Chapter 2

Incinerators and Oxidizers

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2.1 Introduction

In this context, the terms "incineration" and "oxidation" refer to several different thermal treatments of organic substances in waste materials. The term incineration is generally used to describe a process for the combustion of solid and liquid wastes, such as hazardous, medical, municipal, or sewage waste. With respect to gaseous waste streams containing volatile organic compounds (VOCs) and/or organic hazardous air pollutants (HAP), the terms "incinerator" and "oxidizer" are often used interchangeably and generally refer to the use of thermal or catalytic oxidizers.[1] The U.S. Environmental Protection Agency defines any organic compound to be a VOC unless it is specifically determined to have negligible photochemical reactivity.[2] Indeed, a number of commonly used organics (e.g., acetone, methane, and methylene chloride) are specified as not being VOCs and some non-VOC organic compounds (e.g., methylene chloride) are listed as hazardous air pollutants pursuant to section 112(b)(2) of the Clean Air Act. This distinction is important since emissions of VOCs and HAP are regulated, while both VOC and non-VOC organic compounds are combustible and are therefore important in the design of the incinerator or oxidizer. For convenience, we use the term "VOC" in the remainder of the chapter to refer to both VOC and volatile organic HAP.

Incineration, like carbon adsorption, is one of the best-known waste treatment methods for industrial gas. Carbon adsorption allows recovery of organic compounds that may have value as commodity chemicals. In contrast, however, incineration is an ultimate disposal method in that the combustible compounds in the waste gas are destroyed rather than collected. A major advantage of incineration is that virtually any gaseous organic stream can be incinerated safely and cleanly, provided proper engineering design and management are used. In some applications, waste heat from the oxidizer can be recovered and used in other processes or converted to electric power.

The main types of thermal oxidizers are direct fire, catalytic, recuperative, and regenerative. Historically, the most commonly used is the regenerative thermal oxidizer (RTO), although recuperative thermal oxidizers are becoming more common (ICAC, 2016). Table 2.1 provides capital cost estimates for thermal oxidizers in several industry source categories.

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[1] Incinerators should not be confused with flares. Flaring is a combustion control process for in which the gases are piped to a remote, usually elevated, location and burned in an open flame in the open air using a specially designed burner tip, auxiliary fuel, and steam or air to promote mixing for nearly complete destruction. For more information on flares, please review the Flares chapter in the EPA Air Pollution Control Cost Manual.

[2] Volatile organic compound (VOC) is defined in 40 CFR 51.100, which also provides a list of organic compounds that are excluded because they have been determined to have negligible photochemical reactivity.

Table 2.1: Summary of Cost Data for Individual Incinerators and Oxidizers

Source Category	Unit Type	Flow to Incinerator	Units	Capital Cost ($)	Total Annualized Cost ($)	Year	Comments	Reference
Portland Cement	RTO	502,312	dscfm	25,280,000	7,970,758	2010	Cost based on costs of an RTO installed on an existing kiln. Total capital investment includes direct (DC) and indirect costs (IC) estimated based on Cost Manual.	EPA 2010
Plywood Composite - MIN	RTO	7,587	dscfm	NA	358,359	2002	Estimated cost based on RTO cost algorithm developed using: (1) information provided by an RTO vendor with numerous RTO installations at PCWP plants, and (2) the Control Cost Manual. The RTO cost algorithm was used to determine RTO total capital investment (TCI) and total annualized cost (TAC) based on the exhaust flow to be controlled and annual operating hours of 8,000 hours per year.	EPA 2002
Plywood Composite - MAX	RTO	79,483	dscfm	NA	599,447	2002		
Plywood composite	RTO	50,000	dscfm	924,699	NA	1997	Purchased equipment cost based on data provided by vendor.	
		100,000	dscfm	1,350,204	NA			
		200,000	dscfm	2,201,214	NA			
Aerospace	RTO w/out Concentrator	60,000	acfm	NA	347,282	2014	RTO quote from Epcon. Cost were in $/ton	3
Aerospace	RTO w/ Zeolite Concentrator	60,000	acfm	NA	367,276	2014	Concentrator quote from Anguil. Cost were in $/ton	
Spray Finishing	RTO	200,000	scfm	3,300,000	1,050,000	~2008		4
Spray Finishing	RTO w/ concentrator	12,000	scfm	2,500,000	300,000	~2008		
Spray Finishing	RTO w/ conc. & recirculator	3,000	scfm	820,000	42,000	~2008		
Semiconductor	RCTO - CO Catalyst	7600	scfm	121,440	69,208	2014	Based on manufacturer cost information	5
Ethanol Plant	RTO	44,500	scfm	850,000	NA	2005	TO manufacturer	6
Ethanol Plant	Recuperative TO (50%)	57,200	scfm	1,000,000	NA	2005	TO manufacturer	
Specialty Med products	Recuperative TO (65%)	1,500	scfm	145,000	NA	~2006	Capital cost is equipment cost only	Nester
Tire cord coating	RTO	25,000	scfm	450,000	NA	~2006		
Sewage sludge incineration	Fluid Bed Incinerator	4	dry tons/hour	$75 million	NA	2010	Capital cost for 3 fluid bed incinerators, each with capacity 4 dry tons/hour. Units burn undigested sludge and are autogenous. Costs are total for permitting and construction.	7

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[3] See http://www.ecy.wa.gov/programs/air/psd/PSD_PDFS/Final_Boeing_777X_TSD_PSD_14-01_09092014.pdf.
[4] Presentation prepared by Catalytic Combustion, “Air Emissions Control in Spray Finishing Applications.
[5] See http://www.oregon.gov/deq/NWR/Documents/IntelType4permitAppl.pdf.
[6] See http://www3.epa.gov/chp/documents/voc_destruction_white_paper.pdf.
[7] See http://projectgroundwork.org/sustainability/op_enviro/incinerator.html.

2.2 Process Description

Gaseous waste streams may be composed of a complex mixture of organic compounds. This mixture is typically analyzed for carbon, hydrogen, oxygen, and other elements; and an empirical formula is developed which represents the mixture. Combustion of such a mixture of organic compounds containing carbon, hydrogen, and oxygen is described by the overall exothermic reaction:

$$C_xH_yO_z + \left[x + \frac{y}{4} - \frac{z}{2}\right] O_2 \rightarrow xCO_2 + \frac{y}{2} H_2O \tag{2.1}$$

In addition to carbon dioxide and water, exhaust gases from thermal oxidizers may also contain nitrogen oxides, acidic gases, trace metals (e.g., arsenic, beryllium, cadmium, chromium, nickel, and mercury), and other hazardous air pollutants (e.g., dioxins and furans) generated from combustion of compounds present in the waste or from the combustion of supplemental fuels. Nitrogen oxide formation can be reduced through process adjustments (e.g., using low-NO_x burners) or controlled using reducing agents such as ammonia and urea-based scrubbers. Particulates, including trace metals can be controlled through use of mechanical collectors, wet scrubbers, fabric filters, and electrostatic precipitators. Formation of dioxins and furans can be reduced by ensuring adequate combustion temperatures, providing the proper amount and distribution of combustion air, and optimizing the mixing process. Dioxins and furans may be controlled using a spray dryer, water sprays, or injection of carbon in combination with the PM control device. (EPA, 1997) (For information on how to design and estimate costs of particulate controls, please see Section 6 of the Cost Manual).

Waste streams containing halogenated and sulfur compounds will result in emissions of acid gases as the oxidizer converts these compounds to HCl, SO_2 and other acid gases. Plants may need to install an acid gas removal system, such as a wet scrubber, after the oxidizer. An oxidizer followed by a caustic scrubber can be a cost effective and environmentally beneficial approach to control waste streams containing these species. However, the additional cost of installing and operating an acid gas removal system must be considered as it could greatly affect the cost of the incineration system. Scrubbers are sometimes available as part of the main thermal oxidizer package and thus can be included in the equipment cost estimates. Also, if a scrubber is needed at the outlet to the oxidizer, compliance with SO_2 and halogen emission limits will incur additional performance testing and compliance costs for monitoring, possibly continuous monitoring. (For information on how to design and estimate costs of acid gas controls, please see Section 5 of the Cost Manual).

The combustion of waste gases may be accomplished in a thermal oxidizer or in a catalytic oxidizer. In the catalytic oxidizer, a catalyst is used to increase the rate of the combustion reaction, allowing the combustion to occur at lower temperatures. Because the catalytic process operates at a lower temperature than the thermal process, less auxiliary fuel may be required in the catalytic process to preheat the waste gas.

The heart of an incinerator system is a combustion chamber in which the waste stream is burned. Since the inlet waste gas stream temperature is generally much lower than that required for combustion, energy must be supplied to the incinerator to raise the waste gas temperature.

Often, the energy released by the combustion of the total organics (VOCs and others) in the waste gas stream is not sufficient to raise its own temperature to the desired levels. In these cases, auxiliary fuel (e.g., natural gas) must be added to raise the temperature.

Auxiliary fuel requirements may also be decreased, and energy efficiency improved, by providing heat exchange between selected inlet streams and the effluent stream. The effluent stream containing the products of combustion, along with any inert compounds that may have been present in or added to the inlet streams, can be used to preheat the incoming waste stream, auxiliary air, or both via a "primary", or recuperative, heat exchanger. The fractional energy recovery by the preheater or primary heat exchanger is defined as follows:

$$\begin{matrix}\textit{Fractional}\\ \textit{Energy}\\ \textit{Recovery}\end{matrix} = \frac{\textit{Energy actually recovered flue gas}}{\begin{matrix}\textit{Maximum energy recoverable if flue gas approaches}\\ \textit{lowest temperature available to heat exchanger}\end{matrix}} \tag{2.2}$$

The energy actually recovered, the numerator of Equation 2.2, is the increase in sensible heat of the gas, e.g., waste gas or waste gas plus dilution air, being heated. The maximum energy recoverable would be the decrease in sensible heat of the flue gas, if it were cooled to the temperature of the incoming waste gas. While this maximum energy recovery would be attained only with a very large heat exchanger, the concept of fractional energy recovery is useful in expressing the extent of the improvement in energy efficiency using a "primary" heat exchanger.

Energy efficiency can be further improved by placing another ("secondary") exchanger downstream of the primary exchanger to recover additional energy from the effluent stream (e.g., to generate low pressure process steam or hot water). Secondary energy recovery can be economically advantageous where there is a use for the steam or hot water. However, secondary energy recovery is generally not used unless there is a specific on-site use for it.

The majority of industrial gases that contain VOCs are dilute mixtures of combustible gases in air. In some applications, such as air oxidation processes, the waste gas stream is very deficient in oxygen. Depending on the oxygen content of the waste stream, auxiliary air may be required to combust the total organic content of the waste gas as well as any auxiliary fuel that has been used.

The concentration of combustible gas in the waste gas stream plays an integral role in the design and operation of an incinerator. From a cost standpoint, the amount of air in excess of the stoichiometric amounts should be minimized. For safety reasons, however, any mixture within the flammability limits, on either the fuel-rich or fuel-lean side of the stoichiometric mixture, presents an unacceptable fire hazard as a feed stream to the incinerator. The lower, or fuel-lean, explosive limit (LEL) of a given organic compound defines the minimum concentration of that compound in air that can produce more energy than is needed to raise its own temperature to the ignition point. Similarly, the upper, or fuel-rich, explosive limit (UEL) represents the highest concentration of the organic in air that is ignitable. In the latter case, the amount of available oxygen in the air limits the reaction. Both the LEL and the UEL are measured at standard temperature and pressure. The LEL and heating values for some commonly used organic compounds are shown in Table 2.13. Typically, the waste gas stream contains a mixture of hydrocarbons in air. For a typical hydrocarbon mixture, the heating value would be approximately 50 Btu/scf at the LEL (Vatavuk, 1990).

Since the majority of industrial waste gases that contain VOCs are dilute mixtures of combustible gases in air, their heating values are low and their oxygen content exceeds that required to combust both the waste organics (VOCs and others) and the auxiliary fuel. If a waste gas above 50 percent LEL (about 25 Btu/scf) is encountered, it must be diluted to satisfy fire insurance regulations. Generally, the streams are diluted to below 25 percent LEL, although concentrations from 25 percent to 50 percent are permitted provided the waste stream is continuously monitored by LEL monitors. Because air is the usual diluent gas, care must be taken with preheating the diluted stream so that it remains below about 1200°F. (See discussion below on preheating.) Tables showing LEL, UEL, and heats of combustion for selected organic compounds are provided in Appendix A.

