volumetric flow rate (Q') and quantity of carbon required (M_c') for each bed are calculated as follows:

$$Q' = \left(\frac{Q}{N_A}\right) = \left(\frac{10{,}000\,acfm}{2}\right) = 5{,}000\,acfm\ /\,bed$$

$$M_c' = \frac{M_c}{(N_A + N_D)} = \frac{10{,}800\,lbs}{(2+1)} = 3{,}600\,lbs\,/\,bed$$

1.9.2 Cost Estimate Example

Once the carbon adsorption system is sized, the total cost of the system can be estimated as the sum of the total capital investment (TCI) and the total annual costs.

Total Capital Investment: The total capital investment (TCI) costs are estimated as

$$TCI = \sum Direct\ Capital\ Costs + \sum Indirect\ Capital\ Costs$$

Table 1.7 shows the direct and indirect capital costs for the carbon adsorption system based on the factors provided in Section 1.8.

Adsorber Equipment Cost (including costs for carbon and instrumentation): The equipment cost is comprised of the adsorber vessels, carbon, and the condenser, decanter, fan, pumps and other equipment usually included in the adsorber price. The costs of the latter items are based on the combined cost of the vessels and carbon. Combining Equations 1.26 and 1.27, we have:

$$C_A = 5.82\ Q^{-0.133}\left[C_c + \left(N_A + N_D\right)C_v\right] \tag{1.43}$$

Because S falls between 97 and 2,110 ft^2, equation 1.25 can be used to calculate the cost per vessel, C_v. Additionally, since each vessel will be constructed from 304 stainless steel, the adjustment factor for fabrication material (F_m) in Equation 1.25 is 1, and the cost per vessel is calculated as follows:

$$C_v = 271\,F_m\,S^{0.778} = (271)(1)(283)^{0.778} = \$21{,}900\,/\,vessel$$

The cost per vessel, Cv, is then converted to the current year dollars using the Chemical Engineering Plant Cost Index (CEPCI). In this example, the scaling factor for 2017 dollars was 567.5/390.6). Hence, Cv is $31,834 in 2017 dollars.

The total cost of the carbon required for the adsorber (C_c) is calculated using equation 1.16 as follows:

$$C_c = CC \times M_c = \frac{\$4.2}{lb} \times 10{,}800\,lbs = \$45{,}360$$

Substituting the values for C_v and C_c into equation 1.43 yields:

$$C_A = (5.82)(10{,}000)^{-0.133}[45{,}360 + (2+1)(31{,}834)] = \$240{,}800$$

Auxiliary Equipment Costs: Auxiliary equipment includes the costs for the stack and duct work and dampers connecting the coating booth to the carbon adsorber and the carbon adsorber to the stack. The costs for this equipment are estimated using the methods provided in Section 2 of the Manual. For the purposes of this example, assume the costs for the auxiliary equipment have been estimated to be:

Ductwork	$16,500
Dampers	7,200
Stack	8,500
Total	$32,200

Table 1.7: Estimated Total Capital Cost for the Carbon Adsorber

Cost Item	Estimated Cost
Direct Costs	
Purchased equipment costs	
Adsorber vessels and carbon	$240,805
Auxiliary equipment	32,200
Sum = A	$273,005
Instrumentation, 0.1 A[a]	$0
Sales taxes, 0.03 A	$8,190
Freight, 0.05 A	$13,650
Purchased equipment cost (B)	$294,845
Direct installation costs	
Foundations & supports, 0.08 B	$23,588
Handling & erection, 0.14 B	$41,278
Electrical, 0.04 B	$11,794
Piping, 0.02 B	$5,897
Insulation for ductwork, 0.01 B	$2,948
Painting, 0.01 B	$2,948
Direct installation costs	$88,454
Site preparation	$0
Buildings	$0
Total Direct Costs (DC)	**$383,454**
Indirect Costs (installation)	
Engineering, 0.10 B	$29,485
Construction and field expenses, 0.05 B	$14,742
Start-up, 0.02 B	$5,897
Performance test, 0.01 B	$2,948
Total Indirect Cost(IC)	**$53,072**
Contractor Fees, 0.10(DC + IC)	**$43,637**
Contingencies (C), CF(DC + IC + Contractor Fees)[c]	**$48,001**
Total Capital Investment (TCI)[b]	**$528,000**

[a] The costs for instrumentation are included in the adsorber equipment cost.

[b] Value shown has been rounded to three significant figures.

[c] Where "CF" is the contingency factor of 10%, the midpoint of the typical range of 5 to 15% for mature technologies such as this one.

