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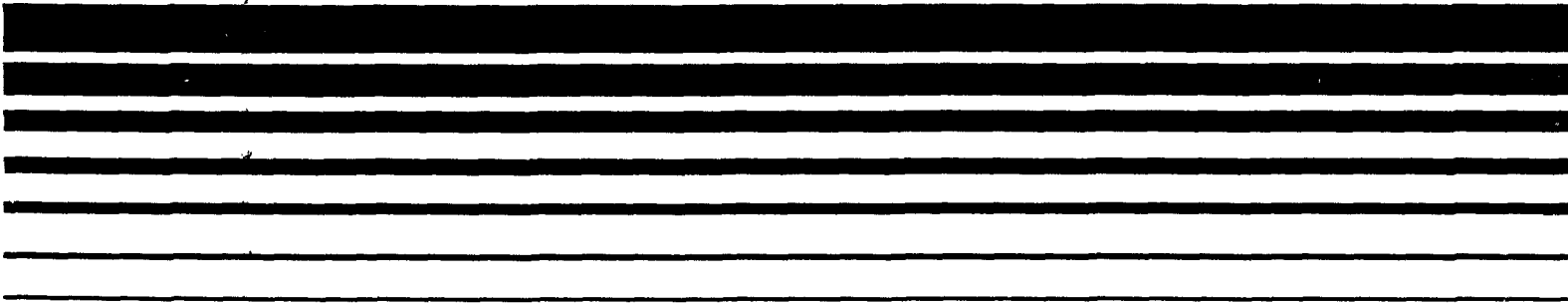
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Alternative Control Technology Document – Carbon Reactivation Processes

ACT

Alternative Control

Technology Document -

Carbon Reactivation

Processes

Emission Standards Division

ALTERNATIVE CONTROL TECHNOLOGY DOCUMENTS

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1.0 INTRODUCTION

Carbon adsorption is one of the most commonly used organic air emission control technologies at hazardous waste treatment, storage, and disposal facilities (TSDF) and industrial manufacturing concerns. Activated carbon is effective in capturing nearly all types of organic vapors by the physical adsorption mechanism. As physical adsorption is a surface phenomenon, activated carbon has a finite or "equilibrium" adsorption capacity (i.e., a finite surface area available for adsorption). When the equilibrium adsorption capacity is reached, the carbon is "saturated" and no further adsorption can occur. At this point, referred to as "breakthrough," the removal efficiency of the carbon adsorption control device approaches zero and the saturated carbon either must be replaced with new carbon or the organic compounds must be removed from the carbon before adsorption can resume. The process of removing organics from saturated carbon can be accomplished by regeneration or reactivation.

In this document, the term "carbon regeneration" refers to the onsite, in situ desorption of organics from the carbon that typically takes place routinely as part of the operation of the carbon adsorption system. Desorption can be achieved by application of either steam or vacuum. In carbon regeneration, the carbon is <u>not</u> removed from the adsorption bed or column. Over time, the carbon loses a significant portion of its reactivity and conventional regeneration is no longer adequate to restore the adsorptive capacity of the carbon. At this point, the carbon must be replaced. The spent carbon can either be disposed of or reactivated.

In carbon reactivation, the carbon is removed from the adsorption bed or column and reactivated in a separate processing unit or furnace. The carbon reactivation process involves exposure of the spent carbon to high temperatures (i.e., thermal reactivation) in an activating atmosphere with sufficient oxygen and steam to remove the organic contaminants from the carbon. Combustion conditions within the furnace are controlled to effect oxidation of the adsorbed material rather than the carbon.

The U.S. Environmental Protection Agency (EPA) issued a statement in the preamble to the boilers and industrial furnaces (BIF) rules (56 FR 7200, Feb. 21, 1991) clarifying the regulatory status of carbon regeneration/reactivation units under the Resource Conservation and Recovery Act (RCRA). Under the definition of a carbon regeneration/reactivation promulgated in the BIF rules (40 CFR 260.10), these units are now regulated as thermal treatment units under the interim status standards of Part 265, Subpart P, and the permit standards of Part 264, Subpart X. Previously, there was some confusion as to whether regeneration/reactivation units had been classified as incinerators, thermal treatment units, or exempt recycling units.

As EPA interprets its rules, the regeneration or reactivation of spent carbon from a carbon adsorption system used in the treatment of a listed hazardous waste or used to capture emissions from a listed hazardous waste is subject to regulation as a RCRA thermal treatment unit because the carbon, with the adsorbed organics, is classified as a hazardous waste under the "derived-from" rule. The derived-from rule (40 CFR 261.3(c)(2)(i)), in short, states that any solid waste generated from the treatment of a listed hazardous waste is a hazardous waste. For other applications of carbon adsorption systems such as the control of air emissions from a production operation or for the treatment of a nonhazardous industrial waste, the regeneration or reactivation of spent carbon is subject to regulation as a RCRA thermal

treatment unit only when the spent carbon is listed as a hazardous waste or exhibits one or more of the hazardous waste characteristics.

To help EPA, State, and local regulators identify the types of air pollutants emitted during carbon reactivation and the kinds of air emission control technologies available for carbon reactivation systems, the EPA has prepared this Alternative Control Technologies (ACT) document. Carbon reactivation systems are a potential source of organic air emissions. For example, as the spent carbon reactivation furnace is operated to minimize the oxidation of the carbon, some of the desorbed organics may be released into the atmosphere prior to complete oxidation. Organic emissions from carbon reactivation systems potentially include photochemically reactive and nonphotochemically reactive organics, some of which are toxic or carcinogenic.

The purpose of this ACT document is to provide information about air emissions (i.e., organics, particulates, toxic metals, sulfur dioxide, and nitrogen oxides) from carbon reactivation processes, some of which are subject to RCRA regulations. It is important to note that the emission control technologies for nonhazardous waste carbon reactivation units are the same as those for hazardous waste carbon reactivation units that are regulated as RCRA thermal treatment units. Information regarding in situ carbon regeneration systems, their emissions, or applicable emission control techniques is not within the scope of this ACT document.

This ACT document presents technical information that Federal, State, and local agencies can use to develop strategies for reducing volatile organic compound (VOC) emissions from carbon reactivation processes. The information in this document will enable writers of permits and State implementation plans to identify carbon reactivation emission sources, identify available control alternatives, and evaluate the VOC reduction and cost of implementing controls.

Chapter 2.0 describes the carbon reactivation industry, the process equipment commonly used for reactivation, and types of emissions associated with carbon reactivation. Chapter 3.0 describes alternative control technologies for the reduction of VOC emissions from carbon reactivation. Chapter 4.0 presents air and cross-media environmental impacts and capital and annual cost analyses of the alternative control technologies.

2.0 INDUSTRY DESCRIPTION, PROCESSES, AND EMISSIONS

This chapter provides an overview of the carbon
reactivation industry and the processes used to reactivate
spent carbon. The sources of organic air emissions and the
types of air pollutants emitted from carbon reactivation
processes are identified. Air emission estimates are provided
when adequate data are available.

2.1 DESCRIPTION OF CARBON REACTIVATION INDUSTRY

For the purposes of this industry overview, it is
beneficial to define a few terms. The carbon reactivation
industry can be divided into two segments: commercial
reactivation and industrial reactivation. The commercial
reactivation industry is comprised of facilities that accept
spent carbon from off site, reactivate it, and then return it
to the originator or sell the reactivated carbon. The
commercial reactivation industry typically includes virgin
carbon manufacturers.

The industrial reactivation industry consists of
manufacturers that employ carbon adsorption on site and
generate enough spent carbon to justify operating their own
onsite carbon reactivation furnaces. Due to the amount and
rate of carbon usage in typical air pollution control
applications, spent carbon from carbon adsorption systems used
for air pollution control is usually reactivated at commercial
facilities.

2.1.1 Commercial Carbon Reactivation Industry

Nationwide, 10 companies, representing a total of 14 facilities, were identified that operate commercial carbon reactivation processes.[1-12] A total of 22 furnaces are operated at these 14 facilities. In addition, four of these companies also produce virgin carbon as a commercial product. Based on information gathered during telephone contacts with selected facilities, the commercial reactivation industry appears to be dominated by two or three large companies. These are Calgon Carbon Corporation and Envirotrol, Inc., on the east coast, with Calgon being the larger of the two, and Cameron-Yakima on the west coast. No figures were available on production or market share for any of the companies.

Interestingly, half (or five) of the commercial reactivation companies identified had new units coming on line or had plans for construction of new units. A total of nine new reactivation furnaces are planned. Five of these furnaces are being built at a single new facility location and a sixth furnace is being built at a second new facility location. The remaining three furnaces are being built to expand existing capabilities at each of three separate facility locations. Table 2-1 lists the commercial carbon reactivation processes identified by EPA in the initial information-gathering phase of this task. This list is not intended to be an exhaustive list of all commercial reactivation facilities, but it includes all the major commercial carbon reactivation facilities in the United States contacted by EPA as a part of information-gathering activities.

Within the commercial carbon reactivation industry, there are two processing routes. One is generally referred to as pooled carbon processing. In this case, the carbon user merely sends the spent carbon to the reactivation facility, and the facility sends the user an amount of replacement carbon from the pool of reactivated carbon at the facility. The spent carbon received is added to the pool of other users' carbon.

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TABLE 2-1. SUMMARY OF COMMERCIAL CARBON REACTIVATION FACILITIES AND PROCESS EQUIPMENT[1-12]

Company/facility location	No. of furnaces	Furnace type	Air pollution equipment
Calgon Carbon Corporation			
Pittsburgh, PA	1[a]	MHF	A, DS, BH
Ashland, KY	1[a]	MHF	A, DS, BH
Envirotrol Incorporated			
Beaver Falls, PA	3[a] (+ 1)[a,b]	RK	A, V, WS
Darlington, PA	4	RK	A, V, WS
Cameron-Yakima			
Yakima, WA	1[a]	MHF	A
	1	RK	A
Barnebey-Sutcliffe			
Columbus, OH	3 (+ 1)[b]	RK	WS
Atochem North America, Incorporated			
Pryor, OK	1	MHF	A
	1[a,b]	RK	A, BH, WS
Adsorption Systems			
Morgantown, WV	1[a]	FB	C, A, V, WS
West States Carbon			
Los Angeles, CA			
(Wilmington, CA)	2	RK	A
Parker, AZ	1[a,b]	MHF	A, WS
California Carbon Company			
Wilmington, CA	2	RK	A
Trans-Pacific Carbon			
Blue Lake, CA	1	RK	A

TABLE 2-1. (continued)

Company/facility location	No. of furnaces	Furnace type	Air pollution equipment
Northwest Carbon			
Anderson, CA	1	RK	A, WS/DS
Red Bluff, CA	1[b]	RK	Unknown[b]
	2[b]	MHF	Unknown[b]
	2[a,b]	IRF	Unknown[b]

A   = Afterburner.
BH = Baghouse.
C   = Cyclone.
DS = Dry scrubber.
FB = Fluidized-bed furnace.
IRF = Infrared furnace.
MHF = Multiple hearth furnace.
RK = Rotary kiln.
V   = Venturi scrubber.
WS = Wet scrubber (e.g., packed column).
[a]Accept, or submitting a permit to accept, hazardous waste.
[b]Under construction.

The second processing route is segregated processing, or custom carbon reactivation. In segregated processing, each customer's spent carbon is processed separately without contacting any other user's carbon. Reactivation companies guarantee the return of the same carbon to the same application. According to the carbon companies, the segregated approach allows them to optimize furnace conditions to suit each customer's spent carbon. This method yields the highest quality reactivated product, typically meeting 90 to 100 percent of virgin carbon activity. The other reported advantage of custom reactivation is that segregated processing keeps the customer's and the reactivation facility's contaminant liability controllable and separate from all other carbon users. Keeping carbon wastes separate provides a

"closed loop" between the customer's plant and the reactivation facility and permits cradle to grave tracking of contaminants. Custom reactivation costs range from one-half to one-third the cost of virgin carbon depending on the quality and intensity of reactivation required.

Another division within the commercial reactivation industry is based on the classification of the spent carbon as hazardous or nonhazardous. Currently, 6 of the 10 commercial reactivation companies report that they do not handle any carbon that is classified as hazardous waste either under RCRA or State regulations. A total of five facilities, representing four different companies, process spent carbon that is considered hazardous waste under RCRA. However, the majority of the processes currently under construction plan to treat spent carbon that is considered to be hazardous waste. If all of the new units are permitted as planned, a total of 8 facility locations, including 6 of the 10 commercial reactivation companies, will have the capability of processing spent carbon considered to be hazardous waste under RCRA or State regulations.

2.1.2 Industrial Carbon Reactivation Industry

Few industrial applications of carbon adsorption systems generate enough spent carbon at a rate sufficient to justify the capital expense of an onsite carbon reactivation process. One notable exception is the corn wet milling industry, which often uses carbon adsorption to remove protein impurities from corn syrup and then reactivates the spent carbon in-house. The other exception is carbon adsorption systems used for the treatment of high-strength wastewaters. Carbon adsorption is commonly used as a polishing step in some wastewater treatment systems, but this application does not usually generate spent carbon at a sufficient rate to justify the capital expense of an onsite carbon reactivation process. Also note that most carbon adsorption systems used for air pollution control do not generate spent carbon at a rate sufficient to justify the capital expense of an onsite carbon reactivation process;

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therefore, the spent carbon from air pollution control equipment is usually reactivated at a commercial facility.

In its information-gathering efforts, the EPA was able to identify four companies that operate industrial carbon reactivation facilities nationally.[14-19] Of these four companies, three represent the corn wet milling industry. The EPA did not attempt to identify every industrial carbon reactivation facility in the United States. Even so, the four industrial companies identified represent 14 separate facilities that operate a total of 21 reactivation furnaces. Thus, from EPA's preliminary review of the industrial segment of the carbon reactivation industry, it would appear that, based on the number of facilities, the number of furnaces in use, and the apparent overall spent carbon treatment capacity, the industrial reactivation industry is comparable to the commercial reactivation industry. However, in contrast to the commercial reactivation industry, no new industrial carbon reactivation processes were identified as being planned or currently under construction. Table 2-2 summarizes the industrial carbon reactivation facilities contacted by EPA and the process equipment used at these facilities.

2.2 CARBON REACTIVATION PROCESSES

This subsection describes the types and operation of reactivation furnaces currently employed by carbon reactivators and discusses the major types and sources of air emissions resulting from carbon reactivation. The air pollution control equipment used to handle the furnace off-gas is discussed in Section 3.

Thermal reactivation, whether conducted at a commercial or an industrial facility, is quite similar to the process used to manufacture new activated carbon. High temperature and steam are combined to produce an activating atmosphere that removes contaminants from the carbon's existing pore structure.

TABLE 2-2. SUMMARY OF INDUSTRIAL CARBON REACTIVATION FACILITIES AND PROCESS EQUIPMENT[14-19]

Company/facility location	No. of furnaces	Furnace type	Air pollution equipment
Mobay Chemical Company			
Kansas City, MO	1	FB	A, V, WS
New Martinsville, WV	1	MHF	A, V, WS
Baytown, TX	1	MHF	A, V
Cargill, Incorporated			
Eddyville, IA	2	MHF	A, V, IS
Dayton, OH	1	MHF	A, V, IS
	1	MHF	A
Cedar Rapids, IA	1	MHF	A, V, IS
Memphis, TN	1	MHF	A
	1	MHF	A, V, IS
A. E. Staley			
Decatur, IL	1	IRF	A
Lowden, TN	1	IRF	A
Lafayette, IN	1	MHF	WHB
Archer Daniels Midland Company (ADM)			
Montezuma, NY	2	MHF	WS
Cedar Rapids, IA	2	MHF	WS
Decatur, IL	2	MHF	WS
Clinton, IA	2	MHF	WS