Currently, some regenerative and catalytic oxidizers include “self-sustaining” features that operate without auxiliary fuel while processing waste gas streams with organic content greater than 3 to 10% of the LEL. These units use two sets of heat absorbing ceramic media beds with excellent thermal energy recovery characteristics. The first set of ceramic media heats the inlet gas, while the second set recovers thermal energy generated during combustion. The direction of gas flow is periodically reversed so that the outlet ceramic bed becomes the inlet bed and vice versa. This system enables high levels of thermal energy to be recovered. Some manufacturers report thermal energy recovery rates as high as 97 percent. This high rate of recovery significantly reduces fuel consumption and reportedly allows some units to operate without auxiliary fuel after initial startup.[8]

The goal of any incineration system is to control by thermal destruction the amount of VOCs released to the environment. Performance of a control device such as an oxidizer can be described by a destruction and removal efficiency (DRE) in percent (%), defined according to the following equation:

$$\mathrm{DRE} = \left[\frac{Inlet\ mass\ rate\ VOC - Outlet\ mass\ rate\ VOC}{Inlet\ mass\ rate\ VOC}\right] \times 100 \qquad (2.3)$$

It is important to note, however, that incomplete combustion of the inlet VOCs could result in the formation of other VOCs not originally present. For example, the incomplete oxidation of dichloroethane can yield vinyl chloride. Both of these compounds are VOCs. The definition given in Equation 2.3 would still be meaningful, however, as long as the newly formed VOC (e.g., vinyl chloride) is detected. This situation necessitates the complete chemical analysis of the inlet and outlet gas streams to confirm compliance with state and Federal regulations.

The outlet VOC concentration, usually in parts per million by volume (ppmv), is sometimes used to assess the performance of an oxidizer. The outlet VOC concentration is typically used to assess performance for sources subject to regulatory limits on the outlet VOC concentration, rather than a percent reduction limit.

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[8] See http://www.anguil.com/oxidizers/regenerative-thermal.aspx

Equipment life for an oxidizer is variable and depends on several factors, including the system design and materials used in its construction, composition of the waste gas stream, and the temperatures experienced by the oxidizer. In general, oxidizers that handle corrosive waste gases or higher levels of particulates will have a shorter operational life. Systems that undergo frequent fluctuations in temperature or more frequent startup-shutdown cycles will have a shorter operational life than systems where a steady temperature is maintained.

There are several different oxidizer designs available. These designs can be broadly classified as thermal systems and catalytic systems. Thermal systems may be direct flame incinerators with no energy recovery, flame incinerators with a recuperative heat exchanger, or regenerative systems that operate in a cyclic mode to achieve high energy recovery. Catalytic systems include fixed-bed (packed-bed or monolith) systems and fluid-bed systems, both of which provide for energy recovery.

Solid waste incinerators operate in a similar manner to oxidizers but receive solid and liquid waste, instead of waste gas streams. There are six different designs of solid waste incinerators, all of which use thermal energy to combust waste materials and destroy VOC and HAP. These designs are grate incinerators (fixed or moving), rotary kilns, multiple hearth incinerators, fluid bed incinerators, controlled-air, and excess air incinerators.

Oxidizers and incinerators vary in size. They may be small, prefabricated, modular designs or large units that must be constructed onsite. Some of the larger units, particularly those used to combust municipal waste, include heat recovery systems that can be used for steam and/or electricity production. The following sections discuss design aspects of these systems.

2.2.1 Solid Waste Incinerators

Solid waste incinerators typically emit hazardous air pollutants, including dioxin, furan, mercury, lead, cadmium, and other heavy metals. For this reason, solid waste incinerators are generally fitted with air pollution controls, such as afterburners to reduce carbon monoxide emissions, scrubbers to remove particulates and acid gases, filters (e.g., electrostatic precipitators, cyclones, or baghouses) to remove particulates, and dry sorbent injection for acid gas control. The types of pollution controls used depend on the composition of the wastes burned and on the design of the solid waste incinerator. For example, fluidized bed incinerators can control NOx by using lower temperatures at which NOx do not form, while sulfur dioxide emissions are controlled by mixing sulfur absorbing materials (e.g., limestone or dolomite) into the fluid bed medium.

There are several different designs of solid waste incinerators. A description of each type is included below.

Fixed Grate Incinerators

Fixed grate incinerators consist of a fixed metal grate over an ash pit. Waste is loaded either from the side or top, and ash removed from the bottom. Many small incinerators used to combust municipal waste have the fixed grate design.

Moving Grate Incinerators

In a moving grate incinerator, waste is loaded onto a moving metal grate that moves through the combustion chamber and deposits ash into a pit at the other end. Air is supplied through the grate from below and through air nozzles located above the grate to provide mixing of gases and surplus oxygen to ensure complete combustion. Operating temperatures are typically maintained at 850 ^{o}C (1,560 ^{o}F) or higher to ensure complete destruction of toxic organic materials. Moving grate incinerators are used to incinerate municipal waste.

Rotary Kilns

Rotary kilns are used to incinerate hazardous waste and hospital, medical, and infectious waste. A rotary kiln consists of an inclined, cylindrical vessel that is rotated slowly about its longitudinal axis. Waste material is fed into the upper end of the vessel and gradually moves to the lower end and is gradually mixed and organic material combusted. Depending on the design, gasses from the combustion process move either in the same direction as the waste or in the opposite direction. The kiln is usually equipped with an auxiliary burner that is used to start combustion and maintain the combustion temperature. Rotary kilns are generally also equipped with a secondary chamber, where combustion of volatile organic compounds generated in the rotary kiln are combusted in excess air. Because the waste is rotated in the primary combustion chamber, rotary kilns typically have higher particulate emissions than other incinerator designs and are usually equipped with scrubbers, fabric filtration systems, or electrostatic precipitators for particulates.

Multiple Hearth Incinerators

Multiple hearth incinerators are used to incinerate sewage sludge from wastewater treatment plants. They consist of a cylindrical vessel constructed of steel and lined with refractory. Inside are a series of horizontally stacked refractory-lined hearths. The typical multiple hearth incinerator has 5 to 12 hearths, with 9 hearths required for complete combustion. The upper hearth tiers are the drying zone, the middle tiers the combustion zone, and the bottom tiers the ash cooling zone. A hollow cast iron shaft passes through the center of the vessel through which air is pumped from the bottom of the vessels to the hearths above before exiting the vessels at the top. Some or all exiting air is recirculated to the bottom of the vessel, while additional ambient air is sometimes pumped into the middle hearth tiers to ensure complete combustion. Rotating arms equipped with teeth (called rabble arms) are located above the hearths. Dewatered sludge is fed into the top of the vessels onto the outer edge of the first tier of hearths. The rabble arms on the upper tier gradually move the sludge from the outside of the vessel to the inside, where it drops through holes onto the hearth below. On the second tier hearth, the rabble arms move the waste from the inside to the outside of the vessel, where the sludge drops through to the next level. The process is repeated through the various hearth tiers until ash is discharged through the bottom of the vessel. Auxiliary heat is provided by burners located on the sides of hearths. Process gases are exhausted through the top of the unit at temperatures between 427 and 760^{o}C (800 and 1,400^{o}F). Temperatures in the combustion zone maintained at about 482^{o}C (900^{o}F) and 75 to 100 percent excess air is used to ensure complete combustion of the sludge. The dewatered sludge consists of between 15 to 35 percent solids. Detention times for solids varies from less than 1 hour to 3 hours, while gas detention times are 1 to 3 seconds. If the sludge is dewatered to about 30 percent, the combustion process can be self-sustaining provided the heat value of the solids is sufficient. However, supplemental fuel is required for initial start-up and during periods when the heating value of the solids is low (EPA 1995 and 2003). In recent years, most multiple hearth incinerators are being replaced with more efficient fluidized bed incinerators.

Fluidized Bed Units

Fluidized bed units are generally used for sewage sludge incineration. They consist of a vertical steel vessel, lined with refractory material and equipped with air nozzles located on the floor of the vessel. A mixture of fuel, waste, sand, and other materials (e.g., limestone to control sulfur dioxide) are fed into the lower part of the vessel, while the jets of air agitate and mix the particles into a fluid-like suspension. The sand is typically between 0.8 to 1.0 meters (2.5 to 3 feet) thick. The result is a fluidized bed of mixed gases and solids that promotes heat transfer and chemical reactions. The injected air may be at ambient temperatures or preheated by passing through a heat exchanger where heat from exhausted gas is used to raise the temperature of the ambient air (EPA 1995).

The fluidized bed is maintained at temperatures between 760 to 871^{o}C (1,400 and 1,600 ^{o}F) (EPA 2003), which results in the evaporation of water and combustion of organic materials. In the area above the fluidized bed, any remaining carbon and combustible gases are burned. The gases are vented at temperatures between 760 and 870 ^{o}C (1,500 and 1,600^{o}F). Fluidized bed incinerators use less fuel and can operate with lower excess air than other incinerators. Fluidized bed incinerators typically operate with 20 to 50 percent excess air (EPA 1995 Retention times for solids vary from 1 to 5 minutes, while gas retention times are between 6 and 8 seconds (EPA 2003).

Due to the lower operating temperatures, NO_x emissions are generally lower than for other incinerators; however, carbon monoxide (CO) and other emissions (e.g., HAP) may be higher due to less complete combustion resulting from the lower combustion temperatures. For this reason, fluidized bed incinerators are sometimes fitted with afterburners to control CO emissions. Because of the lower emissions, most new sewage sludge incinerators are fluidized bed incinerators (EPA 1993). Sewage sludge incinerators are typically fitted with wet flue gas desulfurization (FGD) scrubbers to reduce mercury emissions in compliance with new source performance standard (NSPS), emissions guideline (EG), and federal implementation plan (FIP) that became final in 2016.[9]

Controlled-Air Incinerators

Controlled-air incinerators (sometimes called starved-air incinerators) are used to incinerate municipal solid waste and medical waste (e.g., hospital, medical, and infectious wastes). These units consist of two chambers. The primary combustion chamber is operated at temperatures of 760 to 980^{o}C (1,400 to 1,800^{o}F) with air supplied from beneath the waste bed. The air is supplied at less than the stoichiometric amount of oxygen required for combustion. The secondary combustion chamber is operated with excess air at temperatures of 980 to 1,095^{o}C (1,800 to 2,000^{o}F) (EPA 1993). Waste is fed into the primary combustion chamber, where it is

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[9] These final rules were set for mercury and eight other pollutants under section 129 of the Clean Air Act. These rules are contained in a Federal Register notice available at https://www.gpo.gov/fdsys/pkg/FR-2016-04-29/pdf/2016-09292.pdf.

dried and most residual carbon is burned. Volatile organics generated in the primary chamber enter the second chamber, where the higher temperatures and excess air ensure complete combustion. The secondary chamber is sometimes equipped with auxiliary burners located at the entrance to the chamber to help maintain the combustion temperature when wastes with low heating values are combusted. Controlled-air incinerators are generally not equipped with emissions controls because the flue gas tends to have low levels of particulates due to the relatively low air flow rates and the secondary chamber reduces CO emissions.

Excess-Air Incinerators

Excess air incinerators are small, typically batch units that are used to incinerate municipal solid waste and hospital, medical, and infectious wastes. They are generally operated at excess air levels up to 300 percent. Although designs vary, they generally consist of two or more chambers and baffles. Waste is fed into the primary combustion chamber, where the waste is dried and combusted by the primary chamber burner. Combustion gases, volatile materials, and water vapor exhaust to the secondary chamber. The secondary chamber is equipped with additional burners and an air injection port. The secondary chamber is typically maintained at a temperature between 870 and 980°C (1,600 and 1,800°F) (EPA 1993).

Mass Burn-Waterwall Units

Mass burn-waterwall combustion units are used to incinerate municipal solid waste (MSW) and are the preferred design for large municipal waste combustors (MWC). They range in size from 46 to 900 Mg/day (50 to 1,000 tons per day). The units are similar to the moving grate incinerator discussed above. They consist of primary combustion chamber with grate and a secondary chamber. The upper walls of the primary combustion chamber are equipped with metal tubes that contain circulating pressurized water used to recover heat from the combustion chamber. Heat is also recovered in secondary sections of the combustor. The recovered heat is used to produce steam and/or electricity. Waste is fed into a feed hopper by overhead cranes and then transferred to the combustion chamber either by gravity feeders or hydraulic rams. Reciprocating grates or roller grates are then used to move the waste through the various stages of the combustion chamber. In the first stage, the moisture content of the waste is reduced prior to ignition. The second stage, referred to as the burning grate, is where the majority of active burning takes place. The last stage, referred to as the burnout or finishing grate, is where remaining combustibles in the waste are burned. Ash is discharged from the finishing grate into a water-filled ash quench pit or ram discharger. Air is supplied both from beneath the grate and through rows of high-pressure nozzles located in the side walls of the combustion chamber. The combustion chambers are typically operated with 80 to 100 percent excess air (EPA 1996).

