Chapter 4

CARBON ADSORBERS

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4.1.1 Introduction

In air pollution control, adsorption is employed to remove volatile organic compounds (VOCs) from low to medium concentration gas streams, when a stringent outlet concentration must be met and/or recovery of the VOC is desired. Adsorption itself is a phenomenon where gas molecules passing through a bed of solid particles are selectively held there by attractive forces which are weaker and less specific than those of chemical bonds. During adsorption, a gas molecule migrates from the gas stream to the surface of the solid where it is held by physical attraction releasing energy—the "heat of adsorption", which approximately equals 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.

Some gases form actual chemical bonds with the adsorbent surface groups. This phenomenon is termed "chemisorption".

Most gases ("adsorbates") can be removed ("desorbed") from the adsorbent by heating to a sufficiently high temperature, usually via steam or (increasingly) hot combustion gases, or by reducing the pressure to a sufficiently low value (vacuum desorption). 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. For example, approximately 3 to 5 percent of organics adsorbed on virgin activated carbon is either chemisorbed or very strongly physically adsorbed and, for all intents, cannot be desorbed during regeneration.[1]

Adsorbents in large scale use include activated carbon, silica gel, activated alumina, synthetic zeolites, fuller's earth, and other clays. This chapter is oriented toward the use of activated carbon, a commonly used adsorbent for VOCs.

4.1.2 Types of Adsorbers

Five types of adsorption equipment are used in collecting gases: (1) fixed regenerable beds; (2) disposable/rechargeable canisters; (3) traveling bed adsorbers; (4) fluid bed adsorbers; and (5) chromatographic baghouses.[2] Of these, the most commonly used in air pollution control are the fixed-bed and cannister types. This chapter addresses only fixed-bed and cannister units.

4.1.2.1 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). (For most VOCs, the LEL ranges from 2500 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 can be used both for bed drying and cooling.) At the end of the desorption cycle (which usually lasts 1 to 1½ hours), the unit sits idle until the source starts up again.

In *continuous* operation a regenerated carbon bed is always available for adsorption, so that the controlled source can operate continuously without shut down. For example, two carbon beds can be provided: while one is adsorbing, the second is desorbing/idled. As each bed must be large enough to handle the entire gas flow while adsorbing, 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 beds operating in the system. This can reduce the amount of extra carbon capacity needed or provide some additional benefits, relative to maintaining a low VOC content in the effluent. (See Section 4.2 for a more thorough discussion of this.)

A typical two-bed, continuously operated adsorber system is shown in Figure 4.1. One of the two beds is adsorbing at all times, while the other is desorbing/idled. As shown here, the VOC-laden gas enters vessel #1 through valve A, passes through the carbon bed (shown by the shading) and exits through valve B, from whence it passes to the stack. Meanwhile, vessel #2 is in the *desorption* cycle. Steam enters through valve C, flows through the bed and exits through D. The steam-VOC vapor mixture passes to a condenser, where cooling water condenses the entire mixture. If part of the VOC is immiscible in water, the condensate next passes to a decanter, where the VOC and water layers are separated. The VOC layer is conveyed to storage. If impure, it may receive additional purification by distillation. Depending on its quality (*i.e.*, quantity of dissolved organics), the water layer is usually discharged to a wastewater treatment facility.

Once steaming is completed, valves C and D are closed and valve E is opened, to allow air to enter to dry and cool the bed. After this is done, the bed is placed on standby until vessel #1 reaches the end of its adsorption cycle. At this time, the VOC-laden gas is valved to vessel #2, while vessel #1 begins its desorption cycle, and the above process is repeated.

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[1] Although steam is the most commonly used regenerant, there are situations where it should *not* be used. An example would be a degreasing operation that emits halogenated VOCs. Steaming might cause the VOCs to decompose.