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A   = Afterburner.
FB  = Fluidized-bed furnace.
IRF = Infrared furnace.
IS  = Impingement scrubber.
MHF = Multiple hearth furnace.
V   = Venturi scrubber.
WHB = Waste heat boiler.
WS  = Wet scrubber (e.g., packed column).
```

A carbon reactivation system consists of the following equipment groups:[21]

1. Receiving, conveying, and feeding;
2. Reactivation furnace;
3. Quenching, cooling, and conveying of the carbon;
4. Off-gas equipment (i.e., air pollution controls); and
5. Process controls and instrumentation.

Four types of reactivation furnaces were identified as being used in current reactivation processes: multiple hearth furnaces, rotary kilns, fluidized-bed furnaces, and infrared furnaces. The most prevalent furnace type is the multiple hearth furnace, followed closely by rotary kilns. Only two furnaces were identified that employed fluidized-bed technology and two other furnaces were identified that used infrared technology. All four of these furnace types are simply devices to accomplish heat and mass transfer between gases and solids. Heat transfer is, for the most part, by direct convection and radiation from the gas to the solid. Mass transfer is accomplished solely by convection as the gases pass over the solids.

Thermal reactivation of carbon generally occurs at temperatures of 870 to 1,010 °C (1,600 to 1,850 °F) by the reaction of water vapor and/or carbon dioxide with the adsorbate. As a rule of thumb, 1 pound of steam is fed to the furnace per pound of carbon. Air is used to supply a small amount of excess oxygen, typically 4 to 10 percent of the amount required for burning at stoichiometric conditions. The retention time of carbon in the furnace varies with the furnace type and with the operating conditions for a given furnace. Typically, about 15 minutes are required at the reactivation temperature for optimal reactivation of spent carbon. The different types of reactivation furnaces are described in the next subsection.

2.2.1 Multiple Hearth Furnace[22, 23]

A multiple hearth furnace consists of a number of annular-shaped refractory compartments called hearths, mounted vertically, one on top of the other. Rabble arms connected to a rotating shaft at the center axis of the furnace are used to move the carbon through the furnace. The spent carbon is fed near the center of the upper hearth and is pushed toward the periphery by the rabble arms. At the periphery, the carbon falls to the next hearth where it is then directed back toward the center. It falls to the next hearth, where it is moved back toward the periphery, and so on, until the carbon reaches the final hearth. Burners may be mounted at any hearth to optimize the drying, heating, and reactivation of the spent carbon. A diagram of a multiple hearth furnace is provided in Figure 2-1.

Most multiple hearth furnaces operate countercurrently, with the air stream moving upward through the hearths, but concurrent multiple hearth furnaces are also employed. The sizing of the furnace is based on the sum total of the hearth floor surface areas. The reactivation furnace with the highest spent carbon throughput in the United States is a multiple hearth furnace that processes 40 to 50 ton/day of spent carbon. The typical residence time for spent carbon in a multiple hearth furnace ranges from 30 to 90 minutes.

2.2.2 Rotary Kiln[24]

A rotary kiln is usually a cylindrical, refractory-lined shell that is positioned at a slight incline from horizontal and that rotates around that horizontal axis. A diagram of a rotary kiln is provided in Figure 2-2. The spent carbon is fed at the higher end of the kiln and moves, driven by gravity, down the length of the kiln as the kiln rotates. Most reactivation kilns are fired at the discharge end with air flow countercurrent to the flow of carbon. The rotational speed of the kiln can be varied, and some kilns have two
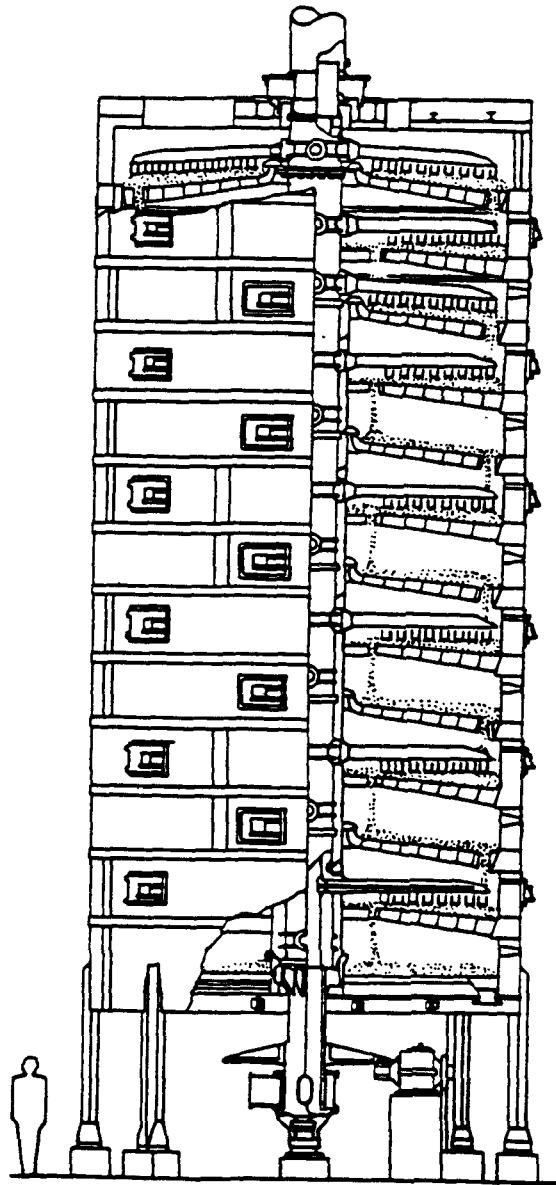
Figure 2-1.   Multiple hearth furnace.
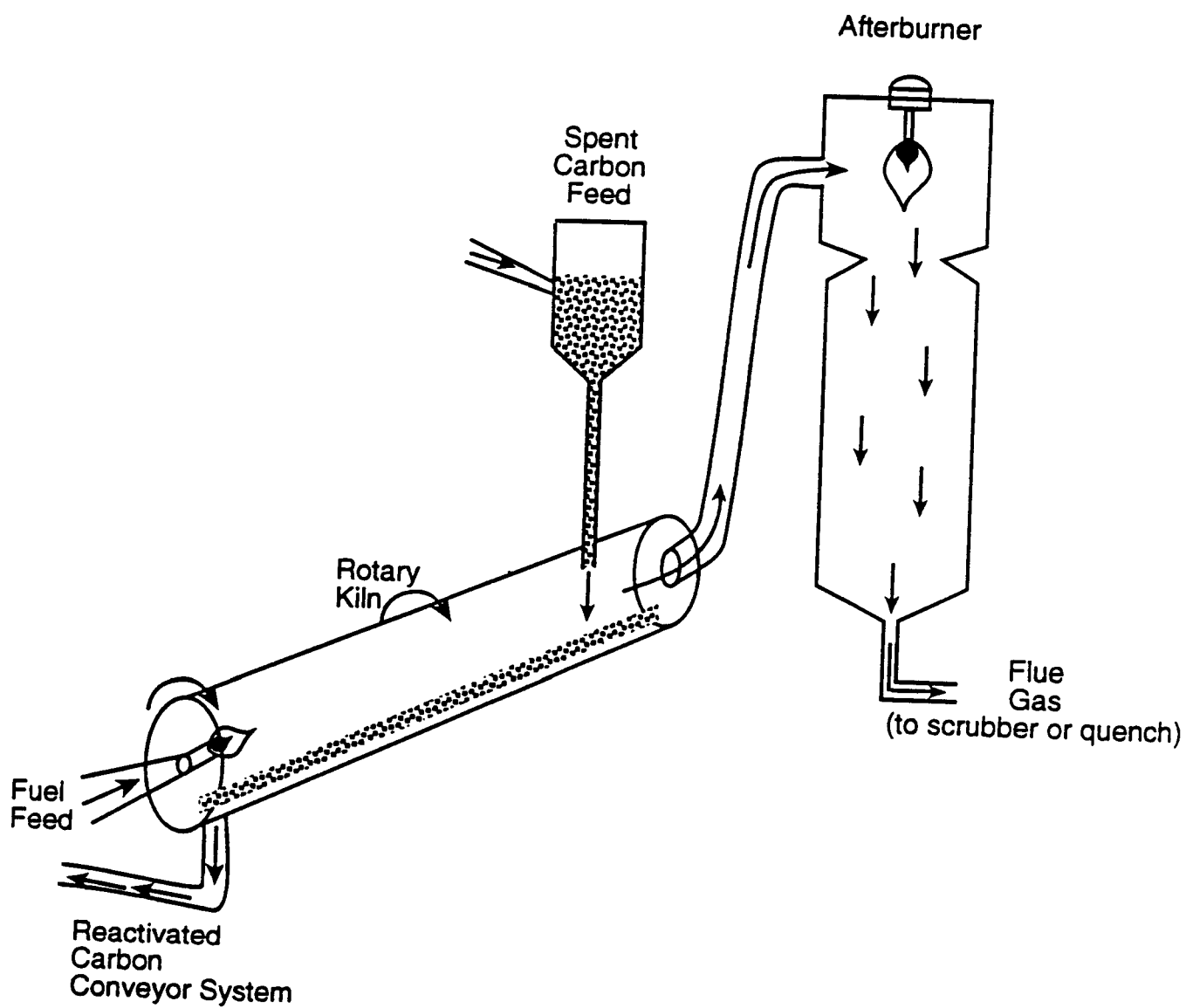
Figure 2-2. Rotary kiln.

different diameters to vary the peripheral speed of the kiln rotation as the carbon moves down the kiln. Peripheral speeds of 0.5 to 2 m/min are typical. The residence time in a rotary kiln is typically less than in a multiple hearth furnace, with average retention times of 20 to 30 minutes.

2.2.3 Fluidized-Bed Furnace[25]

A fluidized-bed furnace is a cylindrical vertical vessel with an air feed at the bottom of the vessel. The air blowing up through the bed lifts or "fluidizes" the carbon, creating a turbulent cloud of carbon particles. Because of the air flow rate required to fluidize the carbon particles, fluidized-bed furnaces have a larger exhaust volume than other types of reactivation furnaces with the same carbon throughput rate. The larger air flow rate tends to increase the amount of carbon fines that are carried over with the exhaust. On the other hand, the high air flow rate creates turbulence in the fluidized bed, and the intimate contact of the air with individual carbon particles provides a very efficient means for heat and mass transfer. A diagram of a fluidized-bed furnace is provided in Figure 2-3.

2.2.4 Infrared Furnace[26]

Infrared furnaces have only recently been used for carbon reactivation. The energy required for the desorption and volatilization of the adsorbed organics is provided by heating a series of heating elements to incandescence using electrical energy. The spent carbon is typically transported through the furnace via a metal grate. The infrared radiation heats the carbon, and an induced draft fan is used to draw air through the furnace and remove the desorbed gases as they are released from the carbon. Residence time in the furnace is controlled by varying the speed of the grate used to transport the carbon through the furnace. Typical residence times for this furnace type when employed for spent carbon reactivation were not available. A diagram of an infrared furnace is provided in Figure 2-4.
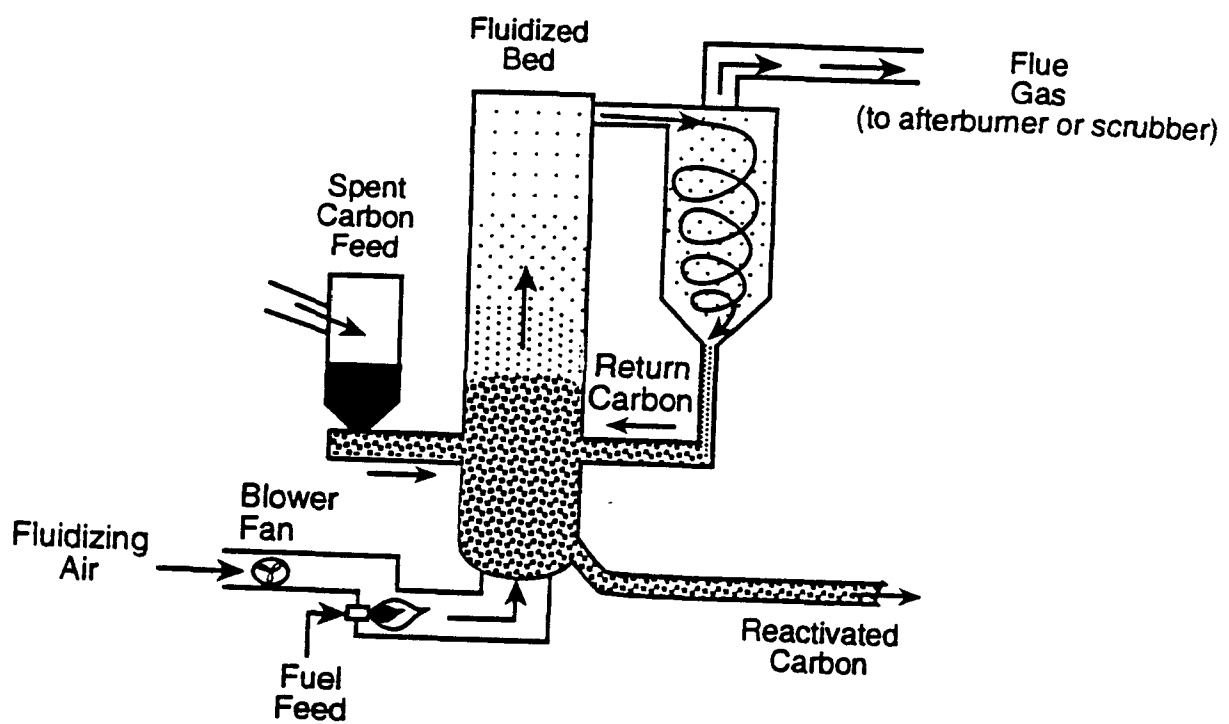
Figure 2-3. Fluidized-bed furnace.
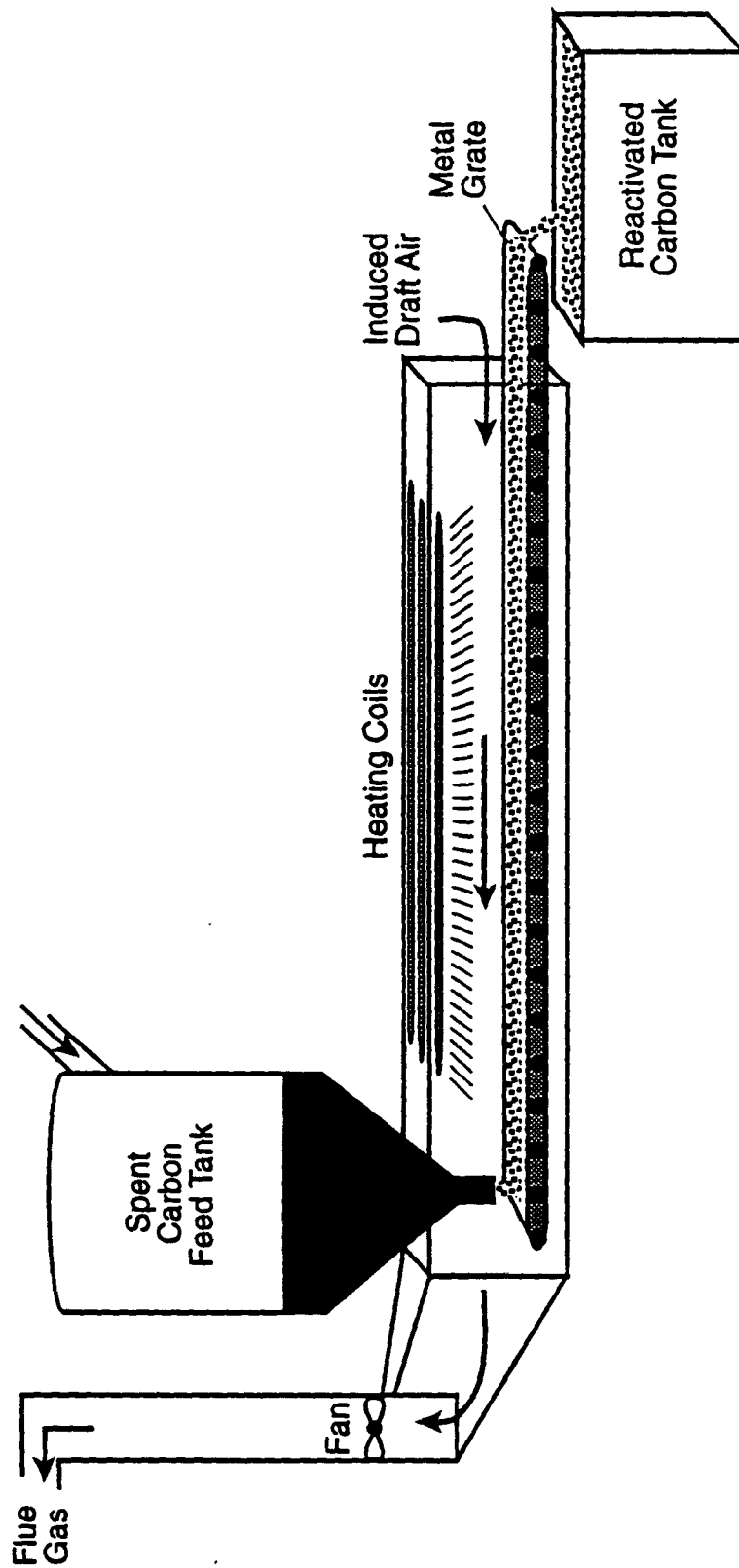
Figure 2-4. Infrared furnace.

2.3 CARBON REACTIVATION AIR EMISSIONS

There are up to four different types of air emissions from carbon reactivation process units: organic emissions, acid gas emissions, particulate emissions, and metal emissions. This section will discuss the sources of these emissions and provide estimates of their emission rates based on carbon throughput rates. Most of the available emission data are from emission measurements made at the final stack exhaust, after emissions passed through all the air pollution control equipment. Therefore, most of the uncontrolled emission rates reported in this section are only order-of-magnitude estimates. An actual emission source test is recommended if more precise emission data are needed for a specific reactivation system.

2.3.1 Organic Air Emissions

The primary source of organic air emissions is the reactivation furnace exhaust gas. The organic emissions from the reactivation furnace consist of organic compounds driven off the spent carbon or formed as products of incomplete combustion. Organic loading on spent carbon can range from a few percent up to 20 percent by weight on a dry basis.

Because of the high temperatures at which carbon reactivation occurs (i.e., 870 to 1,010 °C), a significant portion of desorbed organics pyrolizes or chars on the carbon and is oxidized in the reactivation furnace. No test data for field operating reactivation furnaces were available to determine the actual percentage range of adsorbate oxidized in the reactivation furnace. Estimates provided by the industry varied significantly. The estimated values, based on engineering judgment and operating experience, ranged from a low of about 10 percent to a high of about 90 percent.[4,23,27-29] The variation in the percentage of desorbed organics combusted in the reactivation furnace has a tremendous impact on the estimate of uncontrolled organic emissions. If, for example, the organic loading of the spent carbon averages about

10 percent by weight and the reactivation furnace ranges from 10 to 90 percent efficient in destroying the released or desorbed organics, then the organic air emissions from a carbon reactivation furnace with no afterburner can be estimated as 0.01 to 0.09 times the carbon feed rate on a dry basis. A reactivation unit that processes 11 Mg/day (1,000 lb/hr) of spent carbon may have total organic emissions from the reactivation furnace on the order of 40 to 360 Mg/yr (10 to 90 lb/hr).

Data on total hydrocarbon (as methane) emission rates measured after an afterburner were available for two carbon reactivation systems.[30,31] For these systems, which had average carbon throughput rates of 26 and 48 Mg/day (2,400 and 4,400 lb/hr), the total hydrocarbon emission rates measured for a given run ranged from 0.3 to 7 Mg/yr (0.08 to 1.8 lb/hr). The concentration of total hydrocarbons in the exhaust ranged from 3 to 32 ppm as methane. Based on these data and assuming a 98 percent control efficiency for the afterburner, the uncontrolled hydrocarbon emission rate for a system with a carbon throughput of 11 Mg/day (1,000 lb/hr) is estimated to be 4 to 100 Mg/yr.

2.3.2 Acid Gas Emissions

Acid gases (e.g., hydrogen chloride and hydrogen fluoride) are produced as by-products of the combustion of either the fuel, the adsorbed organics, or the activated carbon itself. As such, the types and amounts of acid gases created in the carbon reactivation furnace (and, if present, the afterburner) are highly variable.

Hydrogen chloride (HCl), hydrogen fluoride (HF), hydrogen bromide (HBr), and hydrogen iodide (HI) are formed by the combustion of halogenated organics contained in the adsorbate. If there are no halogenated organics, there will be no significant halide acid gas emissions. If there are halogenated organics present, a stoichiometric quantity of acid gas in the form of HCl, HF, and/or HBr will be produced.

One facility, while testing the destruction efficiency of its reactivation furnace and afterburner system, measured HCl emissions prior to the acid gas control device.[32] The measurements were made on a system that was processing 9.7 Mg/day (890 lb/hr) of carbon with an organic loading of 1,1,1-trichloroethane at 7 percent (0.68 Mg/day or 62.3 lb/hr of 1,1,1-trichloroethane), which is expected to be an upper limit for halogenated organics adsorbed on carbon. An uncontrolled HCl emission rate of 55 Mg/yr (13.8 lb/hr) was measured. The destruction efficiency of the reactivation furnace and afterburner system based on 1,1,1-trichloroethane mass emissions indicated a destruction efficiency of greater than 99.99 percent. However, if the 1,1,1-trichloroethane had been completely oxidized to carbon dioxide, water, and HCl, the HCl emission rate should have been approximately 200 Mg/yr (51 lb/hr). Thus, estimates based on stoichiometric conversion of halogenated organics to halide acid gases may overestimate the acid gas emissions, but they will provide a reasonable and environmentally conservative estimate of the emission potential. Additionally, these results suggest that there exists a potential problem with products of incomplete combustion (PICs).

Sulfur dioxide (SO_2) is formed from sulfur contained in the reactivation furnace and afterburner fuel and from sulfur contained in the spent carbon. The quantity of sulfur dioxide produced is dependent on the type of fuel used to fire the furnace and afterburner, the sulfur content of the spent carbon, and the amount of carbon combusted during the reactivation process. The sulfur content of carbon typically ranges from 2 to 5 percent. If there is a 5 percent carbon attrition rate (i.e., 5 percent of the carbon is combusted in processing), then for a reactivation unit that processes 11 Mg/day (1,000 lb/hr) of spent carbon, 4 to 10 Mg/yr (1 to 2.5 lb/hr) of sulfur is expected to be released. As the molecular weight of sulfur dioxide is twice that of sulfur,

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approximately 8 to 20 Mg/yr (2 to 5 lb/hr) of sulfur dioxide could be emitted from an uncontrolled reactivation furnace processing 11 Mg/day (1,000 lb/hr) of spent carbon.

Emission source test data from a facility processing 26 Mg/day (2,400 lb/hr) of spent carbon revealed an uncontrolled sulfur dioxide emission rate of 86.8 Mg/yr (21.8 lb/hr).[33] This is slightly higher than expected based on the combustion of the activated carbon, but there are other potential sources of sulfur, such as in the adsorbed organics or in the fuel if oil or coal were used to fire the furnace or afterburner, that can contribute to the total SO_2 emissions. Another facility that does not have an acid gas emission control device has a permit limit of 30 Mg/yr (7.6 lb/hr) of SO_2 for a unit that typically processes 10 Mg/day (900 lb/hr) of spent carbon.[34]

Nitrogen oxides (NO_x) are formed in the reactivation furnace and afterburner combustion processes. The temperatures, residence times, and oxygen contents of typical reactivation operations are just in the range where NO_x formation tends to increase; therefore, it is difficult to make any predictions on NO_x emissions from these carbon reactivation units. The only available NO_x emission data are from measurements taken at one facility subsequent to a dry scrubber used for acid gas removal. The controlled NO_x emission rate for this facility was 9.5 Mg/yr (2.4 lb/hr) while processing approximately 49 Mg/day (4,500 lb/hr) of spent carbon.[35]

2.3.3 Particulate Emissions

There are two sources of particulate emissions from thermal reactivation of spent carbon. From the reactivation furnace itself, the primary particulate emissions consist of carbon fines entrained in the furnace exhaust gas. There are already economic incentives to minimize this source of particulate emissions to minimize the amount of carbon loss. Additionally, the carbon fines are effectively destroyed in

most afterburners. Therefore, particulate emissions from most carbon reactivation furnaces that are equipped with afterburners are typically not significant.

In some reactivation systems, however, a second potential source of particulate emissions exists. Some air pollution control technologies, particularly dry or spray scrubbers used for acid gas removal, may form particulates as a by-product, which subsequently need to be removed from the air stream prior to release to the atmosphere. All available particulate emission measurements for carbon reactivation systems were made downstream of particulate removal devices. From these data, the "controlled" particulate emissions ranged from 0.0013 to 0.057 gr/dscf. Using the exhaust air flow rates for each furnace, the "controlled" particulate emissions ranged from 0.3 to 6 Mg/yr (or 0.08 to 1.5 lb/hr).[36-38]

2.3.4 Metal Emissions