Instrumentation costs are included in the cost of the adsorber. Sales taxes and freight costs are based on the sum of the total equipment costs (i.e., adsorber vessels, carbon, and auxiliary equipment). The sales taxes are assumed for purposes of this cost example to be 3 percent of the equipment costs,[17] while the freight charges are assumed for purposes of this cost example to be 5 percent of the equipment costs. Hence, the purchased equipment cost (B) is given by the following equation:

$$Purchased\ Equipment\ Cost\ (B) = 1.08 \times A = 1.08 \times (\$240{,}805 + \$32{,}200) = \$294{,}845$$

where

A = Sum of adsorber vessels, carbon, and auxiliary equipment costs ($).

Other Direct and Indirect Costs: Table 1.7 shows the calculations for the other direct and indirect costs for the carbon adsorber. These costs are calculated using the factors provided in Section 1.8 and the purchased equipment cost (B) calculated above.

Total Capital Investment: The total capital investment (TCI) is the sum of all the direct and indirect costs and the contingency factor (10% for this example) and is calculated as follows:

$$\text{Total Capital Investment (TCI)} = DC + IC + C + \text{Contractor Fees}$$

$$TCI = 1.48 \times B + CF(1.48 \times B) + 0.1(1.48 \times B) = \$528{,}000$$

1.9.3 Total Annual Costs

Annual Costs, Table 1.8 gives the direct and indirect annual costs for the carbon adsorber system, as calculated from the factors in Section 1.8.

As discussed in Section 1.8.4, the total annual cost (TAC) is comprised of the direct annual costs (DAC), indirect annual costs (IAC), annual disposal costs ($Disposal_{cost}$) and any recovery credits (RC), as described by the equation:

$$TAC = DAC + IAC + Disposal_{Cost} - RC$$

The total direct annual cost (DAC) is the sum of the operator labor, maintenance, carbon replacement, and utility costs.

Operator Labor Costs: The operator costs are the sum of the operator and supervisor costs. As explained in Section 1.8.1.6, the annual operator labor is estimated to be one-half hour per shift. The annual labor cost is calculated as follows:

$$Operator\ \ Labor\ \ Cost = \frac{0.5\ hours}{shift} \times \frac{3\ shifts}{day} \times \frac{360\ \ days}{year} \times \frac{\$27.48}{hour} = \$14{,}839\ /\ year$$

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[17] In some states, pollution control equipment is exempt from sales taxes. This should be accounted for in the TCI estimate if relevant.

The costs for supervisor labor is estimated at 15% of the operator labor cost, as shown below:

$$Supervisor\ \ Labor\ \ Cost\ = 0.15 \times \$14{,}839\ /\ year\ = \$2{,}226\ /\ year$$

Maintenance Costs: The maintenance costs are the sum of the costs of labor and materials. The maintenance labor is estimated to be 0.5 hours/shift. Since the maintenance labor rate is unknown, we estimate the wage rate at 110% of the operating labor rate, as recommended in Section 1 of the Cost Manual. The maintenance labor costs are calculated as follows:

$$Maintenance\ Labor\ Cost = \frac{0.5\ hours}{shift} \times \frac{3\ shifts}{day} \times \frac{360\ days}{year} \times \frac{\$27.48}{hour} \times 1.10 = \$16{,}323$$

Finally, for the maintenance materials, we estimated the materials to be an amount equal to the estimated maintenance labor.

Carbon Replacement Costs: As discussed in Section 1.8.1.4, costs for replacing the carbon are the sum of the labor and the replacement carbon. Because the economic life of the carbon is 5 years, the carbon replacement costs must be calculated separately from the rest of the adsorber system using a capital recovery factor based on 5 years, instead of the 15-year life expected for the absorber vessels and auxiliary equipment. Hence, the costs of replacing the carbon is calculated using equation 1.36.

$$CRC_c = CRF_c\left(1.08\,C_c + C_{cl}\right)$$

Where CRF_c, calculated as follows using a 5% interest rate and 5-year carbon life, is given by:

$$CRF = \frac{\mathrm{i}(1+i)^n}{(1+i)^n - 1} = \frac{(0.05)(1+0.05)^5}{(1+0.05)^5} = 0.2310$$

The labor costs for removing the spent carbon from vessels and replacing it with virgin or regenerated carbon is calculated using equation 1.37. For this example, we used the default factor of 379 lbs/hour for the carbon replacement rate (*CRR*) and labor rate of $30.23/hour (110% of the operator rate):

$$C_{cl} = \frac{LR \times M_C}{CRR} = \frac{(\$30.23 \times (10{,}800 lbs)}{379 lbs/hour} = \$861$$

Hence, the total costs for carbon replacement are calculated as follows using the values of CRF_c, C_c and C_{cl} calculated above:

$$CRC_c = CFR_c\left(1.08C_c + C_{cl}\right) = 0.2310\left[(1.08)(\$45{,}360) + \$861\right] = \$11{,}515$$

Utility Costs: Electricity includes the power for the system fan, bed drying/cooling fan, and the cooling water pump. Since the solvent pump motor is normally very small, its power consumption is negligible and may be neglected.