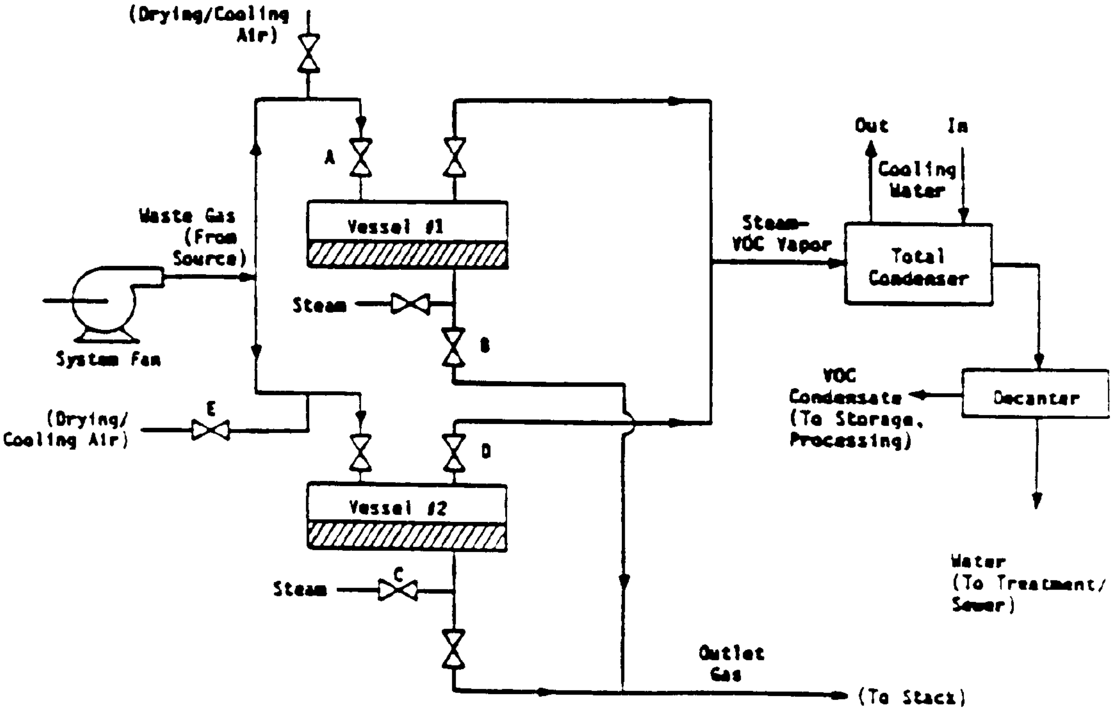

Figure 4.1. Typical-Two-Bed, Continuously Operated Fixed-Bed Carbon Adsorber System

In Figure 4.1, the system fan is shown installed ahead of the vessels, though it could also be placed after them. Further, this figure does not show the pumps needed to bring cooling water to the condenser. Nor does it depict the solvent pump which conveys the VOC condensate to storage. Also missing are preconditioning equipment used to cool, dehumidify, or remove particulate from the inlet gases. Such equipment may or may not be needed, depending on the condition of the inlet gas. In any case, preconditioning equipment will *not* be covered in this chapter.

4.1.2.2 Cannister Units

Cannister-type adsorbers differ from fixed-bed units, in that they are normally limited to controlling low-volume, (typically 100 ft^3/min, maximum) intermittent gas streams, such as those emitted by storage tank vents, where process economics dictate that either toll regeneration or throw-away canisters are appropriate. The carbon canisters are not intended for desorption on-site. However, the carbon may be regenerated at a central facility. Once the carbon reaches a certain VOC content, the unit is shut down, replaced with another, and disposed of or regenerated by the central facility. Each cannister unit consists of a vessel, activated carbon, inlet connection and distributer leading to the carbon bed, and an outlet connection for the purified gas stream.[4] In one design (Calgon's Ventsorb®), 150 lbs of carbon are installed on an 8-inch gravel bed, in a 55-gallon drum. The type of carbon used depends on the nature of the VOC to be treated.

In theory, a cannister unit would remain in service no longer than a regenerable unit would stay in its adsorption cycle. Doing so would help to insure the allowable outlet concentration from being exceeded. In reality, however, poor operating practice may result in the cannister remaining connected until the carbon is near or at saturation. This is because: (1) the carbon (and often the vessel) will probably be disposed of, so there is the temptation to operate it until the carbon is saturated; and (2) unlike fixed-bed units, whose outlet VOC concentrations are usually monitored continuously (via flame ionization detectors, typically), canisters are usually *not* monitored. Thus, the user can only guess at the outlet loading, and could tend to leave a unit in place longer.

4.1.3 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 which relates the mass of adsorbate per unit weight of adsorbent ("equilibrium adsorptivity") to the partial pressure of the VOC in the gas stream. The adsorptivity increases with increasing VOC partial pressure and decreases with increasing temperature.

A family of adsorption isotherms having the shape typical of adsorption on activated carbon is plotted in Figure 4.2. This and other isotherms whose shapes are convex upward throughout, are designated "Type I" isotherms. The Freundlich isotherm, which can be fit to a portion of a Type I curve, is commonly used in industrial design.[2]

$$w_e = kP_m \tag{4.1}$$

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. Except where the VOC in these mixtures have nearly identical adsorption isotherms, one VOC in a mixture will tend to displace another on the carbon surface. Generally, VOCs with lower vapor pressures will displace those with higher vapor pressure, resulting in the former displacing the latter previously adsorbed. 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, one would normally base the adsorption cycle requirements on the *least* adsorbable component in a mixture and the desorption cycle 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, 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 cannister units) replacement and disposal. At the point where the vessel 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 4.1 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 4.1. At very low concentrations, typical of breakthrough concentrations, 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.[5] The vendor presents adsorptivity data in two forms: a set of graphs displaying equilibrium isotherms [5] 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.[6] In this *Manual*, the modified D-R equation is referred to as the Calgon fifth-order polynomial. The data displayed in the Calgon graphs [5] has beenfit

Table 4.1: Parameters for Selected Adsorption Isotherms[*a]

Adsorbate	Adsorption Temp (°F)	Isotherm Parameters *k*	Isotherm Parameters *m*	Range of isotherm[b] (psia)
(1) Benzene	77	0.597	0.176	0.0001-0.05
(2) Chlorobenzene	77	1.05	0.188	0.0001-0.01
(3) Cyclohexane	100	0.508	0.210	0.0001-0.05
(4) Dichloroethane	77	0.976	0.281	0.0001-0.04
(5) Phenol	104	0.855	0.153	0.0001-0.03
(6) Trichloroethane	77	1.06	0.161	0.0001-0.04
(7) Vinyl Chloride	100	0.200	0.477	0.0001-0.05
(8) m-Xylene	77	0.708	0.113	0.0001-0.001
	77	0.527	0.0703	0.001-0.05
(9) Acrylonitrile	100	0.935	0.424	0.0001-0.015
(10) Acetone	100	0.412	0.389	0.0001-0.05
(11) Toluene	77	0.551	0.110	0.0001-0.05

[*] Reference [5].

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

[b] Equations should *not* be extrapolated outside these ranges.

to the Freundlich equation. The resulting Freundlich parameters are shown in Table 4.1 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.