Mass Burn Rotary Waterwall Units

Mass burn-refractory kilns are used to incinerate MSW. They range in size between 180 to 2,400 Mg/day (200 to 2,700 tons per day) and use a rotary combustion chamber followed by a waterwall furnace. The rotary combustion chamber operates in the same manner as the rotary kiln described above. However, in mass burn rotary waterwall units, the rotary combustion chamber is equipped with water-filled tubes for heat recovery. The waste is fed into the inclined rotary combustion chamber using hydraulic rams. Air is injected both from beneath and above the waste bed, with most of the combustion air supplied in the first half of the rotary combustion chamber.

The rest of the combustion air is supplied to the afterburner grate and above the rotary combustor outlet in the furnace. They are typically operated at about 50 percent excess air. As with the mass burn waterwall units, recovered heat is used for steam and/or electricity production (EPA 1996).

2.2.2 Thermal Oxidizers

The heart of the thermal oxidizer is a nozzle-stabilized flame maintained by a combination of auxiliary fuel, waste gas compounds, and supplemental air added when necessary (see Figure 2.1). Upon passing through the flame, the waste gas is heated from its inlet temperature (e.g., 100°F) to its ignition temperature. The ignition temperature varies for different compounds and is usually determined empirically. It is the temperature at which the combustion reaction rate (and consequently the energy production rate) exceeds the rate of heat losses, thereby raising the temperature of the gases to some higher value. Thus, any organic/air mixture will ignite if its temperature is raised to a sufficiently high level.

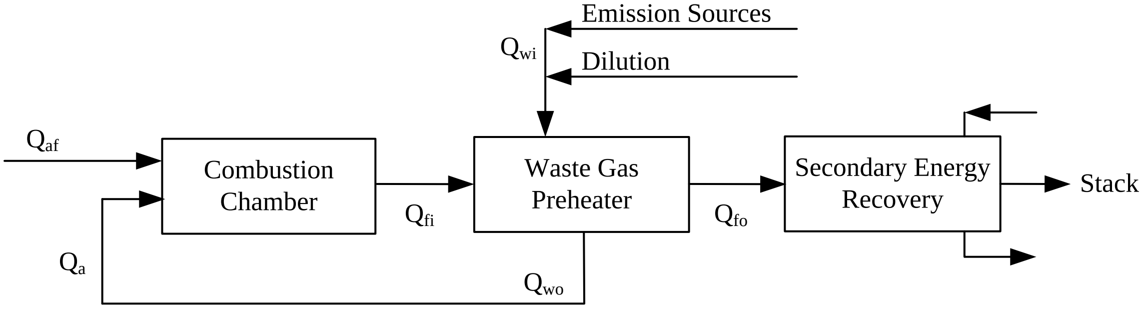

Figure 2.1: Thermal Oxidizer - General Case

The organic-containing mixture ignites at some temperature between the preheat temperature and the reaction temperature. Ignition occurs at some point during the heating of a waste stream as it passes through the nozzle-stabilized flame regardless of its concentration. The mixture continues to react as it flows through the combustion chamber.

The required level of VOC control of the waste gas that must be achieved within the time that it spends in the thermal combustion chamber dictates the reactor temperature. The shorter the residence time, the higher the reactor temperature must be. For example, a 95% destruction efficiency for toluene can be achieved at a temperature of 1351 °F using a thermal oxidizer with a half second residence time. If a thermal oxidizer with a one second residence time were used, a 95% destruction efficiency could be achieved with a temperature of only 1317 °F. To achieve 99% destruction of toluene using a thermal oxidizer with a half a second residence time, the operating temperature would have to be increased to 1372 °F.[10] The nominal residence time of the reacting waste gas in the combustion chamber is defined as the combustion chamber volume divided by the volumetric flow rate of the gas. Most thermal units are designed to provide no more than 1 second of residence time to the waste gas with typical temperatures of 1,200 to

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[10] See http://www.anguil.com/resources/overview-of-emission-control-technologies/.

2,000°F. Once the unit is designed and built, the residence time is not easily changed, so that the required reaction temperature becomes a function of the particular gaseous species and the desired level of control. Table 2.2 illustrates the variability in (theoretical) reactor temperatures that is required to destroy 99.99 percent of the inlet mass of various noxious compounds with excess air for a 1- second reactor residence time (National Academy Press, 1983).

These temperatures cannot be calculated *a priori*, although incinerator vendors can provide guidelines based on their extensive experience. In practice, most streams are mixtures of compounds, thereby further complicating the prediction of this temperature. Other studies (Mascone 1980a, 1980b, and 1980c), which are based on actual field test data, show that commercial incinerators should generally be run at 1600°F with a nominal residence time of 0.75 seconds to ensure 98% destruction of non-halogenated organics. In some permits, the reactor temperature and residence time of the unit are specified rather than attempting to measure actual levels of VOC control. The selected temperature must be maintained for the full, selected residence time for combustion to be complete.

These three studies also conclude that mixing is a critical factor in determining the destruction efficiency. Even though it cannot be measured, mixing is a factor of equal or even greater importance than other parameters, such as temperature. The most feasible and efficient way to improve the mixing in an incinerator is to adjust it after start-up. The 98% control level discussed in the previous paragraph presumes such an adjustment.

Ultimately, once the unit is built, it is the responsibility of the user to operate and maintain the incinerator to insure compliance with applicable regulations.

Direct Flame Incinerators

Many configurations of thermal oxidizers exist with the same goal—to raise the VOC-containing stream to the desired reaction temperature and hold it there for the given reaction time to achieve the required destruction efficiency. The simplest example of such a system is the direct flame incinerator. With reference to Figure 2.1, the direct flame incinerator is comprised only of the combustion chamber. The waste gas preheater and the secondary energy recovery heat exchanger are energy recovery devices and are not included as part of the direct flame incinerator.

Table 2.2: Theoretical Reactor Temperatures Required for 99.99 Percent Destruction by Thermal Incineration for a 1-Second Residence Time (National Academy Press, 1983)

Compound	Temperature, °F
acrylonitrile	1,344
allyl chloride	1,276
benzene	1,350
chlorobenzene	1,407
1,2-dichloroethane	1,368
methyl chloride	1,596
toluene	1,341
vinyl chloride	1,369

Recuperative Thermal Oxidizers

Recuperative oxidizers have improved energy efficiency as a result of placing heat exchangers in the hot outlet gas streams. Regarding Figure 2.1, the recuperative oxidizer is comprised of the combustion chamber, the waste gas preheater, and, if appropriate, the secondary, energy recovery heat exchanger.

Primary Energy Recovery (Preheating Inlet Streams) - Considerable fuel savings can be realized by using the exit (product) gas to preheat the incoming feed stream, combustion air, or both via a heat exchanger, as shown in Figure 2.1 in the so-called “recuperative” incinerator. These heat exchangers can recover up to 70% of the energy (enthalpy) in the product gas.

The two types of heat exchangers most commonly used are plate-and-frame (or simply plate) and shell-and-tube. Plate exchangers offer high efficiency energy recovery at lower cost than shell-and-tube designs. Also, because of their modular configuration, plate units can be built to achieve a variety of efficiencies. But when gas temperatures exceed 1000°F, shell-and-tube exchangers usually have lower purchase costs than plate designs. Moreover, shell-and-tube exchangers offer better long-term structural reliability than plate units (Schmidt, 1989b). In any case, because most incinerators installed are packaged units, the design (and cost) of the recuperative heat exchangers has already been incorporated.

Most heat exchangers are not designed to withstand high temperatures, so that most of the energy needed to reach ignition is supplied by the combustion of fuel in the combustion chamber and only moderate preheat temperatures are sought in practice (<1200°F).

Secondary Energy Recovery (Additional Waste Energy Recovery) - It should be noted, however, that at least some of the energy added by auxiliary fuel in the traditional thermal units (but not recovered in preheating the feed stream) can still be recovered. Additional heat exchangers can be added to provide process heat in the form of low pressure steam or hot water for on-site application. Obviously, an in-plant use for such low-level energy is needed to realize these savings.

The need for this higher level of energy recovery will be dependent upon the plant site. The additional heat exchanger is often provided by the incineration unit vendor. The cost of this additional heat exchanger may be estimated via standard heat exchanger correlations and should be added to the costs estimated using the cost correlations in this section.

Regenerative Thermal Oxidizers

The traditional approach to energy recovery in the recuperative units (shown schematically in Figure 2.1) still requires a significant amount of auxiliary fuel to be burned in the combustion chamber when the waste gas heating values are too low to sustain the desired reaction temperature at the moderate preheat temperature employed. Additional savings can, under these conditions, be realized in units with more complete transfer of exit-stream energy. This is the concept behind the so-called excess-enthalpy or regenerable burner systems (i.e., regenerative thermal oxidizers (RTOs)). These systems use direct contact heat exchangers constructed of a ceramic material that can tolerate the high temperatures needed to achieve ignition of the waste stream. It has been reported that RTOs are the most widely used emission abatement technology and that roughly 80 percent of all recent thermal oxidizer applications are RTOs.[11,12]

The operation of the regenerative system is illustrated in Figure 2.2. As shown in mode A of Figure 2.2, the inlet gas first passes through the first hot ceramic bed thereby heating the stream (and cooling the bed) to its ignition temperature. If the desired temperature is not attainable, a small amount of auxiliary fuel is added in the combustion chamber. The hot gases then react (releasing energy) in the combustion chamber and while passing through the second ceramic bed, thereby heating it to the combustion chamber outlet temperature. When the temperature of the outlet ceramic bed reaches a set temperature, the process flows are reversed so that the inlet gas is now fed into the hot second ceramic bed and exits through the first ceramic bed, as shown in mode B in Figure 2.2. This cyclic process affords very high energy recovery.[13] While regenerative oxidizers generally consist of two ceramic beds, three bed systems are available.[14]

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[11] Anguil. Regenerative Thermal Oxidizer (RTO). Available at “http://www.anguil.com/oxidizers/regenerative-thermal-oxidizer-rto/.

[12] Controlled Environment Equipment Corp. Regenerative Oxidizer Systems in Comprehensive Industrial Solutions. Available at: “http://www.cee-corp.com/VentilationSystems/RegenerativeOxidizer”.

[13] Some manufacturers claim thermal energy recoveries of 97 percent (See http://www.anguil.com/oxidizers/regenerative-thermal.aspx).

[14] See http://www.glenro.com/thermox-regenex.html.