There are two sources of metals in spent carbon: the trace metals that occur naturally in the carbon itself and the metals that are adsorbed on the spent carbon. As reported by the industry, all carbon has some inherent or intrinsic metal content. Most carbons are produced from coal and, in general, have the same basic range/distribution of naturally occurring metals, including mercury (Hg), lead (Pb), barium (Ba), chromium (Cr), and nickel (Ni). The most common process that results in adsorption of metals onto activated carbon is carbon adsorption used in wastewater treatment. Metals other than iron would not likely be encountered when using carbon for treatment of potable water; in addition, it is highly unlikely that metals would be encountered when carbon is used for vapor control.

Metal emissions can occur either as particulate matter or as metal vapors. Metal emissions occurring as particulate matter are primarily metals that remain adsorbed on, or are naturally occurring in, the entrained carbon fines (see Section 2.3.3). Metal particulates can also be formed when

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metal vapors condense prior to the final exhaust. The oxides
of certain metals such as mercury, lead, selenium, and
cadmium have relatively high vapor pressures (volatilities).
When they are formed in the combustion zone, a
disproportionate fraction is present in the postflame as a
vapor. A portion of this metallic vapor is present in the
combustion products; later it condenses as the postflame gases
cool when passing through various ducts and appears as ash,
particulate, or aerosol in the exhaust gas. Before discharge
to the atmosphere, these particulates are typically controlled
by methods described in Section 3.3. The remaining portion of
the metallic vapor typically exits the stack uncontrolled to
the atmosphere.

Carbon reactivators report that they routinely do a
metals analysis of the carbon that enters the facility for
reactivation. Reactivation facilities are most sensitive
about Hg content. If highly volatile metals are present in
"significant" amounts, then the carbon is not accepted for
reactivation. There are exceptions on a case-by-case basis
for treatment of carbons that contain metals; for example, if
the client wants a custom reactivation job where the carbon is
isolated and returned to the client following reactivation.

The fate of metals contained in spent activated carbon
that is thermally reactivated can be traced in three routes.
A portion of the metals will remain in the carbon, including
the collected particulate matter and residual ash; a portion
of the metals present will be controlled or captured in the
wet or dry scrubbing system that is typically present on
nearly all reactivation processes; and a portion of the metals
will be emitted out the stack to the atmosphere.

No data were available to quantify what percentage of
metals are emitted to the atmosphere for carbon reactivation
operations, but there are comparable operations that give some
idea of the variability that occurs from metal to metal. A
study of incineration of waste sludges revealed that: for Hg,

0.4 percent of the Hg remained in the ash, 2 percent was in the scrubber water, and 97.6 percent exited the stack; for Pb, 87 percent remained in the ash, 12 percent was in the scrubber water, and 1 percent was emitted out the stack; for Cd, 80 percent remained in the ash, 20 percent was in the scrubber, and 0 percent was emitted.[39]

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3.0 EMISSION CONTROL TECHNIQUES

This chapter discusses emission control techniques applicable to spent carbon reactivation off-gas streams. These control techniques are grouped into four broad emission categories: organics, acid gases, particulates, and metals. Equipment utilized for the control of these emissions includes afterburners, packed towers, spray towers, venturi scrubbers, impingement scrubbers, cyclones, and baghouses. This equipment is described with respect to its applicability to control carbon reactivation emissions, and equipment design and performance are discussed.

The four emission categories characteristic of carbon reactivation processes present numerous options in terms of the selection and sequencing of air pollution control equipment and techniques. Due to the high temperature of the off-gases from the carbon reactivation furnace, the first control step typically used is an afterburner immediately following the reactivation furnace to destroy the organic materials that were driven from the spent carbon. The next step is typically wet scrubbing by use of either a packed bed or spray tower to: (1) cool the gas, (2) remove the inorganic acid gases and metal vapors from the afterburner exhaust, and (3) remove the particulates. In some cases, especially if a dry scrubber that generates particulates is used, a particulate control device such as a baghouse may also be used as a final control step.

3.1 ORGANIC EMISSION CONTROL

Organic air emissions from carbon reactivation process units consist of organic/hydrocarbon compounds driven from the spent carbon or formed as products of incomplete combustion. Because of the high temperatures of the off-gas exiting reactivation furnaces, this organic matter is usually destroyed by use of a combustion control device. If combustion were complete in these devices, all organics would be converted to carbon dioxide, water, and trace acids. However, since actual combustion is never 100 percent complete, the remaining organics exit from the stack either in vapor form or adsorbed onto particulate matter.

There are two types of combustion control devices: thermal afterburners and catalytic afterburners. Thermal afterburners require less capital investment and therefore are less expensive to install than catalytic afterburners; however, catalytic afterburners operate at lower temperatures and therefore typically require less fuel and are less expensive to operate depending on the life of the catalyst. However, the temperature of the exhaust from the reactivation furnace is already very high (approximately 800 °C [1500 °F] or higher) and carbon fines, which may be entrained with the exhaust, tend to foul most catalysts and thereby shorten the catalyst's useful life. That is, the usual advantages of catalytic afterburners do not apply when considering the control of a reactivation furnace exhaust. Consequently, thermal afterburners are used almost exclusively for the control of organic emissions in the carbon reactivation industry.

A thermal afterburner is shown in Figure 3-1. It usually consists of a refractory-lined chamber that is equipped with one or more sets of burners. The organic-laden stream passes through the burners and is heated above its ignition temperature. The gas is then sent through one or more residence chambers where it is held for a certain length of time to achieve the desired combustion efficiency.
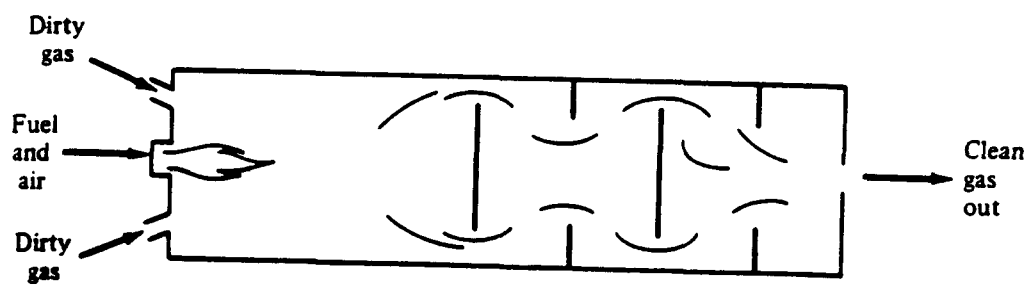
Figure 3-1.   Thermal afterburner.

Residence times of 0.3 to 2 seconds are common when thermal afterburners are used at carbon reactivation facilities.[1]

Thermal afterburners designed specifically for organic incineration with natural gas as the auxiliary fuel may also use a grid-type (distributed) gas burner as shown in Figure 3-2. The tiny gas flame jets on the grid surface ignite the vapors as they pass through the grid. The grid acts as a baffle for mixing the gases entering the chamber. This arrangement ensures burning of all vapors at lower chamber temperature and uses less fuel. This system makes possible a shorter reaction chamber yet maintains high efficiency.

Thermal afterburners operate on the basic principle that any hydrocarbon heated to a high enough temperature in the presence of enough oxygen will be oxidized to CO_2 and water. The theoretical temperature required for thermal oxidation to occur depends on the structure of the chemical involved. Some chemicals are oxidized at temperatures much lower than others. Information collected through site visits to, and telephone contacts with, facilities that have carbon reactivation furnaces with thermal afterburners indicates that typical afterburner temperatures range from 850 to 1,050 °C (1,560 to 1,900 °F).[2-5] The organic destruction efficiency of a thermal afterburner can be affected by variations in chamber temperature, residence time, inlet organic concentration, compound type, and flow regime (mixing). Other parameters affecting afterburner performance are the amount of excess combustion air and the water content and heating value of the off-gas.

The heating value of the off-gas is a measure of the heat available from the combustion of the organics in that stream. Off-gas with heating value below 1.86 MJ/Nm^3 (50 Btu/scf) usually requires auxiliary fuel to maintain combustion temperatures. Off-gas with a heating value above 1.86 MJ/Nm^3 (50 Btu/scf) may support combustion but may need
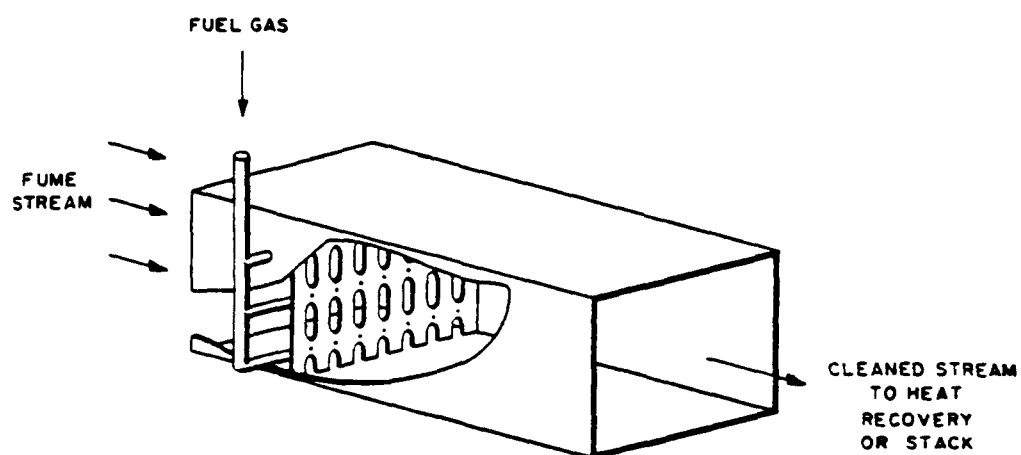
FUEL GAS

FUME STREAM

CLEANED STREAM TO HEAT RECOVERY OR STACK

Figure 3-2. Thermal afterburner with a distributed burner.

3-5

auxiliary fuel for flame stability. Auxiliary fuel requirements can be lessened by the use of recuperative heat exchangers to preheat combustion air. An afterburner handling carbon reactivation off-gases with varying heating values and moisture content requires careful adjustment to maintain the proper chamber temperatures and operating efficiency. Water requires a great deal of heat to vaporize. Entrained water droplets in the off-gas can substantially increase auxiliary fuel requirements due to the additional energy needed to vaporize the water and raise it to the combustion chamber temperature.

Sufficient oxygen to provide for complete theoretical oxidation of both the pollutant and the fuel required must always be ensured when designing the afterburner. The oxygen source for afterburners is air that can enter with the pollutant stream and/or through the afterburner combustion system. The amount of excess air used varies with the fuel and burner type, but it is generally kept to a minimum. Using too much excess air wastes fuel because the additional air must be heated to the combustion chamber temperature. A large amount of excess air also increases flue gas volume and will increase the size and cost of the system. For these reasons, a heat exchanger is sometimes used to preheat the auxiliary air using the afterburner exhaust. In carbon reactivation processes, afterburners used to control organic emissions typically use 6 to 8 percent excess air.[6]

Concerning afterburner capacity, there is no theoretical limit. Units with capacities over 30,000 scfm of process fumes are relatively common. Inlet stream temperatures are typically -7 to 425 °C (20 to 800 °F); however, equipment can be designed to accommodate temperatures outside this range.

The destruction efficiency of thermal afterburners can be affected by various parameters; test results show that they can achieve 98 percent for most organic compounds.[7] The 98 percent efficiency estimate is predicated upon thermal afterburners operated at 870 °C (1,600 °F) with a 0.75-sec

residence time for nonhalogenated organic compounds. If the
vapor stream contains halogenated compounds, a temperature of
1,100 °C (2,000 °F) and a residence time of 1 second is needed
to achieve a 98 percent destruction efficiency. Test results
provided by two commercial carbon reactivators show organic
destruction efficiencies for a reactivation furnace/thermal
afterburner combination greater than 99 percent and as high as
99.9999 percent were achieved.[8, 9]

3.2 ACID GAS EMISSION CONTROL

Acid gas pollutants present in the off-gas from carbon
reactivation process units include hydrogen halides (HCl, HF,
HBr, and HI), NO_x, and SO_2. These pollutants can be
efficiently removed by absorption. An alternate term for
absorption is scrubbing. This section describes the basic
theoretical operating principles of absorption, then describes
the process control equipment used for controlling acid gas
emissions from reactivation furnace off-gas. Control devices
primarily used include packed-bed scrubbers and spray towers.
Venturi scrubbers and impingement scrubbers have limited acid
gas scrubbing capabilities and are also discussed briefly in
this chapter. Removal efficiency and applicability of these
devices to reactivation off-gas streams are also discussed.

3.2.1 Absorption Principles for Acid Gas Emission Control

Absorption is the selective transfer of one or more
components of a gas mixture into a solvent liquid. The
transfer consists of solute diffusion and dissolution into a
solvent. For any given solvent, solute, and set of operating
conditions, there is an equilibrium ratio of solute
concentration in the gas mixture to solute concentration in
the solvent. The driving force for mass transfer at a given
point in an absorption tower is a function of the difference
between the actual concentration ratio and the equilibrium
ratio. The absorbed material may dissolve physically in the
liquid (physical scrubbing) or react chemically with it
(chemical scrubbing).

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In physical scrubbing, the off-gas and scrubbing liquor temperatures, the concentrations of the pollutant in both gaseous and liquid phases, and the solubility of the pollutant in the scrubbing liquor are the major factors that determine effectiveness. The ionic nature of acids, bases, and salts makes these compounds good candidates for removal from off-gases by wet scrubbing. Water by itself is efficient at removing acidic soluble gases such as HCl, HF, and silicon fluoride (SiF_4), especially if the last contact is made with water having a slightly alkaline pH.[10] For pollutant gases with limited water solubility such as SO_2, very large quantities of water are required. For these pollutants, simple wet scrubbing is usually impractical, but may occasionally be employed in unusual circumstances.

Chemical scrubbing, such as use of a caustic solution, often further enhances removal of acid gases and can be used effectively to remove pollutants from reactivation off-gases. Factors that affect chemical reactions, such as reaction rates, temperature, pH of liquor, and concentrations, are also important in determining the effectiveness of this method. Some examples of chemical scrubbing techniques are the absorption of sulfur dioxide using calcium carbonate, calcium oxide, magnesium oxide, sodium hydroxide, and other alkali and the absorption of nitrogen oxides using a urea solution.[11]

The rate of absorption of the acid gas, for either physical or chemical scrubbing, depends on the physical properties of the gas liquid system such as density, diffusivity, equilibrium, solubility, and viscosity. These properties are temperature dependent, and lower temperatures generally favor absorption of gases by the scrubbing liquid.

3.2.2 <u>Absorption Equipment for Acid Gas Emission Control</u>

The most prevalent method of control for inorganic emissions is wet scrubbing. Control devices most commonly used for wet scrubbing are packed-bed and spray tower scrubbers. Tray towers, although potentially applicable, were not used by any of the reactivation facilities identified in

3-8