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 4.1 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. Now, as discussed in Section 4.1.2, the adsorbate concentration is usually kept at 25% of the lower explosive limit (LEL)[2]. 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 4.1.

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

Finally, each set of parameters applies to a fixed adsorption temperature, ranging from 77^0 to 104^0 F. These temperatures reflect typical operting conditions, although adsorption can take place as low as 32^0F and even higher than 104^0F. 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.

The Calgon fifth-order polynomial is somewhat more accurate than the Freundilich parameters from Table 4.1. The polynomial contains a temperature parameter, and it allows one to estimate adsorption isotherms for compounds not shown in Table 4.1 if pure component data are available. The pure component data required are the saturation pressure, liquid molar volume, and the refractive index. It is, however, somewhat more complex to use than the Freundlich equation. The Calgon fifth-oder polynomial is as follows:

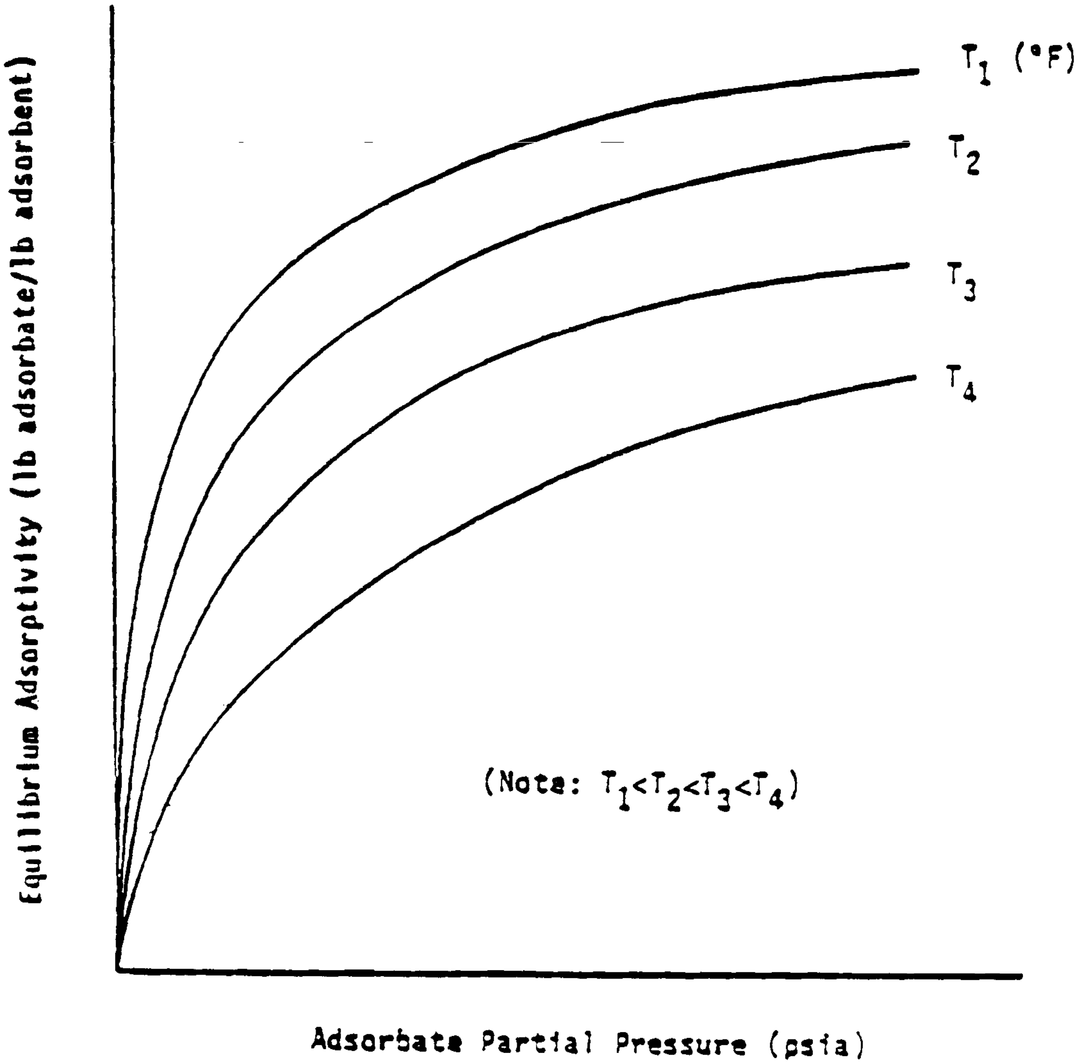

Figure 4.2. Type I Adsorption Isotherms for Hypothetical Adsorbate

The mass loading, w_e, is calculated from

$$w_e = \frac{0.01\,G}{V_m} \times \text{(Molecular Wt of Adsorbate)} \tag{4.2}$$

where

we = mass loading, *i.e.*, equilibrium adsorptivity (g adsorbate per g carbon)*
G = carbon loading at equilibrium (cm^3 liquid adsorbate per 100 g carbon)
V_m = liquid molar volume of adsorbate (cm^3 per g-mole).

Note that the terms in equation 4.2 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:

$$log_{10}(G) = A_0 + A_1Y + A_2Y^2 + A_3Y^3 + A_4Y^4 + A_5Y^5 \tag{4.3}$$

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 χ. This can be done by calculating the adsorption potential, ,:

$$, = RT \ln(P_s/P_i) \tag{4.4}$$

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).

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*This, of course, is equal to *lb* absorbate per *lb* carbon.