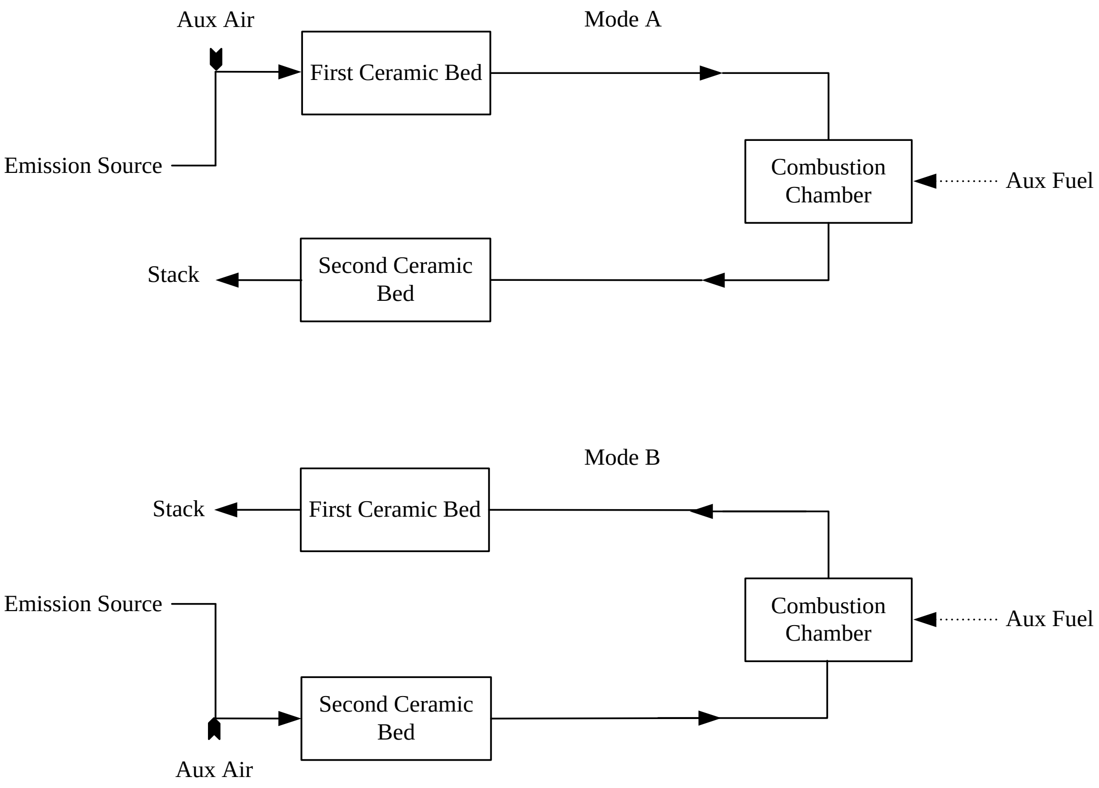

Figure 2.2: Regenerable-Type/Thermal oxidizer

The typical regenerative oxidizer uses valves (typically butterfly or poppet valves)[15] to alternate the airflow direction through the media beds and thereby maximize energy recovery. Since the late 1990s, some RTOs have been designed with a single rotary valve that reduces the number of moving parts and may require less maintenance. The high-energy recovery within these oxidizers reduces the auxiliary fuel requirement and saves operating cost. At organic concentrations as low as 3 to 4 percent of the LEL, some of these oxidizers achieve high destruction efficiency and self-sustaining operation with no auxiliary fuel usage.[16,17]

The higher capital costs associated with these high-performance heat exchangers and combustion chambers may be offset by the increased auxiliary fuel savings to make such a system economical. The costs of these regenerative units will be given separately in the cost correlations presented in Section 2.4. In the past, all regenerative oxidizers were field-erected, but currently packaged units are available in sizes up to 75,000 scfm.[18] The costs given in Section 2.4 for regenerative units are for field-erected units.

One potential issue with regenerative oxidizers is the presence of a slug of non-destroyed process air (known as the “puff”) that exists during valve switching. It occurs when a chamber on “inlet mode” is switched to “outlet mode”, and the process air that just entered the cold face

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[15] See http://airclear.net/regenerative-rto-thermal-oxidizer/.
[16] See http://www.anguil.com/oxidizers/regenerative-thermal.aspx.
[17] See http://airclear.net/regenerative-rto-thermal-oxidizer/.
[18] See http://www.anguil.com/oxidizers/regenerative-thermal-oxidizer-rto/.

plenum and the lower section of the heat recovery media reverses directions and is sent up the stack. Such puffs limit the maximum DRE of the unit. Several techniques have been implemented to minimize puffs. One technique is to minimize the volume and frequency of puffs with low volume cold face plenum and highly efficient heat recovery media. Another technique is to use three or more chambers. In these multi-chambered units the untreated gas is directed to the idle third chamber and then bled through the combustion chamber instead of going directly to the stack. Another way to control the "puff" is by purging chambers with clean air between inlet and outlet modes with the purge gas sent to a third chamber.[19]

In recent years, a new system has been developed that uses a rotating distributor in place of the valve and damper system. In this system, the packed ceramic honeycomb elements of the heat exchanger are divided into sections and then rotated so that different sections of the heat exchanger serve the cooling phase while another section serves the heating phase. The waste gas flows through the RTO's heat exchanger from the bottom to the top, thereby heating the waste gas to the combustion temperature. The hot gases of combustion then flow down through the other side of the rotating heat exchanger where the waste heat is absorbed. This design eliminates the need for complicated valve and damper switching systems and is said to eliminate fluctuations in flow caused by valve switching.[20]

Ceramic Media

Several different types of heat transfer media are used for RTOs; including random packing, monolithic (honeycomb) structured block and corrugated structure packing.

Random packing uses a wide variety of packing materials, including gravel, ceramic balls and shapes of all kinds which are randomly dumped into the RTO to form a media bed. Random arrangement prevents nesting that would constrict flow and cause dead areas that collect particulate. Ceramic "saddles" have proved to be the optimal shape for RTO random packing because it minimizes pressure drop (for lower electricity consumption by the induction fan) and maximizes surface area (for higher heat transfer efficiency). Recently, RTO media suppliers have refined the design of ceramic saddles to provide a high open area and aerodynamic design that limits nesting and reduces pressure drop by 20 percent compared to standard saddle media.

Monolith structure is a form of structured packing that is placed in a formal arrangement, rather than randomly dumped. Cells extend through the block in a straight channel perpendicular to the cold face. The advantage of this design is that it theoretically provides a straight, aerodynamic channel for the air stream. The disadvantage is that if particulate plugs a channel at the cold face, where the inflow enters the block, then this entire channel becomes a dead zone. Vendors recommend adding particulate removal devices upstream of the RTO if the particulate concentration is greater than 0.005 gr/dscf to 0.002 gr/dscf (Raemhild, 2001, and Nester, 2006).[21]

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[19] http://www.nestecinc.com/pdf/PuffReducationTech.pdf

[20] http://www.eisenmann.com/en/products-and-services/environmental-technology/exhaust-air-purification.html.

[21] gr/dscf = grains per dry standard cubic foot

Corrugated structured packing is constructed of corrugated sheets of ceramic. The angle of inclination of the corrugations of adjacent sheets is reversed, ensuring excellent distribution of air flow throughout the media bed. [22]

Manufacturers of ceramic media are constantly developing or improving proprietary products to optimize pressure drop, thermal efficiency, corrosion, plugging/fouling, and/or surface area. Additionally, manufactures state that some of these new material compositions and shapes in media can be substituted without modifying your existing unit or creating additional installation hassles. [23]

Developments to mitigate issues relating to varying VOC loads in gas streams are also available and include air dilution and changing the switching cycle time. In some cases, a hot bypass valve is used to direct the clean (hot) oxidizer exhaust around the heat recovery bed, directly into the stack. Bypassing temporarily reduces the thermal efficiency, thereby allowing higher solvent concentrations to be processed. [24]

Another development in RTO operation yields an increase in the overall Btu value of the process gas stream by injecting a controlled fuel or a premixed air and fuel mixture (controlled to less than 25% of the LEL) directly into the process inlet of the RTO. This additional energy reduces the energy demand on the burners, causing the burners to turn down and run at their low-fire or pilot position, if included. Under this condition, significantly less combustion air and natural gas is introduced directly into the combustion chamber than when the burners are operating. Since the mass imbalance across the heat recovery chambers is virtually eliminated, the measured overall effective thermal efficiency of the RTO unit is markedly improved, and the fuel consumption reduced and the overall performance improved (Bunimovitch, 2010).[25]

2.2.3 Catalytic Oxidizers

Catalytic oxidizers operate very similar to thermal oxidizers, with the primary difference that the gas, after passing through the flame area, passes through a catalyst bed. The catalyst facilitates the overall combustion reaction given in Equation 2.1 by increasing the reaction rate, enabling conversion at lower reaction temperatures than in thermal oxidizer units. Nevertheless, the waste stream must be preheated to a temperature sufficiently high (usually from 300 to 900^{o}F) to initiate the oxidation reactions. The heated gas then passes through the catalyst bed. Oxygen and VOC migrate to the catalyst surface by gas diffusion and are adsorbed onto the active sites on the surface of the catalyst where oxidation then occurs. The oxidation reaction products are then desorbed from the active sites by the gas and transferred by diffusion back into the gas stream. The preheated gas stream is then passed over the catalyst bed. The chemical reaction (combustion) between the oxygen in the gas stream and the gaseous pollutants takes place at the catalyst surface.

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[22] http://www.reliableplant.com/Read/9117/thermal-oxidizers-particulate-buildup

[23] http://www.norpro.saint-gobain.com/uploadedFiles/SGnorpro/Documents/RTO-HeatTransfer.pdf

[24] http://www.epa.gov/cmop/docs/cmm_conference_sept09/07stone.pdf
http://airclear.net/regenerative-rto-thermal-oxidizer/

[25] http://www.nestecinc.com/november-2013-newsletter.html
http://www.anguil.com/case-studies/service-and-maintenance/rto-media-retrofit.aspx

Catalytic incineration can, in principle, be used to destroy essentially any oxidizable compound in an air stream. However, there are practical limits to the types of compounds that can be oxidized due to the poisoning effect some species have on the catalyst. These limits are described below. In addition, most configurations require a low heating value of the inlet gas and low particulate content. The particulate content threshold varies between catalysts and depends on the pore size and volume of the catalyst.

The volumetric gas flow rate and the concentration of combustibles in the gas flowing to the catalytic oxidizer should be constant for optimal operation. Large fluctuations in the flow rate will cause the conversion of the VOCs to fluctuate also. Changes in the concentration or type of organics in the gas stream can also affect the overall conversion of the VOC contaminants. These changes in flow rate, organics concentration, and chemical composition are generally the result of upsets in the manufacturing process generating the waste stream. In situations where the flow rate, concentrations, and chemical composition frequently fluctuate, thermal oxidizers (discussed earlier in this chapter) or carbon adsorption (discussed in Section 3.1 of this Manual) should be evaluated as alternative control technologies.

As was the case for thermal units, it is impossible to predict *a priori* the temperature and residence time (e.g., inverse space velocity) needed to obtain a given level of conversion of a VOC mixture in a catalytic oxidation system. For example, Table 2.3 (Pope et al., 1976) shows the temperature needed for 80% conversion of several VOCs over two oxidation catalysts in a specific reactor design. This table shows that the temperature required for this level of conversion of different VOCs on a given catalyst and of the same VOC on different catalysts can vary significantly.

Table 2.3: Catalyst Temperatures Required for Oxidizing 80% of Inlet VOC to CO_2, ^oF for Two Catalysts$^{(a)}$

Compound	Temperature, ^oF	
	CO_3O_4	**Pt-Honeycomb**
acrolein	382	294
n-butanol	413	440
n-propylamine	460	489
Toluene	476	450$^{(b)}$
n-butyric acid	517	451
1, 1, 1-trichloroethane	661	>661
dimethylsulfide	—	512

(a) Except as specified in (b), data is from Pope, 1976.
(b) ICAC, 2016.

The control efficiency and performance of a catalytic oxidizer system is in part dictated by the permit requirements. More stringent performance requirements and catalyst life requirements require the catalyst suppliers to design for longer life and performance. Suppliers often provide catalysts capable of higher performance that are warranted to the lower level required by the permit. Higher levels of control and performance requirements have a significant impact on the cost. For example, changing from 95% to 99% DRE can cause the catalyst volume to increase by over 50%.

Catalytic oxidizers have been used to control VOC emissions from a variety of commercial applications. In many applications, catalysts have proven to be very durable with catalyst life ranging from 4 years up to as much as 16 years. Many catalysts have operated for over 10 years, with little or no loss in control efficiency. Table 2.4 shows the ranges for catalyst life observed in a range of different applications.

Table 2.4 Typical Ranges for Catalyst Life

Industry	Typical Compounds Treated	Number of Years Before Catalyst Replacement
Can Coating	MIBK, Mineral Spirits, Isophorone, DIBK, Butyl Cellosolve	7 to 14
Metal Coatings	MEK, MIBK, Toluene, i-Butanol	7 to 10
Automotive Paint Bake	MEK, Toluene, Xylene, Isopropyl alcohol	5 to 14
Glove Manufacturing	Formaldehyde, Phenolics	5
Phthalic Anhydride	PA, MA, S	16
Synthetic Fabrics	Scotchguard, Thermosol Dye	5

In many cases, periodic cleaning of the catalyst can restore the catalyst to original or near-original activity levels.

Types of Catalytic Oxidizers

As is the case with thermal oxidizers, there are two types of catalytic oxidizers: recuperative catalytic oxidizers and regenerative catalytic oxidizers.

Recuperative catalytic oxidizer: In a recuperative catalytic oxidizer, the waste stream is preheated either directly in a preheater combustion chamber or indirectly by heat exchange with the incinerator’s effluent or other process heat or both (Figure 2.3).