Section 2. Other wet scrubbing technologies that are available and have been used for emission control at carbon reactivation facilities are venturi scrubbers and impingement scrubbers. These devices, however, are used primarily for particulate removal and are discussed in Section 3.3 on particulate emission controls. An alternative method of control for inorganic emissions is dry scrubbing. This section describes both wet and dry absorption equipment used primarily for acid gas emissions control.

3.2.2.1 Packed Towers.[12] A packed-tower or packed-bed scrubber is shown in Figure 3-3. In these systems, liquid solvent is sprayed through the top of the column on specially designed packing and is allowed to flow through the system. As a result, the large surface area provided by the packing enhances the opportunity for absorption of the pollutant solute. Finely packed beds can remove finer mists, while coarser packing is used to prevent fouling in the presence of coarse particulates. Although gas flow may be countercurrent, crosscurrent or concurrent, most packed towers use a countercurrent gas flow design. The pollutant solute in the gas phase is absorbed by the solvent, usually water, which carries the dissolved solute out the bottom of the tower. Cleaned gas exits at the top for release to the atmosphere or for further treatment as necessary. In some cases, the saturated liquid solvent is directed to a regeneration unit where the solution is treated so that the solvent may be recycled and the pollutant disposed of appropriately or recovered. When water is used as the solvent or absorbent, the solution may be pumped directly to an onsite wastewater treatment system where secondary emissions may occur.

The major packed tower design parameters to be determined for absorbing any substance are column diameter and height, system pressure drop, and liquid flow rate required. These parameters are derived from consideration of the total surface area provided by the tower packing material, the solubility and concentrations of the components, and the quality of the
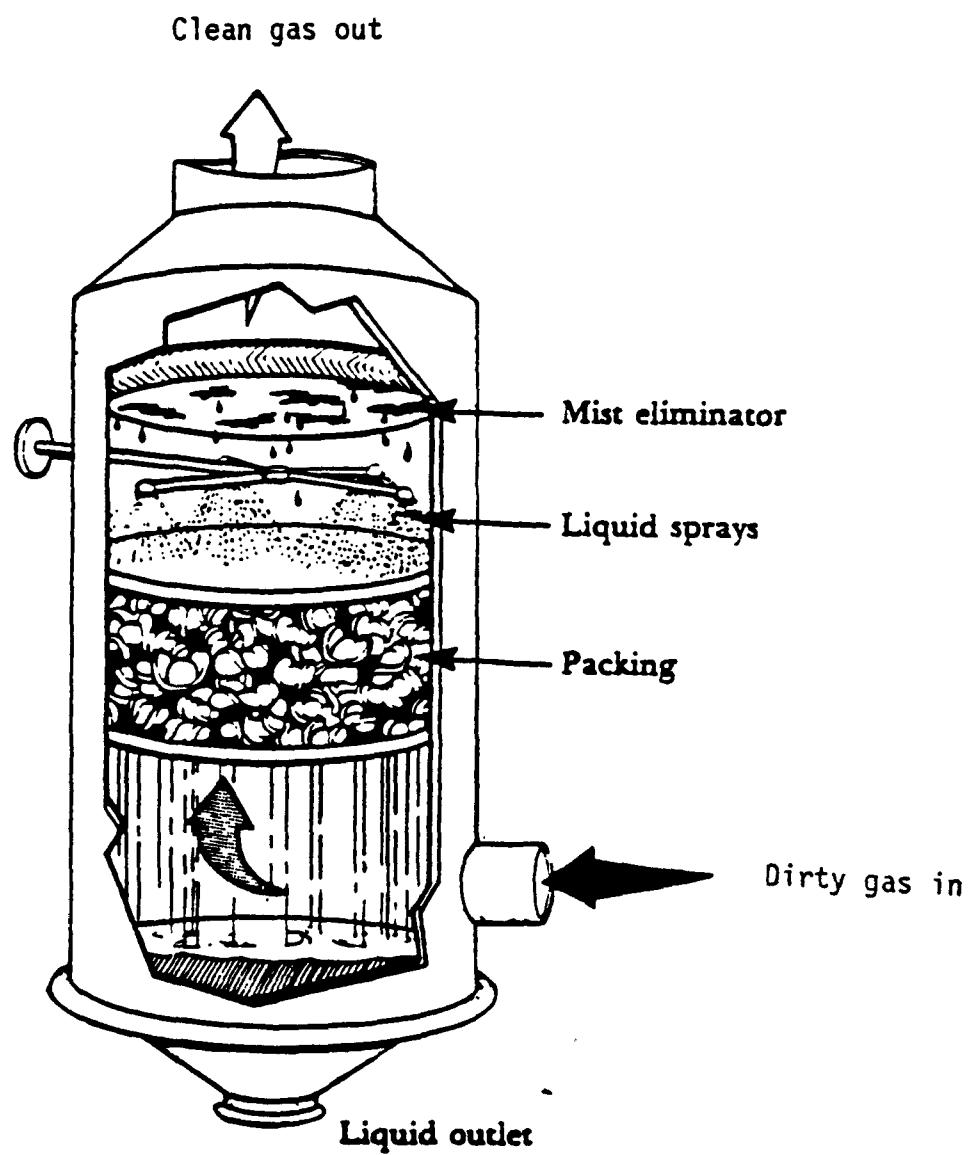
Clean gas out

Mist eliminator

Liquid sprays

Packing

Dirty gas in

Liquid outlet

Figure 3-3. Countercurrent packed tower.

gas to be treated. Liquid-to-gas ratios in packed towers generally range from 0.3 to 3 L/m^3 (2 to 20 gal/1,000 ft^3). Typical pressure drops are 0.12 to 0.25 kPa per foot of packing (0.5 to 1 inches of H_2O/ft).

Packed-tower scrubbers can achieve higher removal efficiencies than spray towers and have relatively lower water consumption requirements. Some disadvantages are high-pressure drops, more fouling potential, potentially higher maintenance cost, and wastewater disposal requirements.

Flooding is an important consideration in the design of packed towers. Flooding occurs when the gas stream approaches the flooding velocity and results in liquid being carried back up the column by the gas stream. Tower diameter should be established based on a superficial gas velocity from 50 to 75 percent of the flooding velocity.

3.2.2.2 Spray Towers.[13] Spray towers or chambers are the simplest and least expensive device employed for acid gas scrubbing. They consist of an empty tower and a set of nozzles to spray liquid. Typically, the contaminated gas stream enters the bottom of the tower and passes up through the device while the solvent liquid is being sprayed at one or more levels by nozzles. Liquid and gas streams typically flow countercurrent to each other (see Figure 3-4).

To provide a large surface for contacting the gas, nozzles are arranged to wet the entire cross section of the tower with fine liquid droplets. After falling short distances, the liquid droplets tend to agglomerate or hit the sides of the tower. This effect reduces the total liquid surface in contact with the gas stream and the residence time. Therefore, spray towers used for acid gas emission control are limited to applications where either the gases are extremely soluble or a high removal efficiency is not required.

Mass transfer is significant when droplets of 500 to 1,000 μm diameter are used.[14] In general, the smaller the droplet size and the greater the turbulence, the more chance for absorption of the gas. Production of fine droplets

3-11

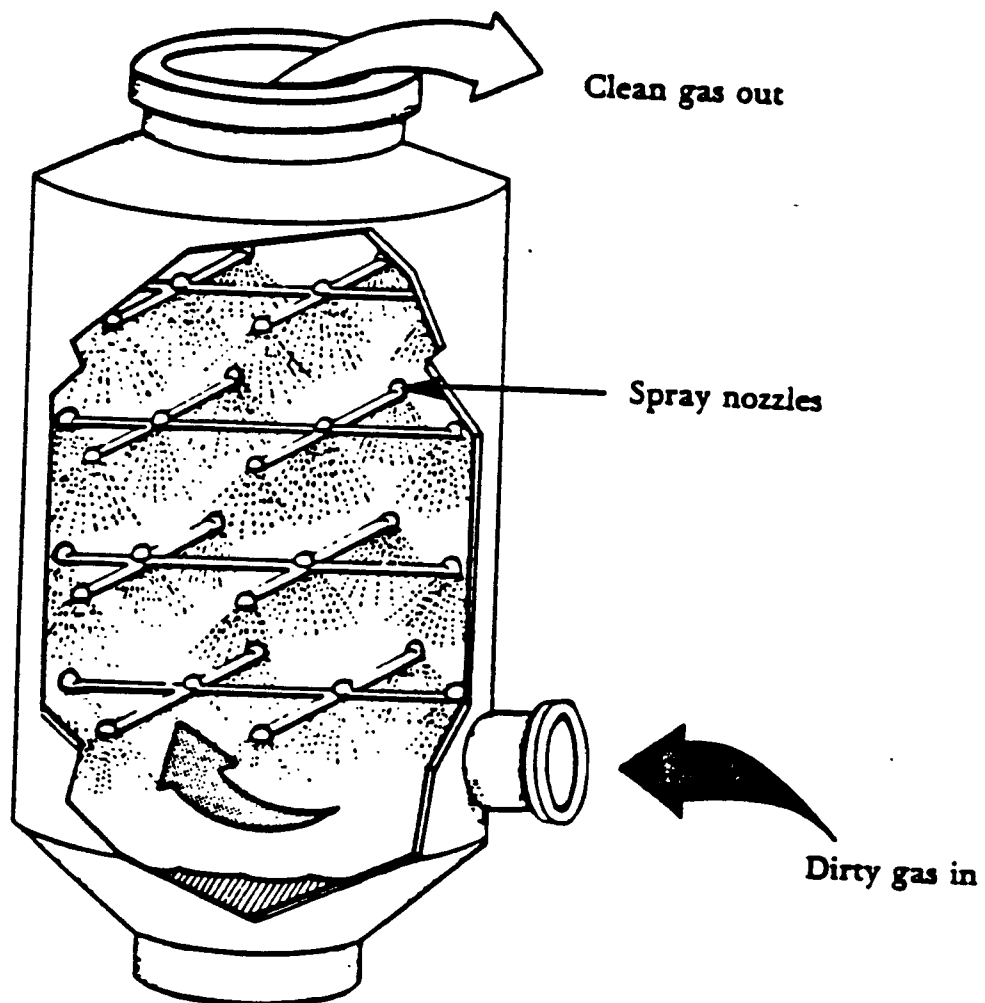Clean gas out

Spray nozzles

Dirty gas in

Figure 3-4. Spray tower.

requires the use of high-pressure spray nozzles that consume
more energy than do low-pressure nozzles.

Spray tower height has an effect on efficiency but is much
less significant than droplet size and liquid-to-gas ratios.
Increasing the height of a given tower only slightly increases
efficiency due to wall losses and agglomeration. Variations
in pressure have little effect on efficiency of these systems
while changes in liquid rate have a large effect on
efficiency. In general, spray towers are not as efficient as
packed towers but may be adequate for specific applications.

Spray towers have a low pressure drop, about 0.5 to 1 kPa
(2 to 4 inches of H_2O). Design liquid-to-gas ratios typically
average between 3 and 13 L/m^3 (20 to 100 gal/1,000 ft^3). As
with packed towers, liquid may be recirculated to reduce
requirements for makeup water.

For acid gas removal in both spray tower and packed-bed
applications, aqueous scrubbing liquids are the most common.
However, equilibrium conditions may favor the use of other
low-volatility solvents, such as nonvolatile hydrocarbon oils.

Flooding is also an important design consideration for
spray towers. As recommended for packed towers, spray tower
diameter should be established based on a superficial gas
velocity from 50 to 75 percent of the flooding velocity.

3.2.2.3 Dry Scrubbers.[15] Dry scrubbing devices are also
frequently used for controlling acid gas emissions, especially
sulfur dioxide. Spray dryers (see Figure 3-5) contact an acid
gas-contaminated stream with a finely sprayed slurry of either
lime or soda ash. The water in this slurry is evaporated by
the heat in the off-gas, yielding a dry waste stream. Dry
sorbent containing the inorganic pollutant, together with
other particulate matter in the waste gases, is captured in a
particulate matter collection system (e.g., a fabric filter
baghouse) connected to the spray dryer.

The SO_2 in the off-gas reacts with the alkali solution or
slurry by adsorption or absorption mechanisms, or both, to
form liquid-phase salts, which are dried to about 1 percent
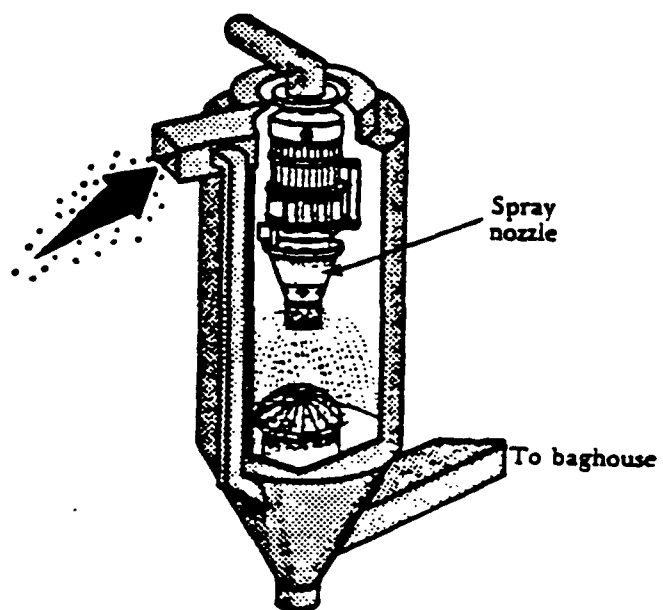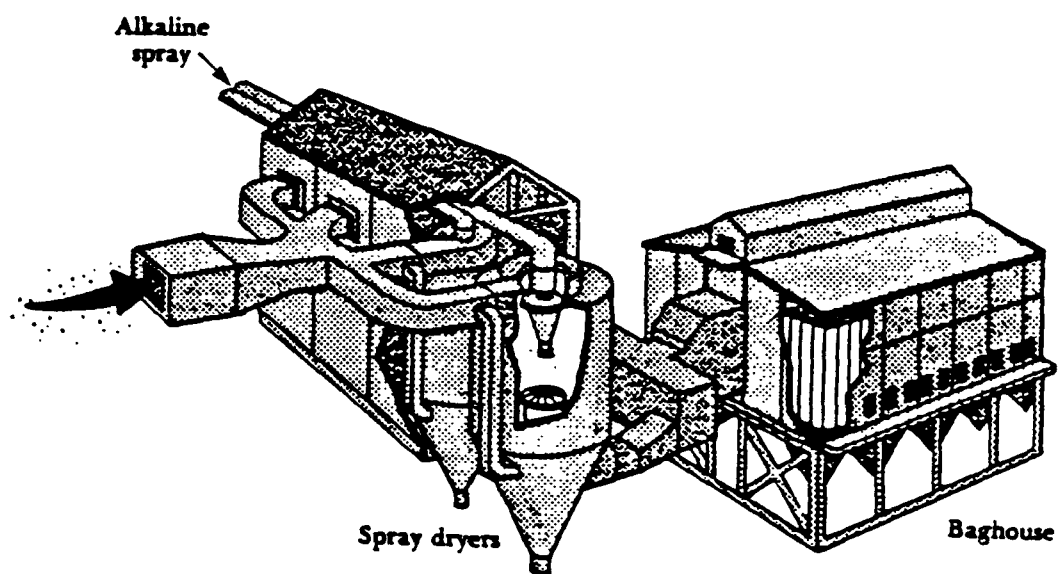
Figure 3-5.   Dry scrubbing system.
(Top: spray dryer with baghouse.   Bottom: spray dryer.)

free moisture by the heat in the off-gas. These solids are carried out of the dryer to a particulate collection device, such as a fabric filter baghouse.

In some designs, a portion of the solids is collected from the bottom of the spray dryer. Usually, the dry waste product is disposed of in landfills. To take advantage of any unreacted reagent left in the solids and reduce fresh reagent requirements, some of the solids can be recycled back to the dryer.

The technology of spray drying is less complex mechanically, and no more complex chemically, than wet calcium or wet sodium-based scrubbing systems. High SO_2 removal efficiencies, typically above 90 percent, can be obtained using either lime or sodium-based sorbents and high molar feed rates of reagent. Sodium-based reagents such as sodium hydroxide and sodium carbonate are more reactive than calcium-based reagents such as lime or limestone but are also more expensive.

3.3 PARTICULATE EMISSION CONTROL[16]

Particulates from carbon reactivation processes are mostly carbon fines and solid material formed from dry scrubbing processes used to remove acid gases. Particulate collection devices applicable for carbon reactivation off-gas include wet scrubbers (e.g., venturi, impingement, and spray towers), baghouses, cyclones, and electrostatic precipitators (ESPs). Although potentially applicable, none of the facilities identified in Secion 2 used ESPs. Therefore, no further discussion is provided regarding ESPs. When heavy particulate loads are present or submicron particulates must be recovered along with gaseous pollutants, it is common to use wet collection devices having high particulate collection efficiencies. In the case of carbon reactivation emissions where particulates are in the presence of acid gases, it is advantageous to apply control devices that also have some capability for gas absorption. Wet scrubbers, such as packed

beds, are commonly used for acid gas removal, but may trap particles effectively as well. The particulates, however, tend to clog such units too quickly to make them useful for particulate removal of gas streams with heavy particulate loads. Venturi scrubbers, as well as impingement scrubbers, are some of the more versatile wet scrubbing devices effective for both particulate and acid gas removal. This section describes each of these particulate control devices, their applicability, design, and efficiency.

3.3.1   Venturi Scrubbers

A typical venturi configuration is shown in Figure 3-6. The particulate laden off-gas enters the scrubber and is accelerated to a high velocity while passing through the converging section and approaching the throat section. The velocity of the gas is at its greatest in the throat section of the scrubber. Atomized droplets are formed by the impact of the high-velocity gas upon the liquid stream in the throat. The particulates and the liquor droplets then collide and agglomerate. After the particles have been trapped within the liquor droplets, the resulting agglomerates are easily removed from the gas stream. Although venturi scrubbers can have collection efficiencies greater than 90 percent for submicron particulates, power costs are relatively high for this device because of the high inlet gas drop. Removal efficiencies are a function of particle sizes and head loss.

Absorption of pollutant gases is limited with venturi collection because of the short time that the gas is in contact with the liquid. The venturi scrubber has, however, been used successfully to remove SO_2 and particulates from the off-gas stream.

Numerous venturi scrubbers are available with the principal differences being in:

1.   Methods for varying throat area for changes in gas flow rate;

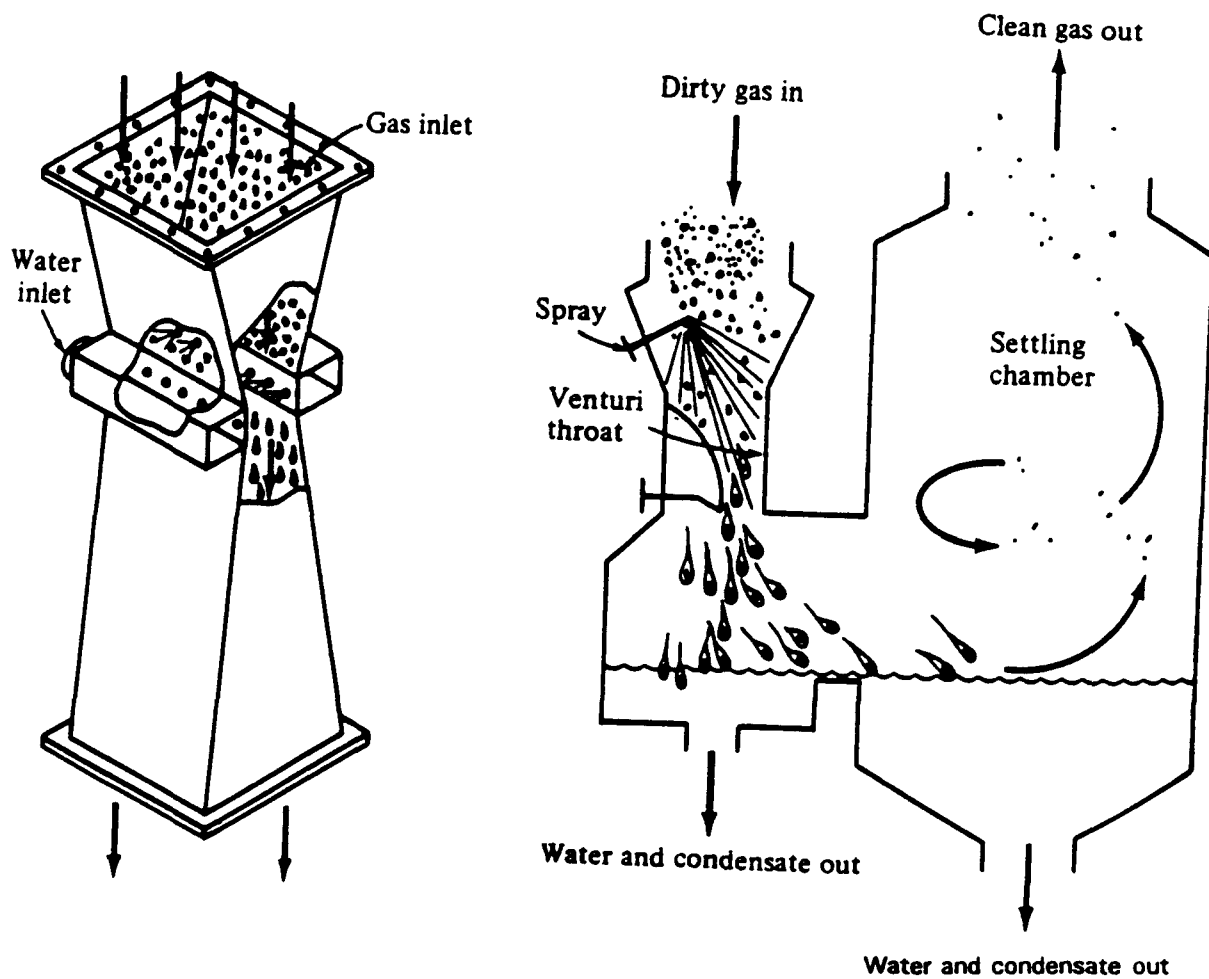2.   Methods for injecting the liquor ahead of the throat;

Figure 3-6. Venturi wet collector.

3.   Presence or absence of a diverging section; and

4.   Design of entrained liquor droplet separator following
      the throat.

Water is injected into the venturi in quantities ranging
from 0.6 to 13 L/m^3 (4 to 100 gal/1,000 ft^3 of gas).
Operating pressure drops ranging from 1.5 to 18 kPa (6 to 70
inches of water).[17] Velocities in the throat can range from
30 to 250 m/sec (100 to 800 ft/sec).

3.3.2   Impingement Scrubbers

Impingement scrubbers are also commonly employed by carbon
reactivation facilities. Although they serve primarily as
particulate control devices, the versatility of impingement
scrubbers makes them suitable for both particulate and acid
gas removal. They are similar in design to spray towers but
have one important difference: a series of baffles have been
added. The major function of these baffles is not to promote
mixing but to provide additional impingement surfaces.

The particulate-laden gas stream is forced to make many
90 degree turns around the baffles, forcing the large
particulates to impinge on the baffles rather than flow with
the air stream around the barriers. Scrubbing liquid, usually
water, is continually added above the baffles. The high
velocity of the gas stream causes the liquid to atomize into
drops that are entrained by the gas and collecting particles.
A typical unit is shown in Figure 3-7. The baffle plate has
numerous orifices and a baffle directly over each orifice. The
orifices provide additional velocity for the gas. The
entrained particles that are not collected on one baffle are
accelerated as they pass through orifices to the next stage.[18]

Particle collection is generally by inertial impaction
caused by impingement on the liquid surface and the atomized
drops. The performance of an impingement scrubber seems to be
comparable to a venturi scrubber operating at the same gas
phase pressure drop. The baffle acts as a multiple venturi.
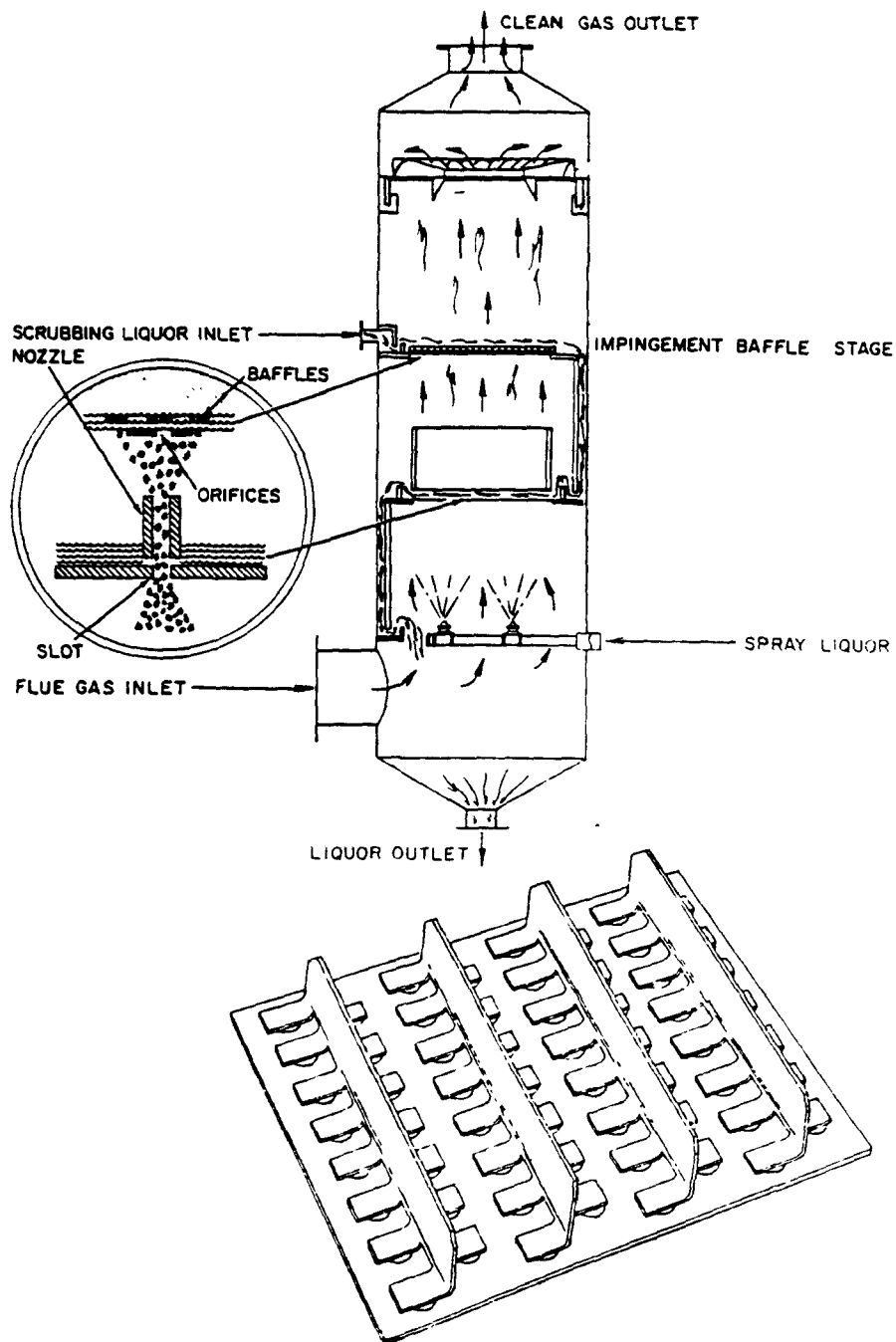One difference between this type of scrubber and a venturi is

Figure 3-7. Impingement scrubber (top);
baffle plate (bottom).

that in the venturi, all the gas is accelerated through one
opening with a resulting large pressure drop and tremendous
particulate acceleration. The impingement scrubber, however,
uses a relatively modest gas phase acceleration in which the
sum of the pressure drop across all of the orifices is less
than that in a venturi. Impingement scrubbers are, therefore,
not as effective for fine particulate removal as venturi
scrubbers.

Several baffle impingement stages may be needed to achieve
the desired degree of performance. Rarely are more than two
or three stages used due to the trade-off between energy
requirements and mass removal efficiency. Each impingement
baffle plate stage has a plate with 600 to 3,000 orifices per
square foot.

The impingement scrubber removes over 97 percent by weight
of particles above 1 μm and a considerable percentage of
smaller particles. The pressure drop through these units is
usually about 1.5 inches of water per stage. Water side
pressure is not too high, about 5 psig with superficial
velocities on the order of a few feet per second to 20 feet
per second. The gas handling capacity of these units is
easily 50,000 ft^3/min.

3.3.3 Spray Towers

Spray towers (described in Section 3.2.2.2) are effective
for dual removal of particulate and gaseous contaminants.
They can handle gases with fairly high concentrations of
particulates without plugging. The units cause very little
pressure loss and can handle large volumes of gases. As the
gas flows upward, entrained particles collide with liquid
droplets sprayed across the flow passage, and liquid droplets
containing the particles settle by gravity to the bottom of
the chamber. They are effective in removing particles in
excess of 10 μm and can be modified to improve efficiencies
for smaller particles.[19]

For all wet collectors, disposal of the wastewater in which the particulates have been collected poses problems. Often the wastewater will require some form of treatment before being charged into a receiving stream. This will typically be a settling tank or pond or a centrifugal device. In areas where water supplies are limited or water costs are unusually high, further treatment before recycling of the water may be necessary.

3.3.4 Cyclones

In industrial applications, cyclones are often used as a precleaner for the more complex air pollution control equipment such as baghouses or wet scrubbers. In carbon reactivation processes, cyclones have been used to collect fine particulates generated in the fluidized-bed reactivation process. Because fluidized-bed furnaces have a higher exhaust volume than other furnace types, more of the fines are carried over in the exhaust. Once the cyclone has removed the fines, the gas can then be further purified by the use of other emission control techniques such as afterburning.