The χ is calculated from:

$$\chi = \epsilon/(2.303RV_m)$$

By substituting for ϵ in the above equation, χ can alternatively be calculated from[**]:

$$\chi = (T/V_m)\log_{10}(P_s/P_i).$$

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

$$\Gamma = \Theta_i/\Theta_o$$

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.

For the adsorbate or the reference compound, using the appropriate refractive index of adsorbate, *n*, the polarizability is calculated from:

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

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

$$Y = \frac{P}{'} \tag{4.5}$$

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.[6] Additional sources of isotherm data include the activated carbon vendors, handbooks (such as *Perry's Chemical Engineer's Handbook*), and the literature.

[**] Alternatively, if the available values for T, P_i, P_s, end V_m are in *English* units, they may be substituted into this equation without conversion. However, to make the result dimensionally consistent with equation 4.3, it would have to be multiplied by a conversion factor, *34.7*.

4.2 Design Procedure

4.2.1 Sizing Parameters

Data received from adsorber vendors indicate that the size and purchase cost of a fixed-bed or cannister carbon adsorber system primarily depend on four 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);
4. The *working capacity* of the activated carbon.

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 which would require the use of extensive pretreatment and/or corrosive-resistant construction materials.

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 internal ducting.

Also important are the VOC inlet and outlet gas stream loadings, the adsorption time, and the working 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 [7] supplied data that illustrate the dependency of the equipment cost on the amount of the carbon charge. 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 4.3.)

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.)

4.2.2 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 circle 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 hours worth 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)[7], 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 were operating *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 must depend 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 one bed were adsorbing, a second would be needed to come on-line when the first was shut down for desorption. In this case, 100% extra capacity would be needed. Similarly, if *five* beds in parallel were operating 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 were 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.

$$M_c = M_{c_I} \times f \tag{4.6}$$

where

M_c, M_{c_I} = 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_{c_I}. Section 4.2.3 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:

$$f = 1 + \frac{N_D}{N_A} \tag{4.7}$$

(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:

$$2_D \quad 2_A \left(\frac{N_D}{N_A}\right) \tag{4.8}$$

(Note: 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 have to be 1-1/3 hours or less (8 hours/6 beds). Otherwise, additional beds would have to be added to provide sufficient extra capacity during desorption.

4.2.3 Estimating Carbon Requirement

4.2.3.1 Overview of Carbon Estimation Procedures

Obtaining the carbon requirement (M_c or M_{c_I}) 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 calculational tools available.

One approach for obtaining the carbon requirement is a rigorous one which considers the unsteady-state energy and mass transfer phenomena occurring in the adsorbent bed. Such a procedure necessarily involves a number of assumptions in formulating and solving the problem. Such a procedure is beyond the scope of this *Manual* at the present time, although ongoing work in the Agency is addressing this approach.

In preparing this chapter of the *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.

4.2.3.2 Carbon Estimation Procedure Used in Manual

The rule-of-thumb carbon estimation procedure is based on the "working capacity" (W_e, 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 4.1.3. 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.

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 cannister adsorber or for an *intermittently* operated fixed-bed adsorber as follows:

$$M_{c_I} = \frac{m_{voc}}{w_e} 2_A \qquad (4.9)$$

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

Combining this with Equations 4.6 and 4.7 yields the general equation for estimating the system total carbon charge for a *continuously* operated system:

$$M_c = \frac{m_{voc}}{w_c} 2_A (1 + \frac{N_D}{N_A}) \qquad (4.10)$$

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:

$$w_c \quad 0.5 w_{e(\max)} \qquad (4.11)$$

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, 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 4.10 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 4.10, 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]

4.3 Estimating Total Capital Investment

Entirely different procedures should be used to estimate the purchased costs of fixed-bed and cannister-type adsorbers. Therefore, they will be discussed separately.

4.3.1 Fixed-Bed Systems

As indicated in the previous section, the purchased cost 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 4.1 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 4.3.1.3.) 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. The costs of these major items will be considered separately.

4.3.1.1 Carbon Cost

This cost (C_c,$) is simply the product of the initial carbon requirement (M_c) and the current price of carbon. As adsorber vendors buy carbon in very large quantities (million-pound lots or larger), their cost is somewhat lower than the list price. A typical vendor cost is \$2.00/lb (fall 1989 dollars).[8] Thus:

$$C_c = 2.00 M_c \tag{4.12}$$

4.3.1.2 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[7], while an activated carbon manufacturer cautions against exceeding 60 ft/min in systems operating at one atmosphere.[5] Another vendor uses a 65 ft/min superficial face velocity in sizing its adsorber vessels.[8] 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.[8]

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 commonly, adsorber vessels are cylindrical in shape and erected horizontally (as in Figure 4.1). Vessels configured in this manner are generally subjected to the constraint that the carbon volume occupies no more than 1/3 of the vessel volume [7,8]. It can be shown that this constraint limits the bed depth to no more than

$$\text{Maximum bed depth} \quad \frac{\mathrm{B}D}{12} \tag{4.13}$$

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:

$$D = \frac{0.127 M_c\, v_b}{Q} \tag{4.14}$$

$$\text{and: } L = \frac{7.87}{M_c}\left(\frac{Q}{v_b}\right)^2 \tag{4.15}$$

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 4.14 and 4.15 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:

$$M_C = \frac{M_e}{(N_A + N_D)}$$

$$Q = \frac{Q}{N_A}$$

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.[8] If so, the vessel diameter can be calculated from the volumetric flow rate per adsorbing vessel and the bed superficial velocity as follows:

$$D = \left(\frac{4Q}{\mathrm{B}v_b}\right)^{1/2} \tag{4.16}$$

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:

$$t_b = \frac{\text{volume of carbon}}{\text{cross-sectional area normal to flow}} = \frac{M_c'/\mathrm{D}_b}{Q'/v_b} \tag{4.17}$$

where

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

The vessel length is, therefore,

$$L = t_b + t_{a,g} \tag{4.18}$$

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:

$$S = \pi D(L + D/2) \tag{4.19}$$

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:[8]

$$C_v = 271 S^{0.778} \tag{4.20}$$

where

C_v = vessel cost (fall 1989 \$), F.O.B. vendor*
and $97 \leq S \leq 2,110$ ft^2.