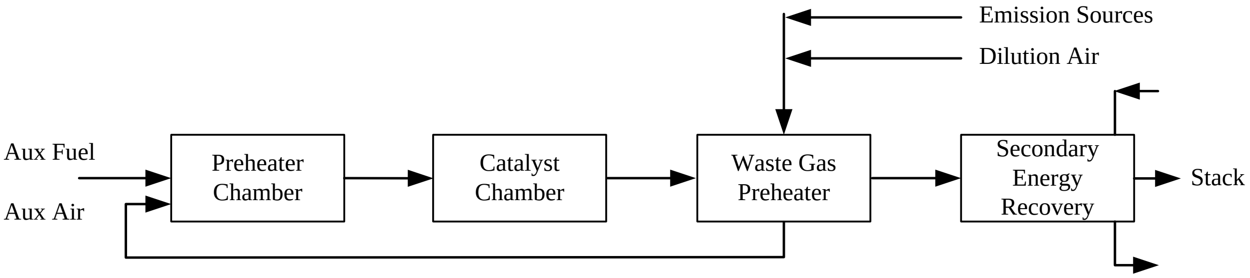

Figure 2.3: Recuperative Catalytic Oxidizer

Regenerative catalytic oxidizer (Bunimovitch, 2010): Regenerative catalytic oxidizers have been available since the mid-1990s. The design is similar to that of a regenerative thermal oxidizer, except that a layer of catalyst is located above the ceramic media. The catalyst is typically the same general type of material as the ceramic media (i.e., random packing, structured media). In many situations, RTOs can be retrofit to RCOs by either adding a layer of catalyst above the ceramic media or by replacing a portion of the ceramic media with catalyst. Energy savings have resulted in reported payback times of retrofits as short as a year or less. RCOs are most suited to controlling low VOC streams because the fuel use can be even lower than for RTOs.[26]

Catalyst Considerations

In the past, the use of catalytic oxidation for control of gaseous pollutants has generally been restricted to organic compounds containing only carbon, hydrogen, and oxygen. Gases containing compounds with chlorine, sulfur, and other atoms that may deactivate the supported noble metal catalysts often used for VOC control were not suitably controlled by catalytic oxidation systems. Catalysts now exist, however, that are tolerant of such compounds. Most of these catalysts are single or mixed metal oxides, often supported on a substrate such as alumina. Most of the development of poison-tolerant catalysts has focused on the oxidation of chlorine-containing VOCs. These compounds are widely used as solvents and degreasers and are often the subject of concern in VOC control in various regulations promulgated by EPA. Catalysts such as chromia/ alumina (Weldon, 1986 and Manning, 1984) cobalt oxide (Pope, 1976), and copper oxide/manganese oxide (Musick, 1974) have been used for oxidation of gases containing chlorinated compounds. Platinum-based catalysts are active for oxidation of sulfur containing VOCs, although they are rapidly deactivated by the presence of chlorine. Compounds containing lead, arsenic, calcium, zinc, sulfur, and phosphorous are generally considered poisons for most oxidation catalysts. However, their concentration may be sufficiently low so that the rate of deactivation and therefore, the catalyst replacement costs, could be low enough to consider catalytic oxidation. Additionally, proprietary precious metal catalysts with high geometric and specific surface area result in high tolerances to catalyst poisons.[27]

Particulate matter, including dissolved minerals in aerosols, can rapidly blind the pores of catalysts and deactivate them over time. Because essentially all the active surface of the catalyst is contained in relatively small pores, the particulate matter need not be large to blind the catalyst.

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[26] http://www.anguil.com/oxidizers/regenerative-catalytic/

[27] http://www.engineerlive.com/content/21226

No general guidelines exist as to particulate concentration and particulate size that can be tolerated by catalysts because the pore size and volume of catalysts vary greatly.

Methods of Contacting Gas Stream and Catalyst

The method by which the VOC-containing stream contacts the catalyst serves to distinguish catalytic incineration systems. Both fixed-bed and fluid-bed systems are available routes for catalyst use in oxidation and incineration. Fixed-bed catalytic oxidizers may use a monolith catalyst or a packed-bed catalyst. Each of these is discussed below.

Fixed Bed Monolith Catalyst Oxidizers - The most widespread method of contacting the VOC- containing stream with the catalyst is the catalyst monolith. In this scheme the catalyst is a porous solid block containing parallel, non-intersecting channels aligned in the direction of the gas flow.

Monoliths offer the advantages of minimal attrition due to thermal expansion/ contraction during startup/shutdown and low overall pressure drop.

Packed-Bed Catalytic Oxidizers - A second contacting scheme is a simple packed-bed in which catalyst particles are supported either in a tube or in shallow trays through which the gases pass. The first scheme is not in widespread use due to its inherently high pressure drop, compared to a monolith, and the breaking of catalyst particles due to thermal expansion when the confined catalyst bed is heated/cooled during startup/shutdown. However, the tray type arrangement, where the catalyst is pelletized is used by several industries (e.g., heat-set web-offset printing). Pelletized catalyst is advantageous where large amounts of such contaminants as phosphorous or silicon compounds are present. (Yarrington, 1989b)

Fluid Bed Catalytic Oxidizers - A third contacting pattern between the gas and catalyst is a fluid-bed. Fluid-beds have the advantage of very high mass transfer rates, although the overall pressure drop is somewhat higher than for a monolith. An additional advantage of fluid-beds is a high bed-side heat transfer as compared to a normal gas heat transfer coefficient. This higher heat transfer rate to heat transfer tubes immersed in the bed allows higher heat release rates per unit volume of gas processed and therefore may allow waste gas with higher heating values to be processed without exceeding maximum permissible temperatures in the catalyst bed. In these reactors, the gas phase temperature rise from gas inlet to gas outlet is low, depending on the extent of heat transfer through imbedded heat transfer surfaces. The catalyst temperatures depend on the rate of reaction occurring at the catalyst surface and the rate of heat exchange between the catalyst and imbedded heat transfer surfaces.

Generally, fluid-bed systems are more tolerant of particulates in the gas stream than either fixed-bed or monolithic catalysts. This is due to the constant abrasion of the fluidized catalyst pellets, which helps remove these particulates from the exterior of the catalysts in a continuous manner. A disadvantage of a fluid-bed is the gradual loss of catalyst by attrition. Attrition-resistant catalysts have been developed to overcome this disadvantage (Sheffer, 1988).

Catalyst Regeneration

In some applications, catalyst activity can be restored by catalyst regeneration. There are currently three methods for regenerating catalyst activity.

Thermal Treatment involves elevating the catalyst to high temperatures sufficient to vaporize or oxidize the organic compounds or char that are masking the catalyst. The increased temperature is achieved by supplying additional heat from the system burner at the inlet to the catalyst chamber.

Physical Treatment uses mechanical means to remove particulates that are deposited on the catalyst. The most common approach is to blow compressed air and/or water across the surface of the catalyst to dislodge particulates.

Chemical Treatment involves cleaning with acid and/or alkaline solutions to remove compounds adhering to the catalyst surface. The catalyst modules are removed from the oxidizer and the chemical treatment performed either at the plant site by the user or returned to the catalyst supplier for treatment at their own cleaning facility. Chemical cleaning is the most frequently used cleaning procedure for catalysts in VOC applications. The treatment does not affect the catalyst composition, but merely removes masking compounds from the catalyst surface. Depending upon the specific application, preventative chemical cleaning maintenance can be scheduled from every six months in applications where catalyst masking is severe, to every three or more years. Chemical cleaning has been used for over 40 years and has proved to be an effective method for extending catalyst life in applications where masking occurs.

Flameless Thermal Oxidizers (FTO)

In this process, the exhaust stream is mixed with air before entering the flameless reactor vessel. The air mixture is evenly distributed into a bed of inert ceramic material coated with a metal catalyst. This bed provides complete mixing of the VOC with oxygen. The VOC oxidizes into carbon dioxide and water vapor once the mixture reaches the combustion temperature. The released combustion energy is absorbed by the ceramic bed and is transferred to the exhaust stream leaving the catalytic oxidizer. The temperature control of the system is very important in effective oxidation of VOCs. This process is a flameless incineration, as opposed to catalytic incineration, which uses an external fuel source. The catalytic oxidizer uses the heat of the exhaust to maintain combustion. To ensure the proper operation of the FTO, the exhaust gas entering the reactor needs to be at least 600°F. Ideally, the hydrocarbon concentration of the waste stream entering the system is high enough to generate the heat required to maintain the reaction.[28] However, FTOs can be used in applications where the hydrocarbon content of the waste stream is insufficient to maintain the optimal operating temperature. In such situations, electric heat or supplemental fuels must be added to the waste stream to maintain the optimal operating temperature.[29]

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[28] http://www.arb.ca.gov/pm/pmmeasures/ceffect/reports/sjvapcd_4692_report.pdf

[29] http://www.lindeus-engineering.com/internet.le.le.usa/en/images/FTO%20Technology%20Datasheet%200716%20Web136_279769.pdf?v=1.0 .

FTO offers the benefit of high temperature thermal destruction combined with the reliability of electric operation. Generally, the pre-heater of the FTO can be turned off as hydrocarbon rates increase. FTOs can also result in reduced power consumption via heat recovery systems that utilizes heat generated from hydrocarbon destruction to pre-heat the incoming vapors.

2.2.4 Other Considerations: Packaged versus Field-Erected Units, Auxiliary Equipment

Packaged vs. Field-Erected Units

Except for regenerative oxidizers, the equipment cost correlations included in this chapter are for packaged units only. They are not valid for field-erected units. For regenerative oxidizers, the correlations are valid for field-erected units only. Packaged units are units that have been shop fabricated and contain all elements necessary for operation, except for connection to facilities at the site, e.g., utilities. The elements include the combustion chamber, preheater, instrumentation, fan, and the necessary structural steel, piping, and electrical equipment. This equipment is assembled and mounted on a “skid” to facilitate installation on a foundation at the plant site. Tie-in to the local emission source is not part of the packaged unit. Units are usually sized to handle flow rates of < 20,000 scfm, but can be built to accommodate flow rates up to 50,000 standard cubic feet per minute (scfm). The cost correlations in this chapter are valid to 50,000 scfm for packaged units, except for fluid-bed units which are valid to 25,000 scfm.

Conversely, field-erected units may be built to any desired size. The combustion chamber, preheater, and other equipment items are designed and fabricated individually, and assembled at the site. However, both the equipment and installation costs of field-erected units are typically higher than those for equivalent-sized packaged units because the factors that improve efficiency of shop-fabrication, such as uniform working environment, availability of tools and equipment, and more efficient work scheduling, are generally not available in the field.

Acid Gas and Venturi Scrubbers

The final outlet stream of any incineration system may contain certain pollutants that must be removed. The combustion of sulfur-containing compounds results in SO_2, while chlorinated compounds yield Cl_2 and HCl in the outlet stream. These acid gases must be removed from the gas stream if they are present at significant concentrations (regulations for limits on these gases vary from state to state; Cl_2 and HCl are also regulated as HAP under CAA section 112 for many industries). These gases can be effectively removed using a packed-bed gas absorber (vertical scrubber) in which the flue gas is contacted with a caustic scrubbing liquid. For fluid-bed catalytic reactors, venturi scrubbers are often used because they also remove particulates from the flue gas. In addition, venturi scrubbers are often used to reduce mercury from sewage sludge incinerators as mentioned earlier in this chapter. In most cases adding a scrubber or absorber significantly increases the cost of the incineration unit, sometimes by a factor of two. More information on acid gas scrubbers is available in the “Acid Gas Scrubbers” chapter of this Manual (see Section 5, Chapter 1). More information on venturi scrubbers and other particulate control devices can be found in Section 6, chapters 1 through 3.

If chlorinated VOCs are present in the waste gas, heat exchangers may require special materials of construction. This added expense is not included in the costing procedures outlined in this section.

In most cases, acid gas scrubbers are located on the outlet of the thermal oxidizer. However, scrubbers may also be used on the inlet to the thermal oxidizer. This arrangement is sometimes used where the waste gas stream contains certain acidic sulfur or halogen compounds. The principal advantages of this arrangement are in reducing corrosion of oxidizer components and avoiding the additional costs associated with upgrading the materials used in oxidizer construction.

Scrubbers are sometimes included by vendors as part of an integrated scrubber/thermal oxidizer package. Although equipment costs quoted for the systems include the cost of the scrubber and thermal oxidizer, the additional costs associated with the purchase, handling, and disposal of the scrubbing medium and spent scrubbing solution and monitoring of scrubber flow rate and pressure drop should be included in the operating costs for the integrated system. For information on estimating the capital and operating costs of scrubbers, please refer to Sections 5 and 6 of the Cost Manual.