Cyclones are centrifugal collectors that employ a centrifugal force instead of gravity to separate particles from the gas stream. Because centrifugal forces can be generated that are several times greater than gravitational forces, particles can be removed in centrifugal collectors that are much smaller than those that can be removed in gravity settling chambers.

A cyclone collector such as the one shown in Figure 3-8 consists of a cylindrical shell, conical base, dust hopper, and inlet where the dust-laden gas enters tangentially. Under the influence of the centrifugal force generated by the spinning gas, the solid particles are thrown to the walls of the cyclone as the gas spirals upward at the inside of the cone. The particles slide down the walls of the cone and into the hopper.

The cyclone has no moving parts and is simple and inexpensive. It is less efficient than a wet scrubber or a

baghouse for removing particulate matter. As a class of equipment, cyclones provide the lowest collection efficiency of devices in general commercial use for control of particles. However, their ability to handle a hot gas without appreciable cooling makes cyclones best suited for the removal of large particulates between the reactivation furnace and an afterburner.

The efficiency of a cyclone depends on the magnitude of the centrifugal force exerted on the particles. The greater the centrifugal force, the greater the separating efficiency. The magnitude of the centrifugal force generated depends on particle mass, gas velocity within the cyclone, and cyclone diameter. Efficiency will increase with an increase in dust particle size or dust particle density, gas inlet velocity, cyclone body length, and ratio of body diameter to gas outlet diameter. Conversely, efficiency will decrease with an increase in gas viscosity or gas density, cyclone diameter, gas outlet diameter, and inlet width or inlet area. Disappointing results from cyclone installations are usually due to overly optimistic estimates of efficiency. Ranges of efficiency to be expected from cyclone collector installations are shown in Table 3-1.[20]

Usually the pressure drop in cyclones varies from 1 to 6 inches of H_2O. Volumetric flows range from 50,000 to 100,000 ft^3/min, depending on the type of cyclone. Concerning temperature limitations, a cyclone is limited only by the

.

TABLE 3-1.   EFFICIENCY RANGE OF CYCLONES

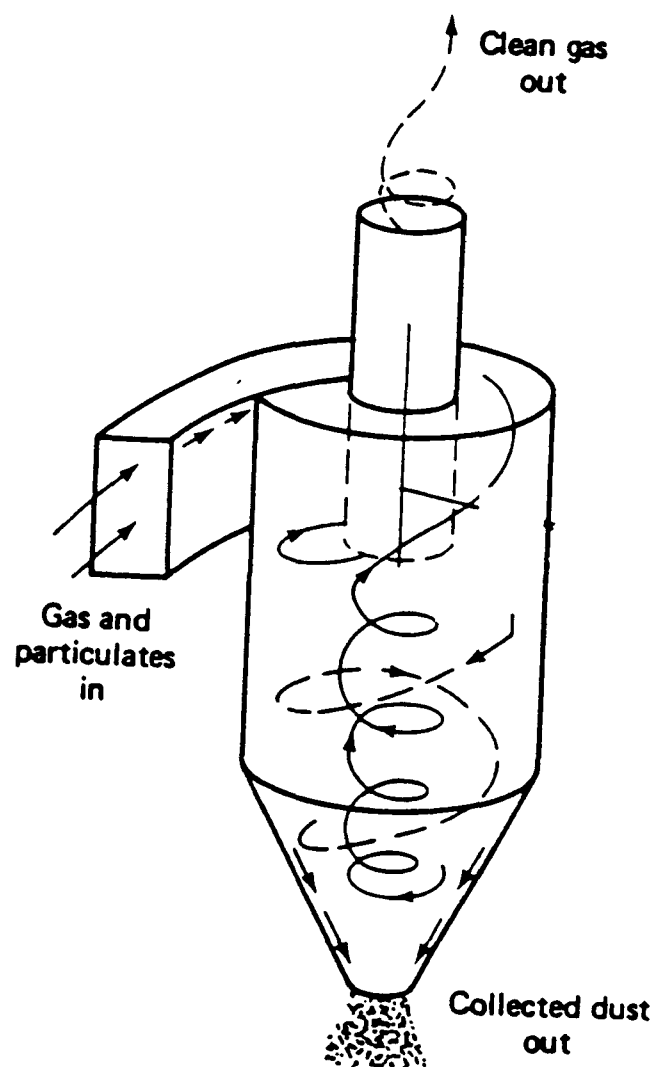| Particle size range (μm)	Efficiency range, weight % collected
Less than 5	Less than 50
5 - 20	50 - 80
15 - 40	80 - 95
Greater than 40	95 - 99

3-22

Figure 3-8. Cyclone collector.

material of which it is constructed, and it can be constructed
of almost any type of material - a distinct advantage if
condensed acids are present. Typical temperature limits are
375 to 550 °C (700 to 1,000 °F);[21] however, cyclones have been
operated at temperatures higher than 1,000 °C using refractory
linings.[22]

Potential problems encountered in operating cyclones
include erosion, fouling, and corrosion. Erosion in cyclones
is caused by the rubbing of dust particles on the cyclone
wall. This condition becomes more severe with high dust
loadings, high inlet velocities, and large or hard dust
particles. Erosion can be minimized by a wise choice of
cyclone diameter size as well as special construction
features. Such construction features include the use of
heavier-gauge metal for cones, the use of abrasion-resistant
removable wear plates at the impingement zone, avoidance of
welds and joints that cannot be ground smooth, and the use of
larger cyclones for prevention of excess gas velocities. Any
defect in cyclone design or operation that tends to
concentrate dust moving at high velocity will accelerate
erosion.

Fouling can also be a problem with cyclones. Fouling
generally occurs either through plugging of the dust outlet or
through the buildup of materials on the cyclone wall. It
results in decreased efficiency, increased erosion, and
increased pressure drop. Corrosion can become a problem if
the cyclone is operating below the condensation point when
reactive gases are present in the off-gas stream. It is
usually caused by chlorides and sulfides in contact with the
metal walls and can be minimized by using the proper
materials--alloys or, in some applications, plastics.

An advantage of cyclones, as well as bag filters, over the
wet scrubbing devices previously described is that the removed
particulates are collected as a dry stream; thus, further
wastewater treatment is not needed. Additionally, since the
primary particulate matter for carbon reactivation off-gases

is carbon fines, especially for fluidized-bed furnaces, cyclones can be used to recover the larger carbon fines and thereby reduce carbon loss during the reactivation process.

3.3.5 Baghouse Filters

Fabric filter dust collectors, commonly referred to as baghouses, are among the oldest and most widely applied particulate emission control devices. They find utility in carbon reactivation processes in several ways. One way is by collecting particulate matter formed when dry scrubbing is used to remove acid gases from the off-gas as discussed in Section 3.2.2.3. Another use for baghouse filters in carbon regeneration facilities is the removal of particulate matter from emissions leaving the reactivation furnace. The temperature of the off-gas is first lowered by a quenching process to eliminate potential fire hazards. A baghouse then removes particulate matter from the cooled gas after which the gas may go to an afterburner or scrubber.

A typical baghouse filter is shown in Figure 3-9. In fabric filter systems, particulate-laden gas streams pass through a fabric that allows gaseous matter to pass though and filters out particulate matter. A dust mat is initially formed from the retention of small particles on the fabric by direct interception, inertial impaction, diffusion, electrostatic attraction, and gravitational settling. Submicron particles are then collected more efficiently by sieving.

As particulates build up on the inside of the bag, the pressure drop across the bag increases and must eventually be relieved. This is typically done by mechanical shaking, reverse-air cleaning, or pulse-jet cleaning of the particulate layer. The particulates then fall into a collection hopper and are removed for disposal.

Several problems are associated with fabric filter use. One potential problem is the possibility of explosion or fire if sparks are discharged in a baghouse area where organic dusts are being filtered. Other problems include the

3-25

Mechanism for shaking, rapping, or vibrating bags

Clean air out

Dirty air in

Tubular filter bags

Clean air out
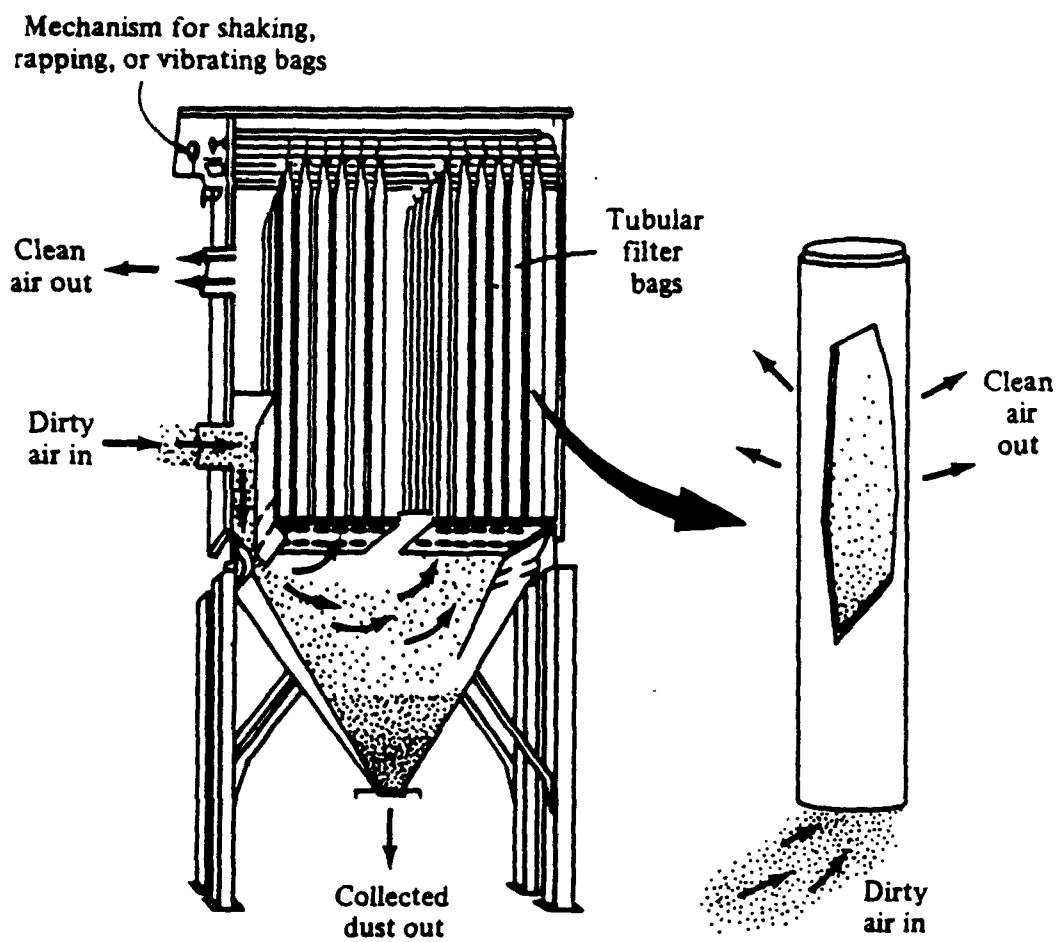
Dirty air in

Collected dust out

Figure 3-9. Baghouse filter.

possibility of rupture or other adverse effects because of temperatures too high for the fabric medium or because of the moisture, acidity, or alkalinity content of the particulate-laden gas stream.[23] Judicious fabric choice can minimize these problems.

Filter bags are usually tubular in shape. Other filter bag shapes include envelopes and pleated cartridges. As to size, a baghouse arrangement may be small enough to fit into an ordinary room or large enough to dwarf many industrial buildings.

Design of baghouses is based on filtering rates, the air-to-cloth ratios, and pressure drops. The air-to-cloth ratio is the volumetric flow rate of the gas stream divided by the surface area of the fabric. The higher the ratio, the smaller the baghouse and the higher the pressure drop. Filtering rates range from 0.5 to 5 m^3 of air per minute per square meter of cloth depending on factors such as the dust loading, fabric material, and method of cleaning.[24]

Pressure drop across the filter is a function of both the velocity of the gas through the filter and the combined resistance of the fabric and accumulated dust layer. In order to avoid operational problems and excessive power requirements, the maximum pressure across the filter should be 5 to 10 inches of water.

Advantages of using baghouses to control particles include performance that is independent of flow rate, uniform collection efficiency over a wide range of particle sizes, and collection efficiency that is independent of particle resistivity. Efficiencies in excess of 99 percent and often as high as 99.99 percent on a weight basis can be expected for well-designed systems. Most inefficiency in these units is the result of either bypass due to damaged fabric, faulty seals, or leaks. Seldom is penetration of particles through the filter cake or fabric the cause of poor performance. Usually, any penetration that does occur is greatest during or immediately after cleaning.

Drawbacks of baghouse systems include: clogging of the
filter medium due to condensation in the gas stream;
cementation of filter cake in humid, low-temperature gases,
especially in the presence of lime from a scrubber; and
excursions of high-particulate concentrations when a bag
breaks. Also, sticky or high-moisture particles can adhere to
the fabric and be difficult to release.

3.4   METAL EMISSION CONTROL

Metal emissions can occur in two forms: either as
particulate matter or as metal vapors. Metal emissions
occurring as particulate matter are primarily metals that
remain adsorbed on, or are naturally occurring in, the
entrained carbon fines. Metal particulates can also be formed
when metal vapors condense prior to the final exhaust. The
oxides of certain metals such as mercury, lead, selenium, and
cadmium have relatively high vapor pressures or volatilities
and, when they are formed in the combustion zone, a
disproportionate fraction is present in the postflame as a
vapor. A portion of this metallic vapor present in the
combustion products later condenses as the postflame gases
cool when passing through various ducts. Prior to discharge
to the atmosphere, these particles are typically controlled
using one of the particulate emission control devices
described in Section 3.3.

Only a limited number of control methods are applicable to
the remaining portion of the metallic vapor. Absorption is
the most widely used and accepted method for inorganic vapor
control. The absorption equipment used to control metal
emissions are described in Section 3.2.2. The removal
efficiency of metal vapors achievable with absorbers can be
greater than 99 percent and is typically determined by the
actual concentrations of the inorganic in the gas and liquid
streams and the corresponding equilibrium concentrations.
Water is often the ideal solvent for vapor control by
absorption. It offers distinct advantages over other

3-28

solvents, particularly its low cost. It is typically used on a once-through basis and then discharged to a wastewater treatment system. The effluent may require pH adjustment to precipitate the metals.

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4.0 ENVIRONMENTAL IMPACTS AND COST ANALYSIS

This chapter presents the emission reductions, costs, and cost-effectiveness values of various options for the control of air emissions from carbon reactivation process units. The different types of carbon reactivation furnaces are described in Section 2, and the technologies available to control the air emissions are described in Section 3. The precise sequence of air pollution control devices used depends, in part, on the type and throughput of the reactivation furnace and the types of pollutants adsorbed on the spent carbon. In this section, example control options typical to the reactivation industry are developed and cost analyses for each control option at low, medium, and high exhaust gas flow rates are presented.

4.1   CONTROL TECHNOLOGY ALTERNATIVES - MODEL SYSTEMS

As discussed in Section 3, there are three types of pollutants emitted from carbon reactivation units: organics, acid gases, and particulates. From the information collected from the reactivation industry (Tables 2-1 and 2-2), the most prevalent organic air emission control device employed is a thermal afterburner; over 70 percent of the reactivation furnaces currently operating already have thermal afterburners in place. Therefore, an afterburner is included for all model systems for control of organic air emissions.

Wet scrubbers are the next most common air pollution control device used for carbon reactivation furnaces. Approximately 50 percent of the currently operating reactivation furnaces have a wet scrubber for either acid gas

or particulate matter emission control. Venturi scrubbers are the most commonly used particulate emission control device; 37 percent of the currently operating reactivation furnaces have a venturi scrubber for particulate emission control. All other emission control devices combined (dry scrubbers, cyclones, impingement scrubbers, and baghouses) represent approximately 25 percent of the population of emission control devices used at operating carbon reactivation systems.

These industry profile data indicate that the majority of the carbon reactivation pollution control systems can be represented by two model systems. The first model system consists simply of an afterburner to control nonhalogenated organic emissions (Model Unit 1). The second model system (Model Unit 2) consists of an afterburner and a packed-bed wet scrubber to control acid gas emissions resulting from the combustion of a halogenated organic exhaust waste stream. Particulate emissions would also be controlled by the scrubber; however, these are considered negligible. For each model system, control units will be designed and their costs estimated based on three different exhaust gas flow rates representative of small, medium, and large carbon reactivation furnaces. The reactivation furnace exhaust rates investigated are: 500 scfm for small furnaces, 1,500 scfm for medium-sized furnaces, and 4,500 scfm for large furnaces.

4.2   CONTROL COSTS

With the exception of fluidized-bed furnaces, which represent only about 5 percent of the operating reactivation furnaces, different types of reactivation furnaces employ similar gas flow rates for similar carbon throughput rates. Thus, some useful generalizations can be made concerning the carbon throughput and the composition of the reactivation furnace exhaust.

Information gathered during the preliminary data-gathering activities and site visits reveals that gas flow rates range from 93 to 287 dscf in the final stack exhaust per pound of

carbon produced, with an average of 183 dscf/lb carbon.[1-8]
These measurements were taken after the afterburner. Because
of the need to add auxiliary air to supply an adequate amount
of oxygen, the final gas flow rate from the afterburner will
always be greater than the reactivation furnace exhaust gas
flow rate. Therefore, it was assumed that a gas flow rate in
the reactivation furnace was 100 dscf/lb carbon produced.
After performing the mass and energy balances, this ratio
between the reactivation furnace exhaust rate and the
afterburner exhaust rate proved to be a reasonable
approximation.

As a rule of thumb, 1 lb of steam is added per pound of
dry carbon in the furnace to create the reactivating
atmosphere.[9,10] Therefore, the moisture content in the
reactivation furnace is approximately 22 scf of steam/100 dscf
of air, based on the 100 dscf/lb carbon assumption, or