These units would be made of 304 stainless steel, which is the most common material used in fabricating adsorber vessels.[7,8] However, to obtain the cost of a vessel fabricated of another material, multiply C_v by an adjustment factor (F_m). A few of these factors are listed below:

Material	**F_m Factor**	**Reference(s)**
Stainless steel, 316	1.3	[7,8,9]
Carpenter 20 CB-3	1.9	[9]
Monel-400	2.3	[7,9]
Nickel-200	3.2	[9]
Titanium	4.5	[9]

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*For information on escalating these prices to more current dollars, refer to the EPA report *Escalation Indexes for Air Pollution Control Costs* and updates thereto, all of which are installed on the OAQPS Technology Transfer Network (CTC Bulletin Board).

4.3.1.3 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 [8], 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:

$$R_c = 5.82Q^{-0.133} \tag{4.21}$$

where

4, 000 Q (acfm) 500,000
Correlation coefficient (r) = 0.872

The total adsorber equipment cost (C_A) would be the product of R_e and the sum of the carbon and vessel costs, or:

$$C_A = R_c[C_c + (N_A + N_D)C_v] \tag{4.22}$$

4.3.1.4 Total Capital Investment

As discussed in Chapter 2, 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 4.2. As Chapter 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.

Table 4.3: Capital Cost Factors for Carbon Adsorbers[a]

Cost Item	Factor
Direct Costs	
Purchased equipment costs	
Adsorber + auxiliary equipment[b]	As estimated, A
Instrumentation[c]	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
Contractor fees	0.10 B
Start-up	0.02 B
Performance test	0.01 B
Contingencies	0.03 B
Total Indirect Costs, IC	0.31 B
Total Capital Investment = DC + IC	1.61 B + SP + Bldg.

[a] Reference [10].

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

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

Note that the installation factor is applied to the total purchased equipment cost, which includes the cost of such auxiliary equipment as the stack and external ductwork and such costs as freight and sales taxes (if applicable). ("External ductwork" is that ducting needed to convey the exhaust gas from the 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.

4.3.2 Cannister Systems

Once the carbon requirement is estimated using the above procedure, the number of cannisters is determined. This is done simply by dividing the total carbon requirement (M_e) by the amount of carbon contained by each cannister (typically, 150 lbs.). This quotient, rounded to the next *highest* digit, yields the required number of cannisters to control the vent in question.

Costs for a typical cannister (Calgon's Ventsorb®) are listed in Table 4.3. These costs include the vessel, carbon, and connections, but do not include taxes, freight, or installation charges.

Note that the cost per unit decreases as the quantity purchased increases. Each cannister contains Calgon's "BPL" carbon (4 x 10 mesh), which is commonly used in industrial adsorption. However, to treat certain VOCs, more expensive specialty carbons (*e.g.*, "FCA 4 x 10") are needed. These carbons can increase the equipment cost by 60% or more.[4] As is indicated in the caption of Table 4.3, these prices are in Spring 1986 dollars. Since then, however, the prices of these cannisters have increased modestly—approximately 10%.[11]

As fewer installation materials and labor are required to install a cannister unit than a fixed-bed system, the composite installation factor is consequently lower. The only costs required are those needed to place the cannisters 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. *Twenty percent* of the sum of the cannister(s) cost, freight charges, and applicable sales taxes would cover this installation cost.

4.4 Estimating Total Annual Cost

As Chapter 2 of this *Manual* explains, the total annual cost is comprised of three components: *direct costs*, *indirect costs*, and *recovery credits*. These will be considered separately.

Table 4.4: Equipment costs (Spring 1986 $) for a Typical Cannister Adsorber[a]

Quantity	Equipment Cost (each)[b]
1-3	$687
4-9	659
10-29	622
30	579

[a] Reference [4].

[b] These costs are F.O.B., Pittsburgh, PA. They do *not* include taxes and freight charges.

4.4.1 Direct Annual Costs

These include the following expenditures: steam, cooling water, electricity, carbon replacement, operating and supervisor labor, and maintenance labor and materials. Of these, only *electricity* and *solid waste disposal* would apply to the cannister-type adsorbers.

4.4.1.1 Steam

As explained in section 4.1, 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.[7,8] Using the midpoint of this range, we can develop the following expression for the annual steam cost:

$$C_s = 3.50 \times 10^{-3} m_{voc} 2_s \, p_s \qquad (4.23)$$

where

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

If steam price data are unavailable, one can estimate its cost at 120% of the fuel cost. For example, if the local price of natural gas were \$5.00/million BTU, the estimated steam price would be \$6.00/million BTU which is approximately \$6.00/thousand lbs. (The 20% factor covers the capital and annual costs of producing the steam.)