Electricity consumed by the system fan is calculated using Equation 1.32, the operating hours per year and the conversion factor from hp to kW:

$$kWh_{sf} = \frac{0.746\,kW}{hp} \times 2.50 \times 10^{-4}\, Q\, \Delta P_s\, \theta_S$$

where

$$\Delta P_s\left(inches\ \ water\right) = \Delta P_b + 1 = t_b\left(0.03679\ v_b + 1.107 \times 10^{-4} v_b^2\right) + 1$$

(The latter expression was derived from Equation 1.30, assuming that the carbon used in this example system is Calgon's "PCB," 4 x 10 mesh size.)

By assuming a carbon bed density, of 30 lb/ft^3, Equation 1.31 can be used to calculate the bed thickness (t_b):

$$Bed\ Thickness\ = t_b = \frac{0.0333\,M'_c}{A_b} = 1.8\ ft$$

Thus:

$$\Delta P_b = 1 + 1.80\left(0.03679 \times 75 + 1.107 \times 10^{-4} \times 75^2\right) = 7.09\ inches$$

And finally:

$$kWh_{sf} = 0.746 \times 2.5 \times 10^{-4} \times 7.09\,in. \times 10{,}000\ \ acfm\ \times 8{,}640\,h/yr\ = 114{,}200\,kWh/yr$$

Bed drying/cooling fan: During the drying/cooling cycle, the pressure drop through the bed also equals P_b. However, as Section 1.6.1.3 indicates, the flow and operating time are different. For the air flow, take the midpoint of the range (100 ft^3 air/lb carbon) and divide by 2 hours (the bed drying/cooling time), yielding: 100 ft^3/lb x 3,600 lbs x 1/120 min = 3,000 acfm. Substituting this into Equation 1.32 results in:

$$2.50 \times 10^{-4} \times 7.09\ inches\ \ \times 3{,}000\ acfm\ = 5.32\ hp$$

From Equation 1.33, we get:

$$\theta_{cf} = 0.4 \times 5h \times 2 \times \frac{8{,}640\,h}{12\,h} = 2{,}880\,h$$

Thus:

$$kWh_{cf} = \frac{0.746\,kW}{hp} \times 5.32\,hp \times 2{,}880\,h = 11{,}430\,kWh/yr$$

Cooling water pump: The cooling water pump horsepower is calculated from Equation 1.34. Here, let η = 63% and H = 100 ft. The cooling water flow (q_{cw}) is the quotient of the annual cooling water requirement and the annual pump operating time. From the data in Table 1.8, we obtain the cooling water requirement: 10,400,000 gal/yr. The pump annual operating time is obtained from Equation 1.33 (substituting 0.6 for 0.4), or θ_{cwp} = (0.6)(5 h)(2)(8,640)/12 = 4,320 h/yr.

Thus:

$$hp_{cwp} = \frac{\left(2.52\times10^{-4}\right)\left(100\ ft\right)}{0.63}\times\frac{10,400,000\,gal/yr}{\frac{4,320\,h}{yr}\times60\ \min/\ hr} = 1.60\,hp$$

And:

$$kWh_{cwp} = 0.746\,kW/hp\times1.60\,hp\times4,320\,h/yr = 5,160\,kWh/yr$$

Summing the individual power consumptions, we get the value shown in Table 1.8: 131,000 kWh/yr.

Recovery Credit: In this example, we have included a credit for the recovery and re-sale of toluene. The quantity of toluene recovered is estimated from the toluene emission rate (100 lbs/hour), the number of operating hours (8,640 hours/year), and the control efficiency of the carbon adsorber (98%).

Equation 1.39 is used to calculate the recovery credit:

$$RC = m_{voc}\,\theta_s\,p_{voc}\,E = \left(\frac{100lbs}{hr}\right)(8640hrs)\left(\frac{\$0.33}{lb}\right)(0.98) = \$279{,}000$$

Hence, a recovery credit of $279,000 can be taken for the recovery and re-sale of 423 tons of recovered toluene. Since all of the recovered toluene can be sold, no disposal costs will be incurred. Hence, for this example the $Disposal_{cost}$ equals zero.