18 percent by volume. This also implies that an average gas
flow rate of 100 dscf/lb carbon is equivalent to 122 scf/lb
carbon.

The organic loading on the spent carbon can range from as
little as 1 to 2 percent for potable water treatment carbons
to as high as 25 percent depending on the application of the
carbon. Typically, however, the organic loading rate ranges
from 8 to 13 weight percent.[11,12] Assuming a 10 weight percent
organic loading on the spent carbon, a 90 percent destruction
efficiency in the reactivation furnace, and using the assumed
gas flow rate of 122 scf/lb carbon produced, the organic
concentration in the exhaust air can range from 250 ppmv for
an organic waste gas stream with an average molecular weight
of 140 g/mol to 1,500 ppmv for an organic waste gas stream
with an average molecular weight of 24 g/mol. To investigate
the influence of the organic loading of the carbon, control
costs were determined at three different organic loading rates
(5, 10, and 15 weight percent). Since the organic destruction
efficiency in the reactivation furnace also directly
influences the organic concentration in the exhaust gas

stream, two values, a low of 10 percent and a high of 90 percent, were used in the model plant design and cost analyses. Sample calculations of the design and cost of each of the model units is provided in Appendix A.

4.2.1 Model Unit 1: An Afterburner

The OAQPS Control Cost Manual[13] was used to investigate different operating conditions and estimate the control costs for Model Unit 1. In this case, a thermal incinerator is used as an afterburner for the control of nonhalogenated organic air emissions. The control cost for an afterburner (thermal incinerator) is largely driven by the total volumetric gas flow rate through the afterburner. This flow rate is the sum of the waste gas flow rate (exhaust from the reactivation furnace), the fuel feed flow rate, and the auxiliary air flow rate needed to maintain the desired oxygen concentration.

From the information gathered from facilities operating carbon reactivation furnaces, the reactivation furnace is typically operated with a minimum of excess oxygen; furnace exhaust oxygen concentrations on the range of 1 to 8 percent are typical of the industry. The afterburner is typically operated with slightly more excess oxygen than the reactivation furnace; afterburner exhaust oxygen concentrations range from 6 to 10 percent.[14-17] Therefore, it is necessary to add auxiliary air to maintain the oxygen concentration in the afterburner.

To minimize the amount of auxiliary air required to be added in the afterburner, the amount of fuel burned needs to be minimized. The waste gas is already very hot (550 to 800 °C), so a preheater for the waste gas is not required. However, the auxiliary air required to maintain sufficient oxygen in the afterburner is presumably at ambient air temperature (25 °C or 77 °F), and a significant amount of the total energy requirement for the afterburner is needed to heat the auxiliary air. For these anticipated gas stream temperatures, a 35 percent efficient heat recovery preheater for the auxiliary air reduces the afterburner fuel requirement

by approximately a factor of four and reduces the auxiliary
air requirement by approximately a factor of two. The
preheater increases the total capital investment (TCI) for the
afterburner compared to a system with no heat recovery, but
this increase in TCI is small because the total afterburner
gas flow rate is less in the system with a preheater.
Additionally, the large decrease in the fuel requirements for
the system with a preheater leads to a lower total annual cost
(TAC) for a system with a preheater compared to a system
without one.

The energy balance and the oxygen mass balance dictate the
design of the afterburner. Because the reactivation furnace
off-gas is a dilute organic waste stream, its composition and
concentration have limited effect on the design of the
afterburner. That is, a compositional change in the furnace
off-gas, although it may provide some impact on the fuel
requirements, will not significantly alter the total gas
throughput in the afterburner and will not, therefore,
significantly affect the TCI. Changes in the fuel requirement
will, however, directly affect the fuel costs, and the fuel
costs become a significant contributor to the TAC for larger
systems, as the total gas flow increases.

The operating parameters used to develop the control costs
are summarized in Table 4-1. These parameters were determined
based on the information gathered from facilities operating
carbon reactivation furnaces, the generalizations discussed in
Section 4.2, and preliminary cost estimates to determine the
optimal heat recovery factor.

The results of the control cost analyses for Model Unit 1
are summarized in Table 4-2 (refer to Appendix A for sample
calculations). At a given concentration, increasing the
exhaust gas flow rate for larger reactivation furnace
throughput causes an increase in the TCI and TAC, but not
proportionally to the increase in the flow rate. Therefore,
the cost effectiveness of the afterburner control device

TABLE 4-1.   DESIGN PARAMETERS FOR THERMAL INCINERATOR

Design parameter	Input value
For reactivation furnace exhaust	
Furnace exhaust flow rate (varied)	
Small furnace throughput	500 scfm
Medium furnace throughput	1,500 scfm
Large furnace throughput	4,500 scfm
Benzene concentration (varied)	
5% org./lb carbon and 90% destruction	216 ppmv
10% org./lb carbon and 90% destruction	456 ppmv
15% org./lb carbon and 90% destruction	725 ppmv
5% org./lb carbon and 10% destruction	1,950 ppmv
10% org./lb carbon and 10% destruction	4,110 ppmv
15% org./lb carbon and 10% destruction	6,520 ppmv
Furnace exhaust temperature	1,100 °F
Furnace exhaust oxygen concentration	4% by volume
Furnace exhaust moisture content	18% by volume
For afterburner	
Fuel inlet temperature	77 °F
Auxiliary air inlet temperature (after 35% heat recovery)	1,500 °F
Afterburner design temperature	1,700 °F
Afterburner exhaust oxygen concentration	8% by volume
Energy recovery (used to preheat aux. air)	35%
Afterburner destruction efficiency	99%

TABLE 4-2. CONTROL COSTS AND COST EFFECTIVENESS FOR MODEL UNIT 1: AFTERBURNER[a]

Compound	Furnace exhaust rate (scfm)	Conc. (ppmv)	Afterburner exhaust rate (scfm)	TCI (1,000 $)	TAC (1,000 $)	Uncontr. emiss. (Mg/yr)	Contr. emiss. (Mg/yr)	Cost effect. ($/Mg)
Benzene	500	216	870	146	82	5	0.05	17,500
		456	860	146	81	10	0.10	8,210
		725	850	145	80	16	0.16	5,090
		1,950	800	143	75	43	0.43	1,770
		4,110	710	138	66	90	0.90	730
		6,520.[b]	760	141	64	143	1.43	450
	1,500	216	2,610	195	133	14	0.14	9,410
		456	2,580	194	130	30	0.30	4,360
		725	2,550	193	126	48	0.48	2,670
		1,950	2,400	190	111	128	1.28	880
		4,110	2,130	184	84	270	7.70	320
		6,520.[b]	2,280	188	80	429	4.29	190
	4,500	216	7,840	259	268	43	0.43	6,350
		456	7,750	258	260	90	0.90	2,910
		725	7,650	258	250	143	1.43	1,760
		1,950	7,200	253	205	384	3.84	540
		4,110	6,380	245	127	811	8.11	160
		6,520.[b]	6,830	250	113	1,290	12.9	90

TCI = Total capital investment.

TAC = Total annual cost.

[a]All costs determined in 1989 U.S. dollars using a 7% capital recovery.

[b]Concentration exceeds 25% of lower explosive limit; additional dilution air was required.

improves with the increased gas flow rate. At a given exhaust gas flow rate, increasing the exhaust gas organic concentration, either through higher organic loading on the spent carbon or through less combustion of desorbed organics in the reactivation furnace, has a very limited effect on the TCI and TAC. Therefore, the cost effectiveness of the afterburner control device improves in proportion to an increase in gas organic concentration.

4.2.2   Model Unit 2:   An Afterburner and Wet Scrubber

For reactivation furnace exhaust air streams that contain halogenated organics, a wet scrubber is needed to remove the acid gases generated as the halogenated organics are incinerated. A chapter of the OAQPS Control Cost Manual is currently being drafted to estimate the costs of a packed-tower wet scrubber. Although the final draft of that chapter was not available when this document was written, the cost curves to determine the equipment costs for the tower were available.[18]

The design inputs for the afterburner in Model Unit 2 use the same exhaust gas flow rates, the same organic carbon loading levels, and the same reactivation furnace organic destruction efficiencies as Model Unit 1 (see Table 4-1). In this case, however, the organic pollutant is 1,1,1-trichloroethane and the exhaust from the afterburner contains HCl, which must be removed. The design for the packed-tower wet scrubber was, therefore, based on the afterburner exhaust gas flow conditions. The total afterburner exhaust gas flow rates were 880 scfm, 2,640 scfm, and 7,910 scfm, corresponding to the reactivation furnace exhaust flow rates of 500 scfm, 1,500 scfm and 4,500 scfm, respectively. The afterburner exhaust was used to preheat the auxiliary air, so the afterburner exhaust temperature exiting the heat exchanger was 593 °C (1,100 °F). Due to the relatively high solubility of HCl in water, the tower design was driven by the energy balance (i.e., the amount of water required to cool the gas stream). The sizing and costing of

the packed-tower wet scrubber was then determined using the design and cost information provided in References 19 through 22. Table 4-3 summarizes the design parameters for Model Unit 2.

The calculated control costs for Model Unit 2, the combined afterburner and packed-tower wet scrubber emission control train, are summarized in Table 4-4 (see Appendix A for sample calculations). The same relative effects caused by changing the exhaust gas flow rate and the exhaust gas organic concentration that were seen in Model Unit 1 are found in Model Unit 2. That is, increasing the exhaust gas flow rate, at a set organic concentration, increases the TCI and the TAC but lowers the cost-effectiveness factor. Increasing the organic pollutant concentration at a set gas flow rate does not affect the TCI or the TAC and, therefore, proportionally lowers the cost-effectiveness factor.

4.3 CROSS-MEDIA AND SECONDARY ENVIRONMENTAL IMPACTS

Although thermal incinerators are used to prevent air pollution, thermal incinerators themselves exhibit an air pollution potential and can produce cross-media impacts and secondary emissions depending on their operation. It is possible to create a major pollution problem when burning any waste gas if the waste gas is not burned with sufficient air for combustion. Improper operating conditions can result in carbon monoxide (CO) generation. If proper combustion conditions are observed, such as correct air-to-fuel ratio, sufficient mixing, adequate residence time, peak flame temperature, and proper cooling rate of combustion products, essentially all the carbon present in the waste gas should end up as carbon dioxide (CO_2). All the hydrogen should result in water as a product of combustion, and unburned hydrocarbons should be minimal.

TABLE 4-3.   DESIGN PARAMETERS FOR MODEL UNIT 2:   THERMAL
INCINERATOR AND WET SCRUBBER

Design parameter	Input value
For reactivation furnace exhaust	
Furnace exhaust flow rate (varied)	
Small furnace throughput	500 scfm
Medium furnace throughput	1,500 scfm
Large furnace throughput	4,500 scfm
1,1,1-Trichloroethane concentr'n (varied)	
5% organic/lb carbon and 90% destruction	127 ppmv
10% organic/lb carbon and 90% destruction	267 ppmv
15% organic/lb carbon and 90% destruction	425 ppmv
5% organic/lb carbon and 10% destruction	1,140 ppmv
10% organic/lb carbon and 10% destruction	2,410 ppmv
15% organic/lb carbon and 10% destruction	3,820 ppmv
Furnace exhaust temperature	1,100 °F
Furnace exhaust oxygen concentration	4% by volume
Furnace exhaust moisture content	20% by volume
For afterburner	
Fuel inlet temperature	77 °F
Auxiliary after air inlet temperature (after 35% heat recovery)	1,500 °F
Afterburner design temperature	1,700 °F
Excess oxygen (exhaust O_2 concentration)	8% by volume
Energy recovery (used to preheat aux. air)	35%
Afterburner destruction efficiency	99%
For afterburner exhaust/wet scrubber feed	
Afterburner exhaust flow rate for:	
Small furnace throughput	880 scfm
Medium furnace throughput	2,640 smfm
Large furnace throughput	7,910 scfm
Afterburner exhaust temperature (i.e., after 35% heat recovery)	1,100 °F
For wet scrubber	
Liquid inlet temperature	70 °F
Liquid outlet temperature	200 °F
Gas outlet temperature	100 °F
Liquid-to-gas ratio (calculated based on energy balance)	2:1 by weight (6.1 gal/1,000 ft^3)
Packing material	1½-in. berl saddles
Scrubber removal efficiency	98%

TABLE 4-4. CONTROL COSTS AND COST EFFECTIVENESS FOR MODEL UNIT 2: AFTERBURNER AND WET SCRUBBER[a]

Compound	Furnace exhaust Rate (scfm)	Conc. (ppmv)	Afterburner exhaust rate (scfm)	TCI (1,000 $)	TAC (1,000 $)	Uncontr. emiss. (Mg/yr)	Contr. emiss. (Mg/yr)	Cost effect. ($/Mg)
1,1,1-TCE	500	127	880	183	127	5	0.05	27,000
		267	880	184	127	10	0.10	12,800
		425	880	184	127	16	0.16	8,070
		1,140	880	183	127	43	0.43	3,000
		2,410	870	183	126	90	0.90	1,410
		3,820	870	183	126	143	1.43	890
	1,500	127	2,640	275	191	14	0.14	13,500
		267	2,640	276	191	30	0.30	6,410
		425	2,640	277	191	48	0.48	4,040
		1,140	2,630	275	189	128	1.28	1,490
		2,410	2,610	275	188	270	2.70	700
		3,820	2,600	275	186	429	4.29	440
	4,500	127	7,910	439	361	43	0.43	8,550
		267	7,910	440	361	90	0.90	4,050
		425	7,910	442	361	143	1.43	2,550
		1,140	7,880	438	358	384	3.84	940
		2,410	7,840	439	354	811	8.11	440
		3,820	7,790	439	349	1,290	12.9	270

TCI = Total capital investment.
TAC = Total annual cost.
1,1,1-TCE = 1,1,1-trichloroethane.
[a]All costs determined in 1989 U.S. dollars using a 7% capital recovery.

NO_x emissions from thermal incinerators are another secondary pollutant impact produced by thermal incinerators. Nitrogen oxides have two sources: nitrogen in the fuel and the reaction between atmospheric nitrogen and oxygen at high temperatures. One problem encountered in reducing emissions from combustion sources is the fact that modifications that reduce carbon monoxide and hydrocarbon emissions generally increase NO_x emissions and vice versa.

As noted above, some adverse effects on air quality can be associated with the use of combustion devices to control organic emissions from spent carbon reactivation units. Pollutants generated by the combustion process--SO_2, CO, and particularly NO_x--may have an unfavorable impact on ambient air quality. One facility was identified that has tested for NO_x emissions. Three 1-hr emission tests were conducted on the final stack, subsequent to an afterburner, venturi scrubber, and impingement scrubber, at a carbon reactivation system processing 41 Mg/day (3,750 lb/hr) of carbon. The NO_x concentrations ranged from 20 to 27 ppmv on a dry basis, and the NO_x emission rates ranged from 1.7 to 2.3 Mg/yr (0.44 to 0.61 lb/hr).[23]