4.4.1.2 Cooling Water

Cooling water is consumed by the condenser in which the steam-VOC mixture leaving the desorbed carbon bed is totally 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.[7] Using the average temperature rise (35°F), we can write:

$$C_{cw} = 3.43 \frac{C_s}{p_s} p_{cw} \tag{4.24}$$

where

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

If the cooling water price is unavailable, use \$0.15 to \$0.30/thousand gallons.

4.4.1.3 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:[5]

$$)P_b / t_b = 0.03679 v_b + 1.1.07 \times 10^{-4} v_b^2 \tag{4.25}$$

where

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

As Equation 4.17 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:

$$t_b = \frac{V_b}{A_b} = \frac{0.0333 M_c^{\prime}}{A_b} \qquad (4.26)$$

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

$$hp_{sf} = 2.50 \times 10^{-4} Q)P_s \qquad (4.27)$$

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.[7][*] 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/coating time per desorption cycle and the number of cycles per year. It can be shown that:

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

$$2_{cf} = 0.42_{D} \, (N_{A} 2_{s} / 2_{A})$$

(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 ($\mathrm{hp}_{\mathrm{cwp}}$) would be computed as follows:

$$\mathrm{hp}_{\mathrm{cwp}} = \frac{2.52 \times 10^{-}4 q_{cw} H s}{0} \tag{4.29}$$

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 4.28, 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 4.29 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 cannister units, use equation 4.27 to calculate the fan horsepower requirement. However, instead of ΔP_b use the following to compute the *total* cannister pressure drop (ΔP_e inches of water):[4]

$$\Delta P_c = 0.0471 Q_c + 9.29 \times 10^{-4} Q_c^2 \tag{4.30}$$

where Q_c = flow through the cannister (acfm).

4.4.1.4 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 Chapter 2, we have:

$$CRC_c = CRF_c (1.08 C_c + C_{cl}) \tag{4.31}$$

where

CRF_c = capital recovery factor for the carbon

1.08 = taxes and freight factor

C_c, C_{cl} = initial cost of carbon (F.O.B. vendor) and carbon replacement labor cost, respectively ($)

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. For example, to remove and replace a 50,000 pound carbon charge would require about 16 person-days, which, at typical wage rates, is equivalent to approximately $0.05/lb replaced.[12]

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—perhaps as low as *two* years—would be likely.[1] For a five-year life and 7% interest rate, $CRF_c = 0.2439$.

4.4.1.5 Solid Waste disposal

Disposal costs are rarely incurred with fixed-bed adsorbers, because the carbon is almost always regenerated in place, not discarded. In certain cases, the carbon in cannister units is also regenerated, either off-site or at a central regeneration facility on-site. However, most cannister adsorbers are disposed of once they become saturated. The entire cannister—carbon, drum, connections, etc.—is shipped to a secure landfill. The cost of landfill disposal could vary considerably, depending on the number of cannisters disposed of, the location of the landfill, etc. Based on data obtained from two large landfills, for instance, the disposal cost would range from approximately $35 to $65 per cannister excluding transportation costs.[13,14]

4.4.1.6 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.[10] The annual labor cost would then be the product of this labor requirement and the operating labor wage rate ($/h) which, naturally, would vary according to the facility location, type of industry, etc. Add to this 15% to cover supervisory labor, as Chapter 2 suggests.

4.4.1.7 Maintenance Labor and Materials

Use 0.5 hours/shift for maintenance labor [10] 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 Chapter 2 suggests. Finally, for maintenance materials, add an amount equal to the maintenance labor, also per Chapter 2.

4.4.2 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 employed. (See Chapter 2 for a thorough discussion of the capital recovery cost and the variables that determine it.) For adsorbers, the system lifetime is typically *ten 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 4.31, we obtain:

$$CRC_s = [TCI - (1.08C_c + C_{cl})]\ CRF_s \tag{4.32}$$

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 cost, respectively ($)
CRF_s = capital recovery factor for adsorber system (defined in Chapter 2).

For a ten-year life and a 7% annual interest rate, the CRF_s would be 0.1424.

As Chapter 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 *cannister* units as well, except that, in most cases, the carbon is not replaced—the entire unit is. Cannisters are generally used in specialized applications. The piping and ducting cost can usually be considered a capital investment with a useful life of ten years. However, whether the cannister 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.

4.4.3 Recovery Credits

These apply to the VOC which is adsorbed, then desorbed, condensed, and separated from the steam condensate. If the recovered VOC is sufficiently pure, it can be sold. However, if the VOC layer contains impurities or is a mixture of compounds, it would require further treatment, such as distillation. Purification and separation costs are beyond the scope of this chapter. Needless to say, the costs of these operations would offset the revenues generated by the sale of the VOC. Finally, as an alternative to reselling it, the VOC could be burned as fuel and valued accordingly. In any case, the following equation can be used to calculate these credits:

$$RC = m_{voc} \theta_s P_{voc} E \tag{4.33}$$

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 absorption cycle. To do this would require knowledge of the temporal variation of the outlet loading during the adsorption cycle. If this knowledge is not available to the *Manual* user, a *conservative* approximation of the efficiency may be made by setting the outlet loading equal to the breakthrough concentration.