Heat Exchangers (Preheaters and Other Waste Energy Recovery Units)

For thermal and catalytic units that have some degree of energy recovery, the cost of the primary heat exchanger is included in the cost, and its design is usually done by the incineration unit vendor. The cost correlations presented in this chapter include units both with and without energy recovery. Secondary energy recovery, if desired, requires an additional heat exchanger, which is also often provided by the incineration unit vendor. Costing procedures for secondary energy recovery are not included in this section.

Concentrators

For processes that generate large volumes of waste gas with lower concentrations of VOC, plants may install concentrators before the oxidizer inlet. These systems use an adsorption system to convert the stream to a smaller volume of highly concentrated gas that is optimized for treatment by the oxidizer. Such optimization of treatment may lead to a smaller sized, and thus less costly, oxidizer. Typical concentrators use carbon or zeolite as the adsorbent. The waste gas entering the condenser passes through a cartridge where emissions are stripped from the air stream and adsorbed on the substrate. A small volume of heated air is passed over the substrate to release the VOC. The heated air with its high concentration of VOC can then be treated in the oxidizer. Various different designs of concentrators are available and may be purchased as part of an integrated oxidizer system.

Other Auxiliary Equipment

Additional auxiliary equipment such as hoods, ductwork, precoolers, cyclones, fans, motors, and stacks are addressed separately in other sections of this Manual.

2.2.5 Technology Comparison

Both the thermal and catalytic incineration systems are designed to provide VOC control through combustion at a level in compliance with applicable state and Federal requirements. Given the wide range of options available, however, it is obvious that not every incinerator will fulfill these requirements at the same cost. This section presents a first step toward deciding how best to deal with VOC emission abatement using incinerators or oxidizers considering some qualitative factors pertinent to the types of incinerators described in this chapter. It is the intent of the remainder of this section to provide a method by which the cost of VOC control for a particular application can be calculated.

A summary of the principal types of incinerators and oxidizers is presented in Table 2.5. From the earlier discussions, the following factors relating to the presence of contaminants should be considered by potential users (Stettenbenz, 1988):

- The fouling of the catalyst in a catalytic system is a possibility. Poisons to the system include heavy metals, phosphorous, sodium, sulfur and most halogens, although catalysts have been developed that are chlorine resistant, such as siloxane.

- The possibility of process upsets that could release any of the above poisons or cause fluctuations in the heating value to the incinerator would favor a thermal system.

- Except for low sulfur distillate fuels, fuel oil should not be used as auxiliary fuel to a catalytic system due to the sulfur and vanadium it may contain (Yarrington, 1989b).

All the above factors would serve to increase the operating expense of a catalytic unit through replacement costs of the catalyst. An additional factor relates to relative energy efficiency of the various types of incinerators:

- Thermal units generally require more auxiliary fuel than catalytic units and operate at temperatures that are roughly 1000^{o}F higher. This difference in fuel requirement increases as the heating value of the waste stream decreases.

In general, a trade-off exists between the higher capital costs of catalytic oxidizers and the higher operating costs of thermal oxidizers. This difference will be illustrated by a design example presented in Section 2.4, which describes both technologies, and cost analysis presented in Section 2.5, which compares the capital and operating costs of the two technologies.

Table 2.5: Principal VOC Incineration Technologies

General Type	Technology
Thermal Systems	Direct Flame Incinerator
	Recuperative oxidizer (Direct Flame with Recuperative Heat Exchanger)
	Regenerative oxidizer Operating in a Cyclic Mode
Catalytic Systems	Fixed Bed (Monolith or Packed-Bed)
	Fluid-Bed

2.3 General Treatment of Material and Energy Balances

In the sizing and costing of oxidizers and the calculation of the auxiliary fuel requirements, it is necessary to make material and energy balances around the entire incinerator unit and around certain parts of the unit, such as the combustion chamber or the preheater. This section presents a general approach to making these balances.

These balances are based on the law of conservation of mass and energy. They can be stated in general equation form as

$$\text{In} - \text{Out} + \text{Generation} = \text{Accumulation} \tag{2.4}$$

Because the incineration process is a steady-state process, the accumulation term is zero and the equation becomes:

$$\text{In} - \text{Out} + \text{Generation} = 0 \tag{2.5}$$

For mass balances, it is useful to restrict the balances to be made on the mass of each atomic species so that for mass balances the generation term becomes zero. However, because the combustion reaction liberates energy, the energy balances around equipment where combustion takes place would include a generation term. Thus, the simplified equations are

$$\text{In} - \text{Out} = 0, \text{for steady – state mass balances} \tag{2.6}$$

$$\text{In} - \text{O u t} + \text{Generation} = 0, \text{ for steady – state energy balances} \tag{2.7}$$

For the incineration process the two terms In and Out are generally mass terms (for a mass balance) of the form,

$$\dot{m} = \rho\, q \tag{2.8}$$

where

ρ = density (mass per unit volume)

Q = volumetric flow rate (volume per unit time) or sensible heat terms (for an energy balance) of the form,

$$\dot{Q} = \dot{m}\ \Delta\mathrm{h} = \rho\ \mathrm{q}\ \mathrm{Cp}\ (\mathrm{T} - T_{ref}) \tag{2.9}$$

where

C_p = heat capacity

T = temperature

The reference temperature, T_{ref}, is often taken to be zero or the temperature of a convenient stream, e.g., the inlet gas stream, in whatever units T is in, so the T_{ref} term may not appear in the equations. When the reference temperature is taken as zero, the sensible heat terms become

$$\dot{Q} = \rho\ Q\ C_p\ T \tag{2.10}$$

Energy losses, H_L, are also part of the *Out* term and, for the incinerator process, are taken here to be 10% of the total energy input to the incinerator.

For the incineration process, the generation term for energy balances accounts for the energy released through the combustion reactions. This term is generally of the form

$$\dot{Q} = \rho\ Q\ (-\Delta h_c) \tag{2.11}$$

Where

($-\Delta h_c$) = heat of combustion.

2.4 Design Procedures

The following procedure is designed to provide parameters for use in developing a study cost estimate (range of accuracy ± 30%) for thermal and catalytic oxidizers. The principal parameters of interest are

- flue gas flow rate, upon which all the equipment cost correlations are based.
- auxiliary fuel requirement, which is important in estimating annual operating costs.

For emissions sources that generate dilute mixtures of VOCs in air (i.e., waste streams that contain a minimum of 20% oxygen), the flue gas flow rate exiting the oxidizer is greater than the inlet waste gas flow rate by the amount of auxiliary fuel and the increase in the moles of gas as a result of the combustion reaction. Because these two factors usually cause only small increases in flow rate, a number of simplifying assumptions can be made in the design calculations. For applications where diluent air must be used to adjust the combustible concentration in the waste gas to 25% LEL and where auxiliary fuel and auxiliary combustion air are needed, more complete mass and energy balances must be made.

The design procedure illustrated below is for waste gas streams that are dilute mixtures of VOCs in air with at least 20% oxygen in the waste gas stream. In this discussion, the design procedure will be illustrated by a sample problem that will be solved step-by-step. In the example, the waste gas streams contain methylene chloride, a halogenated compound that can result in acid gases being exhausted from the oxidizer. In situations where the waste gas stream contains halogenated or sulfur containing compounds, the concentration of acidic gases in the outlet to the oxidizer should be calculated to determine whether an acid gas removal system, such as a wet scrubber, is needed. For information on how to design of acid gas controls, please see Section 5 of Cost Manual.

2.4.1 Steps Common to Thermal and Catalytic Units

This section describes the steps used to determine the basic design characteristics for thermal and catalytic oxidizers. To illustrate how the method should be applied, we have included an example problem. The specifications for the example emissions source are shown in Table 2.6.

Table 2.6: Specifications of Sample Problem

Variable	Value
Preheater Inlet Waste Gas Vol. Flow Rate, Q_{wi}, scfm	20,000
Preheater Inlet Waste Gas Temp., T_{wi}, °F	100
Composition	
• Benzene Content, ppmv	1,000
• Methyl Chloride Content, ppmv	1,000
• Air Content	Balance
Particulate Content	Negligible
Moisture Content	Negligible
Desired DRE, %	99
Desired Percent Energy Recovery, HR%	70

Step 1 - Establish design specifications. The first step in the design procedure is to determine the specifications of the incinerator and the waste gas to be processed. The following parameters of the waste gas stream at the emission source must be available:

- Volumetric flow rate, scfm[30]
- Temperature

---

[30] For the purposes of the example shown in this chapter, standard conditions are defined as a temperature of 77°F and 1 atmosphere pressure. A temperature of 77°F is a reference condition for ambient air quality standards as defined in 40 CFR 50.3 (Reference conditions), but is not standard temperature in other contexts.

- Oxygen content
- Chemical composition of the combustibles
- Inerts content
- Heating value—In certain cases the heating value may act as a surrogate for the chemical composition of the combustibles. This is particularly true for dilute mixtures of combustibles in air.
- Particulate content—The particulate content is important as particulates can coat the catalyst. An upstream filter may suffice if particulate content is too high. Fluid-bed catalytic oxidizers can tolerate higher particulate contents than fixed-bed catalytic oxidizers.

The following parameters must be specified for the incinerator:

- Desired DRE—This efficiency should be based on requirements dictated by relevant state and Federal regulations.
- Combustion chamber outlet temperature—This temperature may also be based on requirements of a regulation or on recommendations developed during regulatory development.
- Desired percent energy recovery—The desired percent energy recovery should be the result of a process optimization in which costs of incinerators with several different levels of energy recovery are estimated and the minimum cost design selected. The tradeoff is between the capital cost of the energy recovery equipment and the operating (fuel) cost.

Step 2 - Verify that the oxygen content of the waste gas is at least 20%. There must be sufficient oxygen in the waste gas to support the combustion of the waste organics (including VOCs) and the auxiliary fuel, if auxiliary fuel is needed. It may be necessary to add auxiliary air if the oxygen content is less than about 20%. In this example, the waste stream contains greater than 20% oxygen, as shown below for 1000 ppmv of benzene and methyl chloride:

$$Air\ Content, Vol.\% = 100 - \frac{1{,}000\ (ppm)}{10^6} \times 100 - \frac{1{,}000\ (ppm)}{10^6} \times 100 = 99.8\% \quad (2.12)$$

This gives the oxygen content in percent as:

$$Oxygen\ Content = 99.8\% \times 0.209 = 20.86\% \quad (2.13)$$

Where 0.209 is the concentration of oxygen in air.

Step 3 - Calculate the LEL and the Percent of the LEL of the gas mixture. If the waste stream contains a significant amount of inert constituents in addition to the nitrogen associated with the oxygen in air, the calculation of LEL (and UEL) loses meaning since the LEL (and UEL) is measured in mixtures of organic with air only. A full chemical analysis is necessary to complete the design procedure in such a case.

The example chosen here is typical, in that there is more than one VOC component in the gas stream. An approximate method to calculate the LEL of a mixture of compounds, LEL_{mix}, is given by Grelecki (Grelecki, 1976) as

$$LEL_{mix} = \left[\sum_{j=1}^{n} \frac{x_j}{\left(\sum_{j=1}^{n} x_i\right) LEL_j}\right]^{-1} \tag{2.14}$$

Where

X_i = volume fraction of combustible component *i*

LEL_i = lower explosive limits of combustible component *j* (ppmv)

N = number of combustible components in mixture

For the example case,

$$\sum_{i=1}^{n} X_i = (1{,}000 + 1{,}000) \times 10^{-6} = 2{,}000 \times 10^{-6}$$

From standard references (Lide, 2005) or from Appendix A,

$$LEL_{Bz} = 14{,}000\ ppmv\ for\ benzene$$
$$LEL_{MC} = 82{,}500\ ppmv\ for\ methyl\ chloride$$

$$LEL_{mix} = \left[\frac{1{,}000}{2{,}000 \times 14{,}000} + \frac{1{,}000}{2{,}000 \times 82{,}500}\right]^{-1} = 23{,}938\ ppmv$$

$$\%\ \mathrm{LEL_{mix}} = \frac{\text{total combustible conc.in mixture}}{\mathrm{LEL_{mix}}} \times 100 \tag{2.15}$$

$$\%\ \mathrm{LEL_{mix}} = \frac{2{,}000}{23{,}938} \times 100 = 8.4\%$$

The percent *LEL* of the mixture is therefore 8.4%. Because this is well below 25%, no dilution air is needed in this example. If the mixture had been above 25% *LEL,* sufficient dilution air would have been needed to bring the concentration of the mixture to less than 25% to satisfy fire insurance regulations.