There is increasing concern about HCl emissions from incinerators owing to the growing amount of halogenated polymers, such as polyvinyl chloride (PVC), and halogenated solvent, such as methylene chloride, used in chemical processes and therefore potentially present in waste streams. In addition, HF emissions arise from the combustion of fluorinated hydrocarbons. Water scrubbing, as used in Model Unit 2, appears to be an effective means of controlling these acid gases. An increase in wastewater, however, will result from the wet scrubbing of these gases. The scrubber wastewater may require neutralization by addition of a caustic before being released into the wastewater treatment and disposal system. The salts produced from the neutralization, though small, may need to be disposed of as hazardous waste.

TABLE 4-5.   ANNUAL ENERGY REQUIREMENTS FOR CONTROL SYSTEMS

Control system/ design air flow rate	Natural gas consumption (thousand ft^3/yr)	Electricity consumption (MWh/yr)
Model Unit 1[a]		
500 scfm	715-5,850	18.8-23.1
1,500 scfm	2,145-17,550	56.4-69.3
4,500 scfm	6,440-56,600	169-208
Model Unit 2[b]		
500 scfm	5,690-6,090	28.7-29.1
4,500 scfm	17,100-18,300	86.0-87.2
4,500 scfm	51,200-54,800	258-262

[a]Model Unit 1 is an afterburner, with operating conditions as provided in Table 4-1. Increasing the concentration of benzene in the reactivation furnace exhaust causes a decrease in both the natural gas consumption rate and the electricity consumption rates up to 25% of the LEL for benzene (4,375 ppm).

[b]Model Unit 2 is an afterburner and packed-tower wet scrubber system, with operating conditions as provided in Table 4-3. Increasing the concentration of 1,1,1-trichloroethane in the reactivation furnace exhaust causes only a slight decrease in the natural gas consumption rate and has less than a 1% effect on the electricity comsumption rate.

Therefore, carbon reactivation may produce a small solid waste impact, but this impact is much smaller than the solid waste impact caused by directly disposing of the spent carbon rather than reactivating the spent carbon.

The use of an afterburner to control organic emissions from carbon reactivation furnaces results in a net energy usage for both model units because supplemental fuel is needed to support combustion. These supplemental fuel requirements in the form of natural gas are presented in Table 4-5. Electrical energy is required to operate the pumps, fans, blowers, and instrumentation that may be necessary to control organics using an afterburner. Electrical energy requirements as estimated by the costing model are presented in Table 4-5 as well. The natural gas consumption varies only slightly with an afterburner of a given size due to the low organic concentrations in the gas stream. Model Unit 2 requires approximately twice the electrical energy, primarily due to the increased pressure drop caused by the packed-tower wet scrubber.

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CARBON REACTIVATION ACT

APPENDIX A

CALCULATION METHODOLOGY FOR COST ANALYSIS

This appendix provides detail regarding the methodologies used to calculate the cost effectiveness of emission controls applied to carbon reactivation furnace off-gas. For the afterburner (thermal incinerator), the cost calculations follow the example provided in Chapter 3 of the OAQPS Control Cost Manual (4th Edition, EPA 450/3-90-006). The major difference between the calculations presented here and those presented in the OAQPS Control Cost Manual (or simply the Cost Manual) for thermal incinerators is that the reactivation furnace exhaust (inlet waste gas for the afterburner) does not contain an adequate amount of oxygen to sustain combustion of the auxiliary fuel (assumed inlet $[O_2]$ = 4 vol%). Therefore, auxiliary air was required to provide the oxygen necessary for combustion of the fuel and the organics in the waste gas.

The cost calculations for the wet scrubber follow the example provided in Chapter 4.7 of the Handbook: Control Technologies for Hazardous Air Pollutants (EPA/625/6-91/014) (referred to hereon as the HAP Handbook). The main difference between the calculations presented here and those presented in the HAP Handbook for wet scrubbers is that the scrubbing water flow rate was driven more by the energy balance than the solubility of the acid gas (HCl in this case).

It may be useful to read (or have available) copies of the Cost Manual and HAP Handbook prior to (or while) studying the sample calculations provided in this appendix.

<u>Step 1</u>. Establish design specifications

From Table 4-1 of this document.

Waste Gas Volumetric Flow Rate	Q_w = 500, 1500, 4500 scfm
Waste Gas Temperature	T_w = 1,100°F
Oxygen Content	[O2] = 4 vol. %
Chemical Composition*	[Bz] = 216, 456, 725 ppm
Heating Value (from Table 3-14)	-Δhc, Bz = 17,446 Btu/lb
	-Δhc,CH_4 = 21,502 Btu/lb
Particulate Content	Not critical for thermal units
Design Control Efficiency	99%
Combustion Chamber Outlet	1700°F
Percent Energy Recovery	35%
Moisture Content	18 vol. %

*The chemical composition of the waste gas is calculated from the organic loading of the spent carbon based on the assumptions outlines in Section 6. Specifically, the gas flow rate is assumed to be 122 scf/lb carbon <u>produced</u>, and 90% of the organics are assumed to be destroyed in the reactivation furnace. Thus, given a 10 wt% organic loading there will be 10 lbs of benzene per 100 lbs of spent carbon or 10 lbs of benzene per 90 lbs of actual carbon (that can be produced). Then the concentration of benzene is:

$$10\% \ organic \ loading = \frac{0.1 \ lb \ Bz}{0.9 \ lb \ Carbon} * \frac{1 \ lb \ Carbon}{122 \ scf} * (1 - 0.9) = 9.1 \ E{-}5 \ 16 \ Bz/scf$$

At standard conditons and noting MWt_{Bz} = 78 g/mol

$$9.1 \ E{-}5 \ \frac{lb \ Bz}{scf} * \frac{454 \ g}{lb} * \frac{mol}{78 \ g} * \frac{24.4 \ L}{mol} * \frac{0.0353 \ scf}{L} = 4.56 \ E{-}4 \ scf/scf$$

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Step 2. Verify oxygen requirements

Since the waste gas oxygen concentration is only 4%, auxiliary air is needed to provide adequate oxygen to burn the fuel. The design thermal incinerator exhaust oxygen concentration if 6%. Oxygen requirements are:

for Methane $CH_4 + 2O_2 = CO_2 + 2 H_2O$ 2 moles O_2/mol CH_4

for Benzene $C_6H_6 + 7.5O_2 = 6CO_2 + 3 H_2O$ 7.5 moles O_2/mol C_6H_6

(Note: assuming ideal gas, a mole balance is equivalent to volumetric balance.)

Oxygen Balance

O2 In	Auxiliary Air = 0.21 Qair
	Waste Gas = 0.04 Qw
O2 Out	Waste Gas = 0.08 (Qair + Qw + Qf)
O2 Consumption	for Methane = 2 Qf
	for Waste Gas = 7.5 [Bz] Qw = 0.0034 Qw (for [Bz] = 456 ppm)

$$In = Out + Consumption$$

$$0.21 \; Qair + 0.04 \; Qw = 0.08 \; (Qair + Qw + Qf) + 2 \; Qf + 0.0034 \; Qw$$

$$Qair = 0.334 \; Qw + 16 \; Qf \qquad \text{Eqn. 1}$$

Step 3. Determine % of LEL

LEL for Benzene = 14,000 ppmv (from Table 3.12 of Cost Manual)

    % LEL = 456/14000 * 100% = 2.3%

Which is well below the desired 25% LEL for fire safety.

<u>Step 4</u>. Calculate the volumetric heat of combustion for waste gas (Δhw)%.

$$-\Delta hc,\ Bz = 17{,}446\ Btu/lb\ *\ \left(\frac{1\ lb}{454\ g}\right)\left(\frac{782}{mol}\right)\left(\frac{1\ mol}{0.861\ scf}\right)$$

$$-\Delta hw = -\Delta hc,Bz\ *\ [Bz] = 3{,}480\ *\ 456\ *\ 10^{-6} = 1.59\ Btu/scf$$

As the waste gas has a moisture content of 18 vol%, the average molecular weight and density of waste gas is :

$$\rho air = 29\ g/mol\ *\ \frac{1\ lb}{454\ g}\ *\frac{1\ mol}{0.861\ scf} = 0.0742\ lb/scf$$

$$\rho H_2O = 18\ g/mol\ *\ \frac{1\ lb}{454\ g}\ *\ \frac{1\ mol}{0.861\ scf} = 0.04605\ lb/scf$$

$$\rho w = 0.18\ \rho H_2O + 0.82\ \rho air = 0.0691\ lb/scf$$

$$\therefore\ -\Delta hw = (1.59\ Btu/scf)/(0.0691\ lb/scf) = 23.0\ Btu/lb$$

<u>Step 5</u>. Establish temperature for the incinerator.

From the survey of the industry, and afterburner temperature of 1700 °F was selected.

<u>Step 6</u>. Calculate the waste gas temperature at the exit of the preheater.

As the reactivation furnace exhaust is typically over 1000 °F to start, preliminary cost estimates discovered that most of the energy requirement for the afterburner was due to the heating of the auxiliary air when the auxiliary air was assumed to be at 77 °F. If was subsequently determined that it was more cost efficient to pre-heat the auxiliary air than to pre-heat (or further heat) the waste gas. It was assumed that the air could be pre-heated to 1500 °F. This assumption will be checked in Step 9.

<u>Step 7</u>. Calculate the auxiliary fuel requirement (Q_f).

For this example, the determination of the auxiliary fuel requirement requires that the energy and oxygen balances be simultaneously balanced.

$$\rho w = 0.0691\ lb/scf$$

$$\rho air = 0.0742\ lb/scf$$

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$$\rho H_2O = 0.0460 \ lb/scf$$

$$Cpw = \frac{0.50 \ (0.18) \ 0.046 + 0.26 \ (0.82) \ 0.0742}{(1 \ scf) \ (0.0691)}$$

$$Cpw = 0.289 \ Btu/lb \ °F$$

$$Tref = 77 \ °F$$

$$Tdesign = 1700 \ °F$$

$$Tave = 900 \ °F$$

$$Cpair = 0.26 \ Btu/lb \ °F$$

$$CpH_2O = 0.50 \ Btu/lb \ °F$$

$$\rho air \ Cpair = 0.0193 \ Btu/scf - °F$$

$$\rho w \ Cpw = 0.0200 \ Btu/scf - °F$$

Note: The product of the density * heat capacity for the afterburner exhaust (waste stream) and the auxiliary air are nearly identical. As these terms always appear as products in the heat balance, it is reasonable to assume that the density * heat capacity for the afterburner exhaust is the same as the waste gas'.

The energy balance is: In - Out + Generation = 0

		Value (Btu/min)

Heat In $\rho QCp(T-Tref)$

Auxiliary Air	$= 0.0193 \ Qair \ (1500-77)$	$= 27.46 \ Qair$
Waste Gas	$= 0.020 \ Qw \ (1100-77)$	$= 20.46 \ Qw$
Auxiliary Fuel	$= 0$ (since inlet temperature = reference temperature	

Heat Out $\rho QCp(T-Tref)$

| Waste Stream | $= 0.020 \ (Qw + Qair + Q_f) \ (1700-77)$ | $= 32.46 \ (Qw + Qair + Q_f)$ |

Heat Out → heat loss = 10% of the total energy input (heat out waste stream) $= 3.25 \ (Qw + Qair + Q_f)$

Heat Generation $(-\Delta h)\rho Q$

Waste Gas	$= (23.0)(0.0691)Q_w$	$= 1.59\ Q_w$
Auxiliary Fuel	$= (21,502)(0.0408)Q_f$	$= 877.3\ Q_f$

Heat Balance Equation:

$$27.46\ Q_{air} + 20.46\ Q_w - 35.71\ (Q_w + Q_{air} + Q_f) + 1.59\ Q_w + 877.3\ Q_f = 0$$

Solving for Qair yields:

$$Q_{air} = 102\ Q_f - 1.656\ Q_w \qquad\qquad Eqn.\ 2$$

Solving Equation 1 and 2 simultaneously and using Qw = 1500 scfm yields

$$0.334\ Q_w + 16\ Q_f = 102\ Q_f - 1.656\ Q_w$$

$$Q_f = 34.7\ scfm$$
$$Q_{air} = 1,055\ scfm$$
$$Q_{tot} = 2,590$$

Step 8. Verify sufficiency auxiliary fuel.

Want heat generated by the fuel to equal or exceed 5% of the total energy input.
Energy input = 32.46 (2,590)   = 84,070 Btu/min
Auxiliary fuel = 877.3 (34.7)   = 30,440 Btu/min

∴ auxiliary fuel provides 36% of energy input, and is therefore sufficient to stabilize the flame.

Step 9. Check total volumetric flow and pre-heater energy balance.

Qtot = 2,590 scfm; Max available heat = 84,070 Btu/min
Heat required to pre-heat air to 1500 °F = 27.46 (1,055) = 28,970 Btu/min
That is a heat recovery of 34.5%. Therefore, the assumed inlet temperature of the air is approximately correct.

Note: Pre-heating the waste stream (instead of the air) does not reduce the total afterburner volumetric flow rate as much as the air stream preheater due to the oxygen balance requirements. As the cost equations provided in the OAQPS Control Costs Manual were not developed for a dual preheater system, the 35% heat recovery (used for preheating air) cost equations used. It may be possible to pre-heat the waste gas, and then pre-heat the air to recover more of the exhaust streams sensible heat energy. To solve the equations for this option requires estimates of inlet temperatures, calculating total gas flow rates, calculate heat recoveries based on the calculated flow rates, estimating new inlet gas temperatures and re-iterating the process until the assumed inlet temperatures equal the actual inlet temperatures for a given level of heat recovery. Then, to determine the most cost effective design, calculate the cost of the systems for each level of heat recovery. This process was done for a single pre-heater system, and the optimal design is presented here. A 50% heat recovery system was more than was needed to pre-heat the air (i.e., could not use the full 50% heat recovery), and subsequently, it did not yield lower costs.

Step 10. Calculate Equipment Cost (EC).

From Equation 3.25 of the Cost Manual:

$$Equipment\ Costs\ =\ EC\ =\ 13,149 Q_{TOT}^{0.2609}$$

$$From\ Step\ 7,\ Q_{TOT}\ =\ 2,590\ scfm\qquad \therefore\ EC\ =\ \$102,200$$

Step 11. Calculate Total Capital Investment (TCI) using Table 3.8 of the Cost Manual and assuuming no auxiliary equipment is required:

$$Purchased\ Equipment\ Cost\ =\ PEC\ =\ 1.18\ EC$$
$$PEC\ =\ \$120,600$$

$$Direct\ Installation\ Cost\ =\ 0.3\ PEC\ =\ \$36,200$$
$$Site\ preparation\ \rightarrow\ not\ required$$
$$Buildings\ \rightarrow\ not\ required$$

$$Therefore,\ Total\ Direct\ Cost\ =\ DC\ =\ 1.30\ PEC\ =\ 156,800$$

$$Total\ Indirect\ Costs\ =\ IC\ =\ 0.31\ PEC\ =\ 37,400$$

$$Total\ Capital\ Investment\ =\ TCI\ =\ DC\ +\ IC\ =\ \$194,200$$

<u>Step 12</u>. Calculate power requirement for fan.

From Table 3.11 of the Cost Manual, the total pressure drop of the system is:

$$\Delta P = \Delta P_{incin} + \Delta P_{Heat\ Ex} = 4 + 4 = 8\ in\ H_2O$$

From Equation 3.37 of the Cost Manual,

$$Power_{fan} = (1.17\ x\ 10^{-4}\ Q_{g,\ a}\ \Delta P)/\in$$

At 1,100 °F $\qquad Q_{g,\ a} = Q_{TOT}\left(\dfrac{460 + 1100}{537}\right) = 2590(2.905) = 7,524\ scfm$

where $\in$ = fan efficiency = 0.6 (assumed)

$$\therefore\ Power_{fan} = 11.74\ kW$$

<u>Step 13</u>. Calculate annual utility costs.

Actual operating hours are assumed to be 24 hours/day, 7 days/wk, 48 wks/yr.
For utility consumption, operating hours are: 8,064 hrs/yr

Electricity cost = $0.059/kWhr * 11.74 kW * 8064 hrs/yr = $5,600/yr

Natural Gas Cost = $3.30/1000 ft^3 * 34.7 scfm * 60 min/hr * 8064 hrs/yr = $55,400/yr

<u>Step 14</u>. Calculate labor and material costs.

Annual operating days = 7 days/wk * 48 wk/yr = 336 days

Operating Labor = $12.96/hr(0.5 hr/shift)(3 shifts/day)(336 days/yr)
$\qquad\qquad\qquad$ = $6,530

Supervisory Labor = 15% of operating labor = $980

Maintenance Labor = $14.26/hr(0.5 hr/shift)(3 shifts/day)(336 days/yr)
$\qquad\qquad\qquad\qquad$ = $7,190

Material = 100% of Maintenance Labor = $7,190

Step 15. Calculate direct annual cost (DAC)

Direct annual cost (DAC) equals the sum of utility, labor, and material costs

DAC   = $5,600 + 55,400 + 6,530 + 980 + 7,190 + 7,190
       = $82,900/yr

Step 16. Calculate Indirect Annual Costs (IAC).

Overhead = 60% of (labor + materials)
        = 0.6 (6,530 + 980 + 7,190 + 7,190)
        = $13,100

Administration, Property Tax, + Insurance   =   4% TCI
                                              = $7,800

Capital Recovery (7% interest; 10 yrs life) = 0.1424 TCI = $27,600

IAC   =   Overhead + Administration + Taxes + Insurance + Capital Recovery
         =   13,100 + 7,800 + 27,600
         = $48,500/yr

Step 17. Calculate Total Annual Cost (TAC)

TAC = DAC + IAC = 82,900 + 48,500
TAC = $131,000/yr

Step 18. Calculate Total Annual Emissions.

Design efficiency of Afterburner is 99%.

$$Uncontrolled\ Emissions = Qw\ [Bz]\ (\rho Bz)\ 60\ min/hr * 8064\ hr/yr$$

$$= 1500\ scfm\ (456 \times 10^{-6}) \left(\frac{78\ E-6\ Mg}{0.861\ scf} \right) 60 * 8064$$

$$= 30\ Mg/yr$$

$$Controlled\ Emissions = Uncontrolled\ Emissions * (1 - Destruction\ Efficiency)$$

$$= 30\ (0.01) = 0.30\ Mg/yr$$

Step 19. Calculate Cost Effectiveness.