4.4.4 Total Annual Cost

Finally, as explained in Chapter 2, the total annual cost (TAC) is the sum of the direct and indirect annual costs, less any recovery credits, or:

$$TAC = DC + IC - RC \tag{4.34}$$

4.4.5 Example Problem

A source at a printing plant emitting 100 lb/h of toluene is to be controlled by a carbon adsorber. The plant proposes to operate the adsorber in a continuous mode for 8,640 h/yr (360 days). While operating, two carbon beds will be adsorbing, while a third will be desorbing/on stand by. For its convenience, the plant has selected adsorption and desorption times of 12 and 5 hours, respectively. The total waste gas flow is 10,000 acfm at the adsorber inlet conditions (one atmosphere and 77°F). The waste gas contains negligible quantities of particulate matter and moisture. Further, the applicable VOC regulation requires the adsorber to achieve a mean removal efficiency of 98% during the entire adsorption cycle. Finally, assume that the recovered toluene is recycled at the source. Estimate the total capital investment and total annual cost for the adsorber system.

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 this partial pressure and the toluene isotherm parameters (from Table. 4.1) into equation 4.1, we obtain an equilibrium capacity of 0.333 lb/lb. By applying the rule-of-thumb discussed above (page 4-19), we obtain a working capacity of 0.167 lb/lb (*i.e.*, 0.333/2).

Carbon Requirement: As stated above, this adsorber would have two beds on-line and a third off-line. Is this a reasonable assumption? Equation 4.8 can answer this question. Substitution of the adsorption time and numbers of adsorbing and desorbing beds yields:

$$\text{Desorption time} = 2_D \quad 2_A(N_D/N_A) = 12\text{ h }(1/2) = 6\text{ h.}$$

Because the stated desorption time (5 hours) is less than 6 hours, the proposed bed configuration is feasible. Next, calculate the carbon requirement (M_c) from equation 4.10:

$$M_c = \frac{m_{voc}}{w_c} 2_A\left(1 + \frac{N_D}{N_A}\right) = \frac{100\,\text{lb/h}\,(12\text{h})(1 + 1/2)}{0.167\,\text{lb/lb}} = 10{,}800\text{ lbs.}$$

From equation 4.12, the carbon cost is:

$$C_c = 2.00M_c = \$21{,}600.$$

Adsorber Vessel Dimensions and Cost: Assume that the vessels will be erected horizontally and select a superficial bed velocity (v_b) of 75 ft/min. Next, calculate the vessel diameter (D), length (L), and surface area (S) from equations 4.14, 4.15, and 4.19, respectively. [Note: In these equations, $M'_c = M_c (N_A + N_D) = 3{,}600$ lb and $Q' = Q/N_A = 5{,}000$ acfm.]

$$D = \frac{0.127\ M_e\ v_b}{Q^{\prime}} = \frac{0.127(3{,}600)(75)}{5{,}000} = 6.86\ \mathrm{ft}$$

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

$$S = \mathrm{B}D(L + D/2) = 283\ ft^2$$

Because S falls between 97 and 2,110 ft^2, equation 4.20 can be used to calculate the cost per vessel, C_v (assuming 304 stainless steel construction). Thus:

$$C_v = 271S^{0.778} = \$21{,}900$$

Adsorber Equipment Cost: Recall that the adsorber 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 cost of the latter items are "factored" from the combined cost of the vessels and carbon. Combining equations 4.21 and 4.22, we have:

$$C_A = 5.82Q^{-0.133}[C_c + (N_A + N_D)C_v]$$

Substitution of the above values yields:

$$C_A = \$149{,}300$$

Cost of Auxiliary Equipment: Assume that costs for the following auxiliary equipment have been estimated from data in other parts of the *Manual*:

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

Total Capital Investment: The total capital investment is factored from the sum of the adsorber unit and auxiliary equipment cost, as displayed in Table 4.4. Note that no line item cost has been shown for instrumentation, for this cost is typically included in the adsorber price.
Therefore:

Purchased Equipment Cost = "B" = 1.08 x "A"
= 1.08 x (\$149,300 + \$32,200) = \$196,000

And:

Total Capital Investment (rounded) = 1.61 x "B" = \$316,000

Annual Costs: Table 4.5 gives the direct and indirect annual costs for the carbon adsorber system, as calculated from the factors in Section 4.4. Except for electricity, the calculations in the table show how these costs were derived. The following discussion will deal with the electricity cost.

First, recall that the electricity includes the power for the system fan, bed drying/cooling fan, and the cooling water pump. (The solvent pump motor is normally so small that its power consumption may be neglected.) These consumptions are calculated as follows:

System fan: From equation 4.27:

Table 4.4: Capital Costs for Carbon Adsorber System
Example Problem

Cost Item	Cost
Direct Costs	
Purchased equipment costs	
Adsorber vessels and carbon	$149,300
Auxiliary equipment	32,200
Sume = A	$181,500
Instrumentation, 0.1A[a]	-------
Sales taxes, 0.03A	5,450
Freight, 0.05A	9,080
Purchased equipment cost, B	$196,000
Direct installation costs	
Foundation and supports, 0.08B	15,680
Handling & erection, 0.14B	27,440
Electrical, 0.04B	7,840
Piping, 0.02B	3,920
Insulation for ductwork, 0.01B	1,960
Painting, 0.01B	1,960
Direct installation cost	$58,800
Site preparation	------
Facilities and building	------
Total Direct Cost	$254,800
Indirect Costs (installation)	
Engineering, 0.10B	19,600
Construction and field expenses, 0.05B	9,800
Contractor fees, 0.10B	19,600
Start-up, 0.02B	3,920
Performance test, 0.01B	1,960
Contingencies, 0.03B	5,880
Total Indirect Cost	$ 60,760
	=======
Total Capital Investment (rounded)	$316,000

[a] The cost for this is included in the adsorber equipment cost.

$$\text{kWh}_{sf} = 0.746\text{kW/hp} \times 2.50 \times 10^{-4} Q) P_s \times 2_s$$