Step 4 - Calculate the volumetric heat of combustion of the waste gas streams, (- Δh_{cw}), Btu/scf. The energy content of the gas stream, expressed in terms of the heat of combustion, is calculated as follows:

$$\left(-\Delta h_{c_w}\right) = \sum_{i=1}^{n}\left(-\Delta h_{c_i}\right) x_i \tag{2.16}$$

where

$(-\Delta h_{cw})$ = heat of combustion of the waste stream (Btu/scf)

$(-\Delta h_{ci})$ = volumetric heat of combustion of component i at 25°C (Btu/scf)

X_i = volume fraction of component i in the waste gas

n = number of combustible components in the waste gas

The heat of combustion that should be used in these calculations is the “lower” heat of combustion, i.e., with gaseous water, rather than liquid water, as a reaction product since water leaves the incinerator in the vapor state. From Appendix A or standard references (Lide, 2005 and Green, 1999) with appropriate conversion of units, the volumetric heat of combustion at 25°C for the two components is calculated to be as follows:

$(-\Delta h_{cBz})$ = 3,475 Btu/scf for benzene

$(-\Delta h_{cMC})$ = 705 Btu/scf for methyl chloride

The compositions specified earlier as ppmv are converted to volume fractions as

X_{Bz} = 1,000 ppmv × $10^{-6} = 10^{-3}$ for benzene

$X,_{MC}$ = 1,000 ppmv × $10^{-6} = 10^{-3}$ for methyl chloride

Using these values of heat of combustion and composition, the heat of combustion of the waste gas stream per standard cubic foot (scf) of incoming gas is

$$\left(-\Delta_{c_w}\right) = (3{,}475)(10^{-3}) + (705)\,(10^{-3}) = 4.18\ \frac{Btu}{scf}$$

Assuming the waste gas is principally air, with a molecular weight of 28.97 g/mole and a corresponding density of 0.0739 lb/scf, the heat of combustion per pound of incoming waste gas is:

$$\left(-\Delta h_{c_w}\right) = 56.6\ Btu/lb$$

The negative heat of combustion, by convention, denotes an exothermic reaction. Also by convention, if one refers to heat of reaction rather than heat of combustion, then a positive value denotes an exothermic reaction.

Empirically, it has been found that 50 Btu/scf roughly corresponds to the *LEL* of organic/air mixtures. Insurance codes require a value below 25% *LEL*, which corresponds to about 13 Btu/scf. However, if *LEL* sensors and monitors are installed, one can incinerate a waste gas with a combustible organic content between 25 and 50% *LEL*, which corresponds to 13 to 25 Btu/scf.

For catalytic applications, the heat of combustion must normally be less than 10 Btu/scf (for VOCs in air) to avoid excessively high temperatures in the catalyst bed. This is, of course, only an approximate guideline and may vary from system to system.

After Step 4, determination of the ($-\Delta h_{cw}$) design procedure for thermal and catalytic oxidizers is discussed separately, beginning with Step 5 for each type of incinerator.

2.4.2 Steps Specific to Thermal Units

Figure 2.1 shows a generic thermal oxidizer with the appropriate streams labeled.

Step 5t - Establish the temperature at which the incinerator will operate. As mentioned in Section 2.2.1, both the reactor temperature and residence time of the waste gas in the reactor determine the level of VOC destruction. In general, state and local regulations specify the required level of destruction that the customer must meet. In this example a destruction efficiency of 99% is specified. Based on Table 2.2, this destruction efficiency can be met in a thermal oxidizer operated at a temperature, T_{fi} of 1600°F and a residence time of 1 second. (Note: This higher efficiency level is the minimum achievable by any new properly designed and operated incinerator. Many incinerators can currently achieve destruction efficiencies of 99% or higher. This is consistent with a statement made by Anguil Environmental, Inc. in November 2015 (Chemical Engineering Magazine, 2015))

Step 6t - Calculate the waste gas temperature at the exit of the preheater. The extent of the heat exchange to be carried out in the preheater is the result of a technical and economic optimization procedure that is not illustrated in this example. As the VOC stream temperature leaving the heat exchanger, T_{wo}, increases, the auxiliary fuel requirement decreases, but at the expense of a larger heat exchanger. However, there are several important limits on T_{wo}. First, T_{wo} must not be close to the ignition temperature of the organic-containing gas to prevent damaging temperature excursions inside the heat exchanger should the gas ignite. Second, for gases containing halogens, sulfur, and phosphorous (or other acid-forming atoms), the flue gas temperature after the heat exchanger, T_{fo}, must not drop below the acid dew point. Both limitations minimize the amount of possible heat exchange and thus the maximum value of T_{wo}. The calculation of the acid dew point is not simple. It is recommended that vendor guidance be sought to ensure that the dew point is not reached. Condensation of acid gases will result in corrosion of many of the metallic construction materials used in heat exchangers. As an example, fuel sulfur contents of 1 to 2 percent can give acid dew points of about 200 to 270°F. Increasing the sulfur content to 4 percent can raise the dew to about 290°F. Chlorine and phosphorous have a much smaller effect on acid dew elevation.

With the following assumptions, one can estimate T_{wo} using Equation 2.2, the definition of fractional energy recovery for a heat exchanger:

- The fractional energy recovery is specified.

- The amount of auxiliary fuel, Q_{af} and auxiliary combustion air, Q_a, are small relative to the waste gas, Q_w, so that the mass flow rates of gases, p_iQ_f and p_fQ_f, on both sides of the preheater are approximately the same, or:

$$\rho_w \, Q_w = \, \rho_f \, Q_f \qquad (2.17)$$

- The heat capacities of the gases on both sides of the preheater are approximately the same, regardless of composition. This is true for waste streams which are dilute mixtures of organics in air, the properties of the streams changing only slightly on combustion.

- The mean heat capacities above the reference temperature of the gases on both sides of the preheater are approximately the same regardless of temperature.

With these assumptions, the equation for fractional energy recovery for a heat exchanger becomes

$$Fractional\ Energy\ Recovery = \frac{T_{wo} - T_{wi}}{T_{fi} - T_{wi}} \qquad (2.18)$$

For this example with a fractional energy recovery of 0.70, an incinerator operating temperature, T_{fi}, of 1600°F, and a waste gas inlet temperature, T_{wi} of 100°F, the waste gas temperature at the end of the preheater becomes:

$$T_{wo} = 1{,}150°F$$

The temperature of the exhaust gas, T_{fo}, can be determined by an energy balance on the preheater, which, with the same assumption as used in deriving Equation 2.18 regarding the mass flow rates and average heat capacities of the gases involved, results in the following equation:

$$T_{f_i} - T_{f_o} = T_{w_o} - T_{w_i} \qquad (2.19)$$

e.g., the temperature rise in the waste gas is approximately equal to the temperature decrease in the flue gas with which it is exchanged. For this example, this results in

$$T_{f_o} = \ 550°F$$

This value of T_{f_o} should be well above the acid dew point of the flue gas stream.

It should be remembered that T_{wo} should be well below the ignition temperature of the VOC stream to prevent unwanted temperature excursions in the preheater. This must be verified even if the stream is well below the *LEL* because flammability limits can be expanded by raising the reactant stream temperature. A sufficiently high preheat temperature, T_{wo} could initiate reaction (with heat release) in the preheater. This would ordinarily be detrimental to the materials of construction in the heat exchanger. The one exception is the thermal oxidizer of the regenerable type described in Section 2.1. The 95-percent energy recovery, obtainable in regenerable systems would result in this example in a T_{wo} of 1,525°F. The significant reaction rate that would occur at this temperature in the ceramic packing of the heat exchanger/reactor is by design.

Step 7t - Calculate the auxiliary fuel requirement, Q_{af}. Auxiliary fuel will almost invariably be needed for startup of the unit. However, at steady state, if the energy released by combustion of the organics present in the waste stream is sufficient to maintain the reactor temperature (1600°F in the example), only a small amount of auxiliary fuel (< 5% of the total energy input) is needed to stabilize the flame. In most cases, however, more fuel than just this stabilizing fuel will be required to maintain the reactor temperature.

With the following assumptions, one can estimate Q_{af} using a mass and energy balance around the combustion chamber and following the principles discussed in Section 2.2, with reference to Figure 2.1.

- The reference temperature, T_{ref}, is taken as the inlet temperature of the auxiliary fuel, T_{af},
- No auxiliary air, Q_a, is required.
- Energy losses, H_L, are assumed to be 10% of the total energy input to the incinerator above ambient conditions (Schmidt, 1989a, and Yarrington, 1989a). Thus, if the reference temperature is near ambient condition

$$H_L = 0.1\,\rho_{fi}\,Q_{fi}\,\,C_{pm_{fi}}\,(T_{fi} - T_{ref}) \tag{2.20}$$

- The heat capacities of the waste gases entering and leaving the combustion chamber are approximately the same, regardless of *composition*. This is true for waste streams which are dilute mixtures of organics in air, the properties of the streams changing only slightly on combustion.
- The mean heat capacities above the reference temperature of the waste gases entering and leaving the combustion chamber are approximately the same regardless of temperature. Thus the mean heat capacity for the waste gas stream entering or leaving the combustion chamber should be evaluated at the average of T_{wo} and T_{fi}. For air this assumption introduces an error of, at most, 5% over the temperatures of interest.

With these assumptions, the mass and energy balance around the combustion chamber reduces to the following equation:

$$\rho_{af} Q_{af} = \frac{\rho_{w_o} Q_{w_o}\left[C_{pm_{air}}\left(1.1T_{f_i} - T_{w_o} - 0.1T_{ref}\right) - \left(-\Delta h_{c_{w_o}}\right)\right]}{\left(-\Delta h_{c_{a_f}}\right) - 1.1C_{pm_{air}}\left(T_{f_i} - T_{ref}\right)} \tag{2.21}$$

The waste stream is essentially air so that

$$\rho_{wo} = \rho_{wi} = 0.0739 \text{ lb/scf, air at 77°F, 1 atm}$$

C_{pmair} = 0.255 Btu/lb F, the mean heat capacity of air between 77°F and 1375°F (the average pm air temperature of the waste gas entering the leaving the combustion chamber)

Other input data to Equation 2.21 include

Q_{wo} = Q_{wi} = 20,000 scfm

$(-\Delta h_{caf})$ = 21,502 Btu/lb, for methane

$T_{af} = T_{ref}$ = 77°F, assume ambient conditions

ρ_{af} = 0.0408 lb/ft^3, methane at 77°F

T_{fi} = 1,600°F, Step 5t

T_{wo} = 1,150°F, Step 6t

$(-\Delta h_{cwo})$ = 56.6 Btu/lb, Step 4

Substituting the above values into Equation 2.21 results in:

$$Q_{af} = 167 \text{ scfm}$$

The values of the parameters in the energy balance are summarized in Table 2.7.

It is instructive to examine the magnitude of the various terms in the energy balance around the combustor for the sample problem. This is done in Table 2.8. The energy balance shown does not quite add to zero due to round-off-error and simplifying assumptions. Table 2.8 shows that the largest inlet term is the sensible heat of the incoming waste gas. The heat of combustion of the organics contained in the waste gas stream is somewhat smaller than that of the auxiliary methane because of the relatively small amount of organics in the waste gas stream. The largest term in the outlet stream is the sensible heat of the outgoing waste stream. The overall energy losses are based on an assumption, but are relatively small. Because the sensible heat contents of the entering and leaving waste stream are so large, it is apparent that energy recovery is an important factor in achieving energy efficiency. In fact, with zero energy recovery in the sample problem, the auxiliary fuel requirements would be 605 scfm, about four times the energy requirements based on 70% energy recovery.

Table 2.7: Summary of Example Problem Variable Valuation T_{ref} = 77°F

Stream		Subscript,j	ρ_j, lb/scf	Q_j, scfm	C_{pmj}, Btu/lb°F	T_j,°F
IN--Sensible Heat						
	Auxiliary Air	a	na	na	na	na
	Auxiliary Fuel	af	0.0408	167	1	77
	Waste Gas	w_o	0.0739	20,000	0.255	1,150
OUT--Sensible Heat						
	Waste Stream	f_i	0.0739	20,167	0.255	1,600

$(-\Delta h)$, waste gas = 56.6 Btu/lb
$(-\Delta h)$, auxiliary fuel = 21,502 Btu/lb na = Not Applicable
1 = Not used because reference temperature is taken equal to auxiliary fuel temperature.