The cost effectiveness is calculated as:

$$Cost\ effectiveness = TAC/Emission\ Reduction$$
$$= 131,000/(30 - 0.30)$$
$$= \$4,400/Mg$$

Design and Cost of Incinerator/Wet Scrubber System

To control the air emissions from the reactivation of spent carbon containing halogenated organics, and incinerator/wet scrubber system was designed. The design and cost of the incinerator follow the example for benzene containing waste gas previously provided.

The oxidation reaction of 1,1,1-trichloroethane (TCE)

$$CH_3\ CCl_3 + 2O_2 \rightarrow 2CO_2 + 3HCl$$
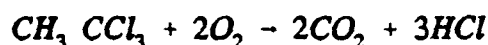
Therefore, for every mole of 1,1,1-TCE oxidized, 3 moles of HCl are produced.

At 10% organic loading, the maximum amount of 1,1,1-TCE is:

$$\frac{0.1\ g\ TCE}{0.9\ g\ carbon} \left(\frac{1\ lb\ carbon}{122\ scf}\right) \left(\frac{454\ g}{lb}\right) \left(\frac{1\ mol\ TCE}{133.1\ g}\right) = 3.1\ x\ 10^{-3}\ mol\ TCE/scf$$

$\therefore$ Max Amount HCl = $9.3\ x\ 10^{-3}$ mol HCl/scf

The reactivation furnace off-gas, 90% destruction efficiency is:

$$[HCl] = 0.9\ (9.3\ x\ 10^{-3}\ mol) \left(\frac{0.861\ scf}{mol}\right) = 7,200\ ppmv$$

Using the reactivation system with a 1,500 scfm flow rate, and the properties of 1,1,1-TCE (i.e., MWt_{TCE} = 133.1 g/mol; -Δhc = 938.9 Btu/lb), the following gas flow rates are calculated:

$$Qw = 1,500\ scfm$$
$$Q_f = 38\ scfm$$
$$Qair = 1,099\ scfm$$
$$Q_{TOT} = 2,637\ scfm$$

Note: additional fuel was required for the 1,1,1-TCE waste gas than for the Benzene waste gas due to 1,1,1-TCE's lower heat of combustion.

Assuming 99% destruction in the after burner, the afterburner exhaust HCl concentration is:

$$HCl = [7{,}200 + 10^6\ (0.99)\ (0.1)\ (9.3\ x\ 10^{-3})\ 0.861] * \left(\frac{1{,}500}{2{,}637}\right)$$
$$= 4{,}550\ ppmv$$

Thus, the auxiliary air diluted the concentration of HCl. Nonetheless, the total mass flow rate of HCl is higher in the afterburner exhaust than the reactivation furnace exhaust.

Two factors were considered when designing the packed tower: 1) needed to ensure sufficient water flow rate to absorb the HCl; and 2) needed to use sufficient water to cool the hot gas stream. As the solubility of HCl is temperature dependent, the scrubber water exiting the packed tower needed to be low enough to achieve adequate absorption.

Step 1a. Calculate the inlet gas temperature.

Energy needed to heat air stream to 1500 °C is

$$E_a = Q_{\rho a}\ C_{pn}\ (T - T_i) = 1{,}099(0.0193)(1500 - 77)$$
$$= 30{,}180\ Btu/min$$

Energy available from the waste stream is:

$$E_T = Q_{TOT}\ \rho w\ C_{pw}\ (T - T_i) = 2{,}637\ (0.0200)\ (1700 - 77)$$
$$= 85{,}600\ Btu/min$$

Energy available from a 35% recovery heat exchanger:

$$E_{35\%T} = 29{,}960\ Btu/min$$

Therefore, the actual auxiliary air temperature entering the afterburner is 1490 °F, which was close enough to the assumed value of 1500 °F that no further iterations were performed. The waste stream temperature exiting the heat exchanger was calculated using the 35% heat recovery value as follows:

$$E_{35\%T} = 29{,}960 = Q_{TOT}(\rho C_{pa})(1700 - T_g)$$
$$T_g = 1700 - \frac{29{,}960}{2{,}637(0.0200)}$$
$$T_g = 1{,}130\ °F$$

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<u>Step 2a</u>. Calculate Liquid flow rate using energy balance.

Due to the relatively high solubility of HCl in water, the design approach used was to:
1) assume desired outlet water and gas temperatures; 2) determine the liquid flow rate needed
to achieve that temperature, then check to determine if sufficient water is being used to
achieve the desired control efficiency.

$$\begin{aligned}
\textit{Using} \qquad T_g &= T_{g,i} = 1{,}130\ °F \\
\textit{and Assuming} \qquad T_{L,i} &= \quad 77\ °F \\
T_{L,o} &= \quad 200\ °F \\
T_{g,o} &= \quad 100\ °F
\end{aligned}$$

(Wanted this cool to minimize
safety hazard of gas exhaust)

For water @ 140 °F (average temperature):

$$\rho_L = 0.9832\ g/cm^3 = 61.43\ lb/ft^3$$

$$C_{pL} = \quad 1.05\ Btu/lb\ °F$$

$$\rho_L\ C_{pL} = 64.5\ Btu/ft^3\ °F$$

Energy needed to cool to gas stream:

$$\begin{aligned}
E_g &= Q_g\ \rho_g\ C_{pg}\ (T_{g,i} - T_{g,o}) \\
&= 2{,}637(0.0200)(1{,}130-100) \\
&= 54{,}322\ Btu/min
\end{aligned}$$

The liquid flow rate required to match that energy requirement is:

$$\begin{aligned}
Q_L &= \frac{E_g}{\rho_L\ C_{pL}\ (T_{L,o} - T_{L,i})} \\
&= \frac{54{,}322}{[64.5(200-77)]} \\
&= 6.85\ ft^3/min
\end{aligned}$$

To check the sufficiency of this flow rate, the calculation provided in HAP Control Technology Handbook, Section 4.7, are used. First, need to calculate the molar flow rates:

$$MWt_g = 3\% \ CO_2 + 13\% \ H_2O + 4\% \ O_2 + 80\% \ N_2 = 27.3 \ lb/lb \ mol$$
$$= 6.67 \ lb \ mol/min$$

$$L = Q_L \rho_L / MWt_L = 6.85(61.43)/(18)$$
$$= 23.4 \ lb-mol/min$$

Step 3a. Check liquid flow rate using equilibrium absorption data

To calculate the slope of the equilibrium curve, the following data was calculated from the data given in Perry's Chemical Engineer Handbook, 5th Edition, p. 33-98. (Note: Partial pressure of HCl is $760*1400^{-6} = 1.06$ mm Hg.)

Solubility For Given Temperature and Partial Pressure

Temperature	P_v = 1.06 mm Hg	P_v = 0.053 mm Hg (95% removal)
50 °C (122 °F)	21 g/100 g H_2O	96 g/100 g H_2O
80 °C (176 °F)	12 g/100 g H_2O	3 g/100 g H_2O

$$M = \left(\frac{mol \ fraction \ in \ gas}{mol \ fraction \ in \ liquid \ at \ equilibrium} \right) = \frac{y}{x}$$

$$MWt \ HCl = 36.46 \ g/mol$$

$$At \ 80 \ °C, \ Pv = 1.06 \ mm \ Hg: \quad X = \frac{(12/36.46)}{(12/36.46 + 100/18)} = 0.056$$

$$Y = 1.4 \ x \ 10^{-3}$$

$$M = \frac{Y}{X} = 1.4 \ x10^{-3} = 0.025$$

Of the conditions presented, this is the largest value of m. To ensure sufficient liquid flow for absorption want:

$$L > 1.6 \ m \ G$$
$$L = 23.4 \ lb-mol/min > 1.6(0.025)6.67 = 0.27 \ lb \ mol/min$$

Therefore, the liquid flow rate is more than sufficient to achieve high HCl removal efficiencies (>98%).

Step 4a. Calculate the Column Diameter using Figure 4.7-2 of HAP Handbook.

First calculate the abscissa (ABS):

$$ABS = \frac{(L \; MWt_L)}{(G \; MWt_g)} \left(\frac{\rho_g}{\rho_L}\right)^{0.5}$$

The average gas temperature is 600 °F, therefore

$$\rho_g = 0.0691 * \left(\frac{460 + 777}{460 + 600}\right) = 0.035 \; lb/acf$$

$$ABS = \left(\frac{421 \; lb/min}{182 \; lb/min}\right) \left(\frac{0.035}{61.43}\right)^{0.5}$$
$$= 2.313(0.0239)$$
$$= 0.055$$

From Figure 4.7-2, the ordinate (ORD) is:
$$ORD = 0.14$$

$$G_{a,f} = \left\{\frac{ORD \; \rho_g \; \rho_L \; gc}{[(a/e^3) \; \mu_L^{0.2}]}\right\}^{0.5}$$

$$For \; 1\frac{1}{2}'' \; Berl \; Saddles: \quad a = 44$$
$$e = 0.75$$
$$e^3 = 0.422$$

The gravitational constant, g_c, is: 32.2 ft/sec² ●

$$Viscosity \; of \; water \; is: \quad \mu_L = 0.495 \; cp \; @ \; 140 \; °F \; (Perry's \; p. \; 3\text{-}213)$$
$$\mu_L^{0.2} = 0.869$$

$$G_{a,f} = \left[\frac{0.14(0.035)(61.43)(32.2)}{(44/0.422)(0.869)}\right]^{0.5}$$
$$G_{a,f} = 0.327 \; lb/ft^2\text{-sec}$$

A-14

Assuming a fraction of flooding velocity, f, of 0.60

$$G_a = f \, G_{a,f} = 0.60(0.327)$$

$$G_a = 0.196 \; lb/ft^2\text{-sec}$$

The column area is: $\quad A_c = \dfrac{G \; MWtg}{(60 \; sec/min)(G_a)}$

$$A_c = \dfrac{(182 \; lb/min)}{(60 \; sec/min)(0.196)}$$

$$A_c = 15.46 \; ft^2$$

The column diameter is then:

$$D_c = \left[\left(\frac{4}{\pi}\right)A_c\right]^{0.5} = 1.13 \; A_c^{0.5} = 4.44 \; ft \quad \sim 4.5 \; ft$$

<u>Step 5a</u>. Calculate column height.

$$H_{pack} = N_{oG} \, H_{oG}$$

Design for 98% removal: $(HAP_e/HAP_o) = 50$

For the flow rates used, the worst (largest) m occurs at the water outlet

$$(Temp = 200 \; °C; \; X_{sat} = \dfrac{8_g \; HCl}{100 \; g \; H_2O} = 0.038 \; moles/mole; \; m = 0.037)$$

$\therefore (1/AF) = 0.037$.

Then $N_{oG} = 3.91$ (from Eqn 4.7-13 HAP Handbook)

$$H_{oG} = H_G + (\frac{1}{AF}) \, H_L$$

$$H_G = [b(3600G_a)^c/(L')^d] \; (Sc_g)^{0.5}$$

For $1\frac{1}{2}"$ *Berl Saddles:* $\quad b = 5.05$

$$c = 0.32$$
$$d = 0.45$$

$$L'' = 60L(MWt_L)/A_c = 60(421)/(15.46) = 1,634 \; lb/hr\text{-}ft^2$$

Schmidt number $= Sc_g = \mu_g/(\rho_g \, D_{AB})$

@600 °F μ_g = 0.0287 cp (from Perry's p.3-21) = 2.87 x 10^{-4} g/cm-sec

(315 °C or 588 °K) ρ_g = 0.035 lb/ft³ = 5.61 x 10-4 g/cm³

$$D_{AB} = \frac{BT^{3/2} \, (1/M_1 + 1/M_2)^{0.5}}{\rho r_{1,2}^2 \, I_D} \qquad \textit{(Perry's p. 3-231)}$$

$$B = [10.85 - 2.5(1/M_1 + 1/M_2)] \; x \; 10^{-4} \qquad M = Molecular \; Wts$$

$$= [10.85 - 2.5 \, (1/27.3 + 1/36.46)^{0.5}] \; x \; 10^{-4}$$

$$= [10.85 - 2.5 \, (0.253)] \; x \; 10^{-4}$$

$$= 1.02 \; x \; 10^{-3}$$

$T^{3/2} = (588)^{1.5} = 14,258$

$P = 1 \; atm$

$$r_{12} = \frac{(r_o)_1 + (r_o)_2}{2} = \frac{3.6117 + 3.305}{2} \qquad \textit{(from Perry's Table 3-308 p. 3-234)}$$

$r_{12} = 3.461$

$r_{12}^2 = 11.98$

To calculate I_D: $\in/K_1 = 97.0$; $\in/K_2 = 360$ (from Perry's Table 3-308 p. 3-234)

$$\epsilon_{12}/K = [(\epsilon/K_1)(\epsilon/K_2)]^{0.5} = (97.360)^{0.5} = 187$$

$$KT/\epsilon_{12} = \frac{588}{187} = 3.14$$

From Table 3-309 Perry's p. 3-234

$I_D = 0.4688$

$D_{AB} = 1.02 \; x \; 10^{-3} \, (14,258) \, (0.253)/[1(11.98) \, (0.4688)]$

$D_{AB} = 0.655 \; cm^2/sec$

$$Scg = \frac{2.87 \times 10^{-4}}{[(5.61 \times 10^{-4})(0.655)]}$$
$$= 0.781$$

$\therefore H_G = [5.05[(3600)0.196]^{0.32}/(1634)^{0.45}] \, (0.781)^{0.5}$

$H_G = 1.30$

$H_L = Y(L''/\mu_L)^s \, (Sc_L)^{0.5}$

For $1\frac{1}{2}''$ *Berl Saddles:* $Y = 0.00625$

$s = 0.28$

(*From Perry's Fig 3.45, p. 3-213*) $\mu_L = 0.495 \, cp = 4.95 \times 10^{-3} \, g/cm\text{-sec} \ @ \ 140 \ {}^\circ F$

$\mu_L = 1.197 \, lb/ft\text{-}hr$

$Sc_L = (\mu_L/\rho_L \times D_{AB})$

$T = 60{}^\circ F \ (T = 333 \ {}^\circ K)$

$$D_{AB} = 8.931 \times 10^{-10} \, T\left(\frac{\ell_+ \, \ell_-}{\Lambda^o}\right)\left(\frac{Z_+ + Z_-}{Z_+ Z_1}\right) \qquad (\textit{Eqn 3-34 Perry's p. 3-235})$$

For H^+ $\ell_+ = 349.8 + 4.816\Delta T = 1.031 \times 10^{-2}(\Delta T)^2 - 7.67 \times 10^{-5}(\Delta T)^3$

$\Delta T = 60 - 25 = 35$

$\therefore \ell_+ = 502.4$

For Cl^- $\ell_- = 76.35 + 1.54 \, \Delta T = 4.65 \times 10^{-3}(\Delta T)^2 = 1.28 \times 10^{-5}(\Delta T)^3$

$= 135.4$

$\Lambda^o = \ell_+ + \ell_- = 637.8$

$Z_+ = Z_- = 1$

$$D_{AB} = 8.931 \ 10^{-3}(333)\frac{(502.4)(135.4)}{637.8}(2)$$

$D_{AB} = 6.344 \times 10^{-5} \, cm^2/sec$

$$Sc_L = \frac{4.95 \times 10^{-3} \, g/cm\text{-sec}}{0.9832 \, g/cm^3 \ast 6.344 \times 10^{-5} \, cm^2/sec}$$

$Sc_L = 79.4$

$$\therefore H_L = 0.00625 \left(\frac{1634}{1.197}\right)^{0.28} (79.4)^{0.5}$$

$$H_L = 0.42$$

$$\therefore H_{OG} = 1.3 + \left(\frac{1}{AF}\right) (0.42) = 1.32$$

$$H_{pack} = N_{OG} \, H_{OG}$$

$$H_{pack} = 3.91(1.32)$$

$$H_{pack} = 5.2 \, ft$$

Step 6a. Calculate volume of packing.

$$V_{pack} = \pi/4 \; (D_c)^2 \; (H_{pack})$$
$$V_{pack} = 82.7 \, ft^3$$

Step 7a. Calculate Pressure Drop through column.

$$P_a = g \; x \; 10^{-8} \left[10^{(rL'/\rho_L)}\right] (3600 \; G_{area})^2/\rho_g$$

For $1\frac{1}{2}''$ *Berl Saddles:* $\quad g = 8.10$

$$r = 0.0025$$

$$P_a = 8.1 \; x \; 10^{-8} \left[10^{(0.0025)(1634)/(61.43)}\right] (705.6)^2/0.035$$

$$P_a = 1.343 \, lb/ft^2\text{-}ft$$

$$P_{TOT} = P_a \, H_{pack} = 1.343(5.2)$$

$$P_{TOT} = 6.98 \, lb/ft^2$$

$$P_{TOT} = 1.34 \, inches \; H_2O$$

Step 8a. Calculate the surface area of the column.

$S = \pi D_c (H_{TOT} + D_c/2)$ *Eqn 2 of Vatavuk memo*

$H_{TOT} = 2.81 + 1.4 \, H_{pack} + 1.02 \, D_c$ *Eqn 3 of Vatavuk memo*

$\qquad = 2.81 + 1.4(5.2) + 102(4.5)$

$\qquad = 14.68 \, ft$

$S = \pi(4.5) \; (15 + 4.5/2)$

$S = 244 \, ft^2$

<u>Step 9a</u>. Calculate Column Cost.

$$C_{col} = 115S = \$28,060 \qquad\qquad\textit{Eqn 1 of memo}$$

<u>Step 10a</u>. Calculate equipment costs. EC.

$$EC = C_{col} + C_{pack} + C_{aux.}$$

(Using Porcelain Flexisaddles for Berl Saddle = 1½" in Table 4.7-2 of HAP Handbook)

$$C_{pack} = V_{pack} \, (CF)_{pack}$$

$$(CF)_{pack} = \$17.75/ft$$

$$C_{pack} = 82.7(17.75)$$

$$C_{pack} = \$1,470$$

$$C_{aux} = \text{Cost Auxiliary Equipment.}$$

Assuming 40 feet of ductwork is needed, then using Equation 4.12-6 (p. 4-100) of HAP Handbook.

$$d_{duct} = 12\left[\left(\frac{4}{\pi}\right)\left(\frac{Q_{g,\,a}}{U_{duct}}\right)\right]^{0.5} = 0.3028 \, Q_{ea}^{0.5}$$

This equation assumes a duct velocity (U_{duct}) of 2,000 ft/min.
At 1,100 °F:

$$Q_{g,\,a} = Q_{TOT}\left(\frac{460 + 1100}{460 + 77}\right) = 2,637 \; scfm(2.905)$$

$$Q_{g,\,a} = 7,660 \, ft^3/\text{min}$$

$$d_{duct} = 26.5 \; inches$$

$$P_{FRP} = 24\left(\frac{d_{duct}}{12}\right) = \$53/ft$$

$$C_{aux} = 40 \, ft \, (\$53/ft) = \$2,120$$

$$EC = 28,060 + 1,470 + 2,120$$

$$EC = \$31,650$$

Step 11a. Calculate total capital investment (TCI)

Purchased equipment cost (PEC) is:
$$PEC = 1.18EC = \$37,350$$

Direct Cost (DC) assuming no site preparation and no building is:
$$DC = 1.85PEC = \$69,100$$

Indirect cost (IC) is:
$$IC = 0.35PEC = \$13,070$$

Then, $\qquad TCI = DC + IC$
$$TCI = \$82,200$$

Step 12a. Calculate power requirement for fan from Step 7a:
$$Power = 1.17 \times 10^{-4} \; Q_{g,a} \; \Delta P/\epsilon$$

From Step 7a: $\Delta P \approx 1.5$ *in* H_2O *and using* $\epsilon = 0.6$ *then,*
$$Power = 1.95 \times 10^{-4} \; (7660)1.5$$
$$= 2.24 \; kW$$

Step 13a. Calculate annual utility costs.

$$Electricity \; cost = \$0.059/kW\text{-}hr * 2.24 * 8,064$$
$$= \$1,070$$

$$Water \; cost = \$0.20/1000 \; gal \; (6.85 \; ft^3/min) \left(\frac{60 \; min}{hr}\right) \left(\frac{7.48 \; gal}{ft^3}\right) * 8,064 \; hr$$
$$= \$4,960$$

Step 14a. Calculate labor and material costs.

These are the same as Step 14 for the afterburner.

$$
\begin{aligned}
Operating \; Labor &= \$6,530 \\
Supervisory \; Labor &= 980 \\
Maintinance \; Labor &= 7,190 \\
Material &= 7,190
\end{aligned}
$$

Step 15a. Calculate direct annual cost (DAC).

$$DAC = \text{sum of utility, labor, and material costs}$$
$$= 1,070 + 4,960 + 6,530 + 980 + 7,190 + 7,190$$
$$= \$27,920$$

Step 16a. Calculate indirect annual cost (IAC).

$$Overhead = 0.6(labor + materials) = \$13,100$$
$$Admin, Taxes, and\ Insurance = 4\%TCI = \quad 3,300$$
$$Capital\ Recovery\ (7\%;\ 10\ yrs) = 0.1424TCI = \quad 11,700$$
$$IAC = \text{sum of the above} = \quad \$28,100$$

Step 17a. Calculate total annual cost.

$$TAC = DAC + IAC$$
$$= 27,920 + 28,100$$
$$= \$56,000/yr$$

Step 18a. Calculate TCI and TAC for afterburner/wet scrubber system.

$$TCI_{sys} = TCI_a + TCI_{ws}$$
$$= 194,000 + 82,000$$
$$= \$276,000$$

$$TAC_{sys} = TAC_a + TAC_{ws}$$
$$= 135,000 + 56,000$$
$$= \$191,000/yr$$

Step 19a. Calculate cost effectiveness.

$$\text{Emission reduction of } VO = 30.0 - 0.30 = 29.7\ Mg/yr$$

$$Cost\ Effectivenss = TAC_{sys}/Emission\ Reduction$$
$$= 191,000/29.7$$
$$= \$6,400/Mg$$

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(Please read instructions on the reverse before completing)

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16. ABSTRACT

The purpose of this Alternative Control Technology (ACT) document is to provide technical information to address air emissions of volatile organic compounds (VOC) from carbon reactivation processes, some of which are subject to RCRA regulations. This document contains technical information on carbon reactivation process operations, air emission rates, control technologies, and environmental and cost impacts of alternative control technologies.

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