Total Annual Cost: The sum of the direct annual costs, indirect annual costs and annual disposal costs, less the toluene recovery credit, yields a net total annual credit of $60,400. Clearly, total annual cost is very sensitive to the amount and value of the recovered toluene. For instance, if the market price for toluene was $0.15/lb, then the recovery credit would be $127,000 and the total annual cost would be $92,000. Thus, when incorporating recovery credits, it is imperative to select the value of the recovered product carefully.

Table 1.8: Estimated Annual Costs for the Carbon Adsorber

Cost Item	Factors	Unit Cost	Calculation	Estimated Cost
Direct Annual Costs (DAC)				
Operator Labor Costs:				
Operator	0.5 hours/shift	$27.48/hour[d]	0.5h/shift x 3 shifts/day x 360 days/yr x $27.48/hr	$14,839
Supervisor	15% of operator cost		0.15 x 14,839	$2,226
Operating Materials	—			
Maintenance Costs:				
Labor	0.5h/shift	$27.48/hour[d]	0.5h/shift x 3 shifts/day x 360 days/yr x $27.48/hr x 1.1	$16,323
Material	100% of maintenance labor			$16,323
Carbon Replacement (assuming a 5-year life):				
Labor[b]	CRF	$30.23/hour[d]	0.2310 x $30.23/hour x 10,800 lb/379lbs/hour	$199
Carbon[a,b]	CRF x 1.08	$4.20/lb	0.2310 x $4.20/lb x 10,800 x 1.08	$11,311
Utilities				
Electricity	—	$0.0676/kWh	$0.0676 /kWh x 130,835 kWh/yr	$8,838
Steam	—	$5.00/1,000 lbs	3.5 lb/lb VOC x $5/1000 lb x 100 lb VOC/hr x 8,640 hr/yr	$15,120
Cooling Water	—	$3.55/1,000 gallons[e]	3.43 gal/lb steam x [(3.5 x 100 x 8,640) lb steam x $3.55/1000 gal] / yr	$36,822
Total Direct Annual Cost (DAC)				$122,004
Indirect Annual Costs (IAC)				
Overhead	60% of sum of operating labor, maintenance labor, & maintenance materials		0.6 (14,839 + 2,226 + 16,323 + 16,323)	$29,828
Administrative charges	2% of TCI		0.02 ($445,000)	$10,560
Property tax	1% of TCI		0.01 ($445,000)	$5,280
Insurance	1% of TCI		0.01 ($445,000)	$5,280
Capital recovery[a,b]	CRF[TCI – (1.08Cc + Labor Cost for Carbon Replacement)]		0.0963 [445,000 – (1.08 x 45,360) + 861]	$46,069
Total Indirect Annual Cost (IAC)				$97,017
Recovery Credit for toluene				($279,418)
Total Annual Cost (TAC)[c]				($60,400)

[a] The 1.08 factor applied to the carbon replacement cost is for freight and sales taxes.

[b] The capital recovery cost factor, CRF, is a function of the carbon or equipment life and the opportunity cost of the capital (i.e., interest rate). The CRF is calculated using the following equation: $i(1+i)^n/((1+i)^n-1)$, where n is equipment life and i is the interest rate. For a 5-year carbon life and 5% interest rate, the CFR is 0.2310. For a 15 year equipment life (absorber vessel and auxiliary equipment) and a 5% interest rate, the CRF = 0.0963.

[c] Value shown has been rounded to three significant figures.

[d] Bureau of Labor Statistics, *May 2017 National Occupational Employment and Wage Estimates – United States*, May 2017 (http://www.bls.gov/oes/current/oes_nat.htm).

[e] Based on industrial water rates for users with greater than 15,000 gal monthly usage who purchase water from a municipality. Industrial users that have their own water source or supply with likely have lower water rates [27].

Annual Cost Effectiveness: The total amount of VOC removed can be calculated as follows:

$$VOC\ \mathrm{Re}\,moved\ /\ year\ = \frac{100\,\frac{lb}{hour} \times 0.98 \times \frac{8640\,hours}{year}}{2{,}000\,\frac{lb}{ton}} = 423\,\frac{tons}{year}$$

And the annual cost in terms of VOC removed, or cost effectiveness, is calculated using equation 1.41 and results in a credit of $170/ton of toluene recovered:

$$Annual\ Cost\ Effectiveness = \frac{TAC}{VOC\ Removed} = \frac{(\$60{,}400)}{423\ tons} = (\$143)/ton$$

1.10 Acknowledgements

The authors gratefully acknowledge the following companies for contributing data to this chapter:

- Calgon Carbon Corporation (Moon Township, PA)
- Carbtrol Corporation (Bridgeport, CT)

1.11 References

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