But:

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

(The latter expression was derived from equation 4.25, 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 4.26 can be used to calculate the bed thickness (t_b):

$$\text{Bed thickness} = t_b = \frac{0.0333\, M_c^{/}}{A_b} = \frac{0.0333\, M_c^{/}}{LD} = 1.80 \text{ ft}$$

Thus:

$$)P_s = 1 + 1.80(0.03679 \times 75 + 1.107 \times 10^{-4} \times 75^2) = 7.09 \text{ inches}$$

and finally:

$$\text{kWh}_{sf} = 0.746 \times 2.50 \times 10^{-4} \times 7.09 \text{ in.} \times 10{,}000 \text{ acfm} \times 8640 \text{ h/yr}$$
$$\text{kWh}_{sf} = 114{,}200 \text{ kWh/yr}$$

Bed drying/cooling fan: During the drying/cooling cycle, the pressure drop through the bed also equals ΔP_b. However, as section 4.4.1.3 indicates, the flow and operating time are different. For the air flow, take the midpoint of the range given on page 4-30 (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 4.27 results in:

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

From equation 4.28, we get:

$$\Theta_{cf} = (0.4)(5 \text{ h})(2)(8{,}640 \text{ h})/12 \text{ h} = 2{,}880 \text{ h}$$

Thus:

$$kWh_{cf} = 0.746 \text{ kW/hp} \times 5.32 \text{ hp} \times 2{,}880 \text{ h} = 11{,}400 \text{ kWh/yr}$$

Cooling water pump: The cooling water pump horsepower is calculated from equation 4.29. Here, let $\eta = 63\%$ and $H = 100$ ft. The cooling water flow (qcw) is the quotient of the annual cooling water requirement and the annual pump operating time. From the data in Table 4.5, we obtain the cooling water requirement: 10,400,000 gal/yr. The pump annual operating time is obtained from equation 4.28 (substituting 0.6 for 0.4), or $\theta cwp = (0.6)(5\text{ h})(2)(8{,}640)/12 = 4{,}320$ h/yr.

Thus:

$$hp_{cwp} = \frac{(2.52 \times 10^{-4})(100ft)}{0.63} \times \frac{10{,}400{,}000 \text{ gal/yr}}{4{,}320 \text{ h/yr} \times 60 \text{ min/yr}} = 1.60 \text{ hp}$$

And:

$$kWh_{cwp} = 0.746 \text{ kW/h} \times 1.60 \text{ hp} \times 4{,}320 \text{ h/yr} = 5{,}160 \text{ kWh/yr}$$

Summing the individual power consumptions, we get the value shown in Table 4.5:131,000kWh/yr **Recovery Credit**: As Table 4.5 indicates, a credit for the recovered toluene has been taken. However, to account for miscellaneous losses and contamination, the toluene is arbitrarily valued at one-half the November 1989 market price of \$0.0533/lb(= \$111/ton).[15]

Total Annual Cost: The sum of the direct and indirect annual costs, less the toluene recovery credit, yields a net total annual cost of \$76,100. Clearly, this "bottom line" is very sensitive to the recovery credit and, in turn, the value given the recovered toluene. For instance, if it had been valued at the full market price (\$221/ton), the credit would have doubled and the total annual cost would have been \$29,200. Thus when incorporating recovery credits, it is imperative to select the value of the recovered product carefully.

Table 4.6: Annual Costs for Carbon Adsorber System Example Problem

Cost Item	Calculations	Cost
Direct Annual Costs, DC		
Operating Labor		
Operator	0.5h/shift x 3 shi/day x 360 days/yr x $12/h	$6,480
Supervisor	15% of operator = .15 x 6,480	970
Operating materials		—
Maintenance		
Labor	0.5h/shift x 3sh/day x 360days/yr x $13.20/hr	7,130
Material	100% of maintenance labor	7,130
Replacement parts, carbon	(5-year life)	
Replacement labor	0.2439 ($0.05/lb x 10,800 lb)	130
Carbon cost[a]	0.2439 ($21,600 x 1.08)	5,690
Utilities		
Electricity	$0.06/kWh x 131,000k Wh/yr	7,860
Steam	3.5lb/lbVOCx \$6/10^3lb x 100lbVOC/h x 8640h/yr	18,140
Cooling water	3.43gal/lbsteam x $\frac{(3.5 \times 100 \times 8640)\text{lb steam}}{\text{yr}}$ x \$0.20/10^3gal	2,070
Total DC		$55,600
Indirect Annual Costs, IC		
Overhead	60% of sum of operating, supv., & maint., labor & maint. materials = 0.6 (6,480 + 970 + 7,130 + 7,130)	13,030
Administrative charges	2% of Total Capital Investment = 0.02($316,000)	6,320
Property tax	1% of Total Capital Investment = 0.01($316,000)	3,160
Insurance	1% of Total Capital Investment = 0.01($316,000)	3,160
Capital recovery[b]	0.1424 [316,000 - 0.05(10,800) - 1.08(21,600)]	41,600
Total IC		$67,270
Recovery Credit (toluene)		(46,820)
Total Annual Cost (rounded)		$76,100

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[a] The 1.08 factor is for freight and sales taxes.

[b] The capital recovery cost factor, CRF, is a function of the adsorber or equipment life and the opportunity cost of the capital (i.e., interest rate). For example, for a 10 year equipment life and a 7% interest rate, CRF = 0.1424.

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[15] *Chemical Marketing Reporter*, December 2, 1989.