Table 2.8: Terms in Energy Balance Around Combuster – Example Problem

Stream		Subscript,j	Value, Btu/min
IN--Sensible Heat $\rho_j Q_j C_{pmj}$ $(T_i - T_{ref})$			
	Auxiliary Air	a	0
	Waste Gas	w_o	404,403
OUT--Sensible Heat $\rho_j Q_j C_{pmj}$ $(T_i - T_{ref})$			
	Waste Stream	f_i	578,796
OUT--Losses			
	10% of total energy input		57,800
GENERATION-- Heat of Combustion, $\rho_j Q_j$ $(-\Delta h_{cj})$			
	Waste Gas	w_o	83,655
	Auxiliary Fuel	af	146,506

Step 8t - Verify that the auxiliary fuel requirement is sufficient to stabilize the burner flame. Only a small amount of auxiliary fuel (< 5% of the total energy input) is needed to stabilize the burner flame. In general, more fuel than just this stabilizing fuel will be required to maintain the reactor temperature. It is wise to verify that the auxiliary fuel requirement calculated in Step 7t is sufficient for stabilization. If it is insufficient, then a minimum amount of auxiliary fuel must be used, and the amount of energy recovery, specified earlier must be reduced to avoid exceeding the specified temperature at which the incinerator will operate (Step 5t).

This check is made by calculating 5% of the total energy input to the incinerator and comparing it with the auxiliary fuel energy input. The total energy input is given as follows:

$$Total\ Energy\ Input = \rho_{fi}\, Q_{fi}\, C_{pmfi}\, (T_{fi} - T_{ref}) \tag{2.22}$$

$$Auxiliary\ Fuel\ Energy\ Input = \rho_{af} Q_{af} (-\Delta h_{caf}) \tag{2.23}$$

The auxiliary fuel used in the design, Q_{af}, should be the larger of 5% of the total energy input (28,900 Btu/min.) and the auxiliary fuel energy input (146,500 Btu/min). The auxiliary fuel used easily meets this criterion.

Step 9t - Calculate the total volumetric flow rate of gas through the incinerator, Q_{fi}.. The total volumetric flow rate of gas leaving the incinerator is referred to as the flue gas flow rate, Q_{fi}, and is the gas rate on which the incinerator sizing and cost correlations are based. The flue gas flow rate measured at the standard conditions of 77°F and 1 atmosphere, where the increase in volumetric throughput due to an increase in the number of moles of gas as a result of combustion is neglected, is the sum of the inlet streams to the incinerator.

$$\begin{aligned} Q_{fi} &= Q_{wo} + Q_a + Q_{af} \\ &= 20{,}000 + 0 + 167 \\ &= 20{,}167\ scfm \end{aligned} \tag{2.24}$$

This result conforms to the assumptions stated in Step 6t, e.g., the mass (and volume) flow rates on both sides of the preheater are approximately equal. Finally, it must be emphasized that steps 5t to 9t apply to thermal recuperative oxidizers only. To calculate the auxiliary fuel requirements for other types of thermal oxidizers (e.g., regenerative), a different procedure must be used. (See Appendix B.)

2.4.3 Steps Specific to Catalytic Units

Figure 2.3 shows a generic catalytic oxidizer with the appropriate streams labeled. The approach used in the calculations on the catalytic oxidizer is somewhat different than that used in the thermal oxidizer. This difference arises because of additional constraints which are placed on the catalytic oxidizer. These constraints are as follows:

- The desired catalyst bed outlet temperature is typically 700 to 900°F. The maximum temperature to which the catalyst bed can be exposed continuously is limited to about 1200°F. Therefore, the combustible content of the waste gas is limited, and the amount of heat exchange that occurs in the primary heat exchanger may be limited.

- The inlet temperature to the catalyst bed itself must be above the catalytic ignition temperature required to give the desired destruction efficiency in the incinerator. Therefore, the combustible content of the waste gas is further limited to that which, when combusted, will raise the temperature in the catalyst bed no more than the ΔT between the required reactor bed inlet temperature, and the desired reactor bed outlet temperature.

- Auxiliary fuel, in combination with the preheater from the primary heat exchanger, is used to preheat the waste gas to the reactor inlet temperature. A minimum amount of auxiliary fuel (< 5% of the total energy input) must be used to stabilize the flame in the preheat combustion chamber. This has the effect of further limiting the combustible content of the waste gas stream and the amount of heat exchange permissible in the primary heat exchanger.

The steps outlined below represent one approach to recognizing these constraints and incorporating them into the calculation procedures.

Step 5c - Establish the desired outlet temperature of the catalyst bed, T_{fi}. The energy released by the oxidation of the VOCs in the catalyst bed will raise the temperature of the gases by an amount, ΔT, as the gases pass through the catalyst bed. An outlet temperature from the catalyst, and thus from the reactor, must be specified that will ensure the desired level of destruction of the VOC stream. As in thermal oxidizers, this value varies from compound to compound and also varies from catalyst to catalyst. Final design of the incinerator should be done by firms with experience in incinerator design. Manufacturers indicate that values from 400 to 900°F result in destruction efficiencies above 95 percent (Nester). To prevent deactivation of the catalyst a maximum bed temperature of 1200°F should not be exceeded. In the example problem, the catalyst outlet temperature, T_{fi}, is selected to be 900°F.

Step 6c - Calculate the waste gas temperature at the exit of the preheater (primary) heat exchanger. The waste gas temperature at the exit of the primary heat exchanger is estimated in the same manner as for the thermal oxidizer. The equation for fractional energy recovery Equation 2.18, is used, with the same assumptions as used for the thermal oxidizer. For the example problem with a fractional energy recovery of 0.70, a catalyst bed outlet temperature, T_{fi}, of 900°F, and a waste gas inlet temperature, T_{wi}, of 100°F becomes

$$T_{wo} = 660°\text{F}$$

The same considerations regarding the closeness of the temperature of the exhaust gas, T_{fa}, to its dew point apply to the catalytic oxidizer as they did to the thermal oxidizer.

Step 7c - Calculate the auxiliary fuel requirement, Q_{af}. The auxiliary fuel requirement, Q_{af}, is calculated by making mass and energy balances around the preheater combustion chamber and the catalyst chamber. The auxiliary fuel requirement calculated in this manner must be checked to insure that it falls within the constraints imposed by design considerations of the catalytic oxidizer. These constraints are as follows:

- The auxiliary fuel requirement must be positive. A negative fuel requirement indicates that the heat of combustion of the waste gas, ($-\Delta hc$), is too high for the fractional energy recovery in the primary heat exchanger that was selected.

- The auxiliary fuel amount must be high enough to provide a stable flame in the preheater combustion chamber (See Step 8c below).

An energy balance around the preheater combustion chamber and the catalyst chamber, taken together, results in Equation 2.21, the same equation used in the thermal oxidizer calculations. The input data for Equation 2.21 for the catalytic oxidizer example problem are summarized below:

- The waste stream is essentially air so that

ρ_{wo} = ρ_{wi} = 0.0739 lb/scf, air at 77°F, 1 atm

C_{pmair} = 0.248 Btu/lb F, the mean heat capacity of air between 77°F and 780°F (the average of the preheater exit and catalyst bed outlet temperatures)

- Other input data to Equation 2.21 include:

$$Q_{wo} = Q_{wi} = 20{,}000 \text{ scfm}$$

($-\Delta h_{caf}$) = 21,502 Btu/lb, for methane

$T_{af} = T_{ref}$ = 77°F, assume ambient conditions

ρ_{af} = 0.0408 lb/ft3, methane at 77°F

T_{fi} = 900°F, from Step 5c

T_{wo} = 660°F, from Step 6c

$(-\Delta h_{cwo})$ = 56.6 Btu/lb, from Step 4

Substituting the above values into Equation 2.21 results in

$$Q_{af} = 40 \text{ scfm}$$

If the outlet temperature of the catalyst bed, T_{fi}, is 800°F, then Q_{af}, decreases to -6.7 scfm. In other words, no auxiliary fuel would, theoretically, be required at this bed temperature. However, as discussed above in Step 8t, a certain quantity of auxiliary fuel would be required to maintain burner stability.

At 70% energy recovery and 900°F outlet catalyst bed temperature, a waste gas with a heat combustion $(-\Delta h_{cwo})$, of about 79.9 Btu/lb would cause the auxiliary fuel requirement, Q_{af}, to become negative, indicating the catalyst bed would exceed 900°F. At 70% energy recovery and 800°F outlet catalyst bed temperature, the same result occurs with a $(-\Delta h_{cwo})$ of 52.7 Btu/lb. Both of these heats of combustion are relatively low for typical waste gases. These results are, of course, dependent on the assumption of energy losses from the combustion chamber. The lower the energy losses, the lower the allowable waste gas heat of combustion before overheating occurs in the catalyst bed.

Step 8c - Verify that the auxiliary fuel requirement is sufficient to stabilize the burner flame. Only a small amount of auxiliary fuel (< 5% of the total energy input) is needed to stabilize the burner flame. In general, more fuel than just this stabilizing fuel will be required to maintain the reactor temperature. It is wise to verify that the auxiliary fuel requirement calculated in Step 7c is sufficient for stabilization. If it is insufficient, then a minimum amount of auxiliary fuel must be used and the amount of energy recovery specified earlier must be reduced to avoid exceeding the specified temperature at which the incinerator will operate (Step 5c).

This check is made in the same manner as that in Step 8t of the thermal oxidizer calculation. The results of this check indicate that the auxiliary fuel requirement is more than sufficient to stabilize the burner flame.

Step 9c - Estimate the inlet temperature to the catalyst bed, T_{ri}. The inlet temperature to the catalyst bed must be calculated to ensure that the inlet temperature is above that necessary to ignite the combustible organic compounds in the catalyst that was selected for use.

The inlet temperature to the catalyst bed, T_{ri}, should be such that, when the temperature rise through the catalyst bed, T, is added to it, the resulting temperature is T_{fi}, 900°F. Thus,

$$\Delta T = T_{fi} - T_{ri} \qquad (2.25)$$

The value of ΔT is determined by an energy balance around the preheater portion of the combustor. The preheater is required to heat the gases up to the catalyst bed inlet temperature using auxiliary fuel.[31] This energy balance is prepared with the assumptions made earlier in deriving Equation 2.21 and further assuming that only auxiliary fuel is combusted in the preheater portion. The resulting equation is very similar to Equation 2.21 except that (1) the terms with an f_i subscript become terms with r_i subscripts to denote a catalytic reactor inlet stream rather than a combustor outlet (flue gas inlet to the primary heat exchanger) and (2) the term for combustion of the waste gas organics does not appear. The resulting equation is as follows:

$$\rho_{af}Q_{af} = \frac{\rho_{w_o}Q_{w_o}\left[C_{pm_{air}}\left(1.1T_{r_i} - T_{w_o} - 0.1T_{ref}\right)\right]}{\left(-\Delta h_{c_{af}}\right) - 1.1C_{pm_{air}}\left(T_{r_i} - T_{ref}\right)} \tag{2.26}$$

This equation may be rearranged to solve for T_{ri} explicitly. This produces an equation that is somewhat complex and non-intuitive.

$$T_{r_i} = \frac{\rho_{af}Q_{af}\left[\left(-\Delta h_{c_{af}}\right) + 1.1C_{pm_{air}}T_{ref}\right] + \rho_{w_o}Q_{w_o}C_{pm_{air}}\left(T_{w_o} + 0.1T_{ref}\right)}{1.1C_{pm_{air}}\left(\rho_{af}Q_{af} + \rho_{w_o}Q_{w_o}\right)} \tag{2.27}$$

After substituting the example problem parameters into Equation 2.28, we obtain a value for T_{ri} of 693°F. Based on ignition temperatures shown in Table 2.3, this reactor inlet temperature should be satisfactory. Prior to a more definitive design, the ignition temperatures for the specific chemicals should be verified.

The temperature rise across the catalyst bed is thus (900 - 693) or 207°F. These temperatures are somewhat sensitive to the assumption for energy losses from the combustor. The assumption for energy losses is perhaps somewhat conservative, i.e., it causes a larger Q_{af} to be estimated than would a less conservative assumption, and becomes more conservative as the combustor size and insulation are increased.

Step 10c - Calculate the total volumetric flow rate of gas through the incinerator, $\boldsymbol{Q_{fi}}$**.** The total volumetric flow rate of gas leaving the incinerator is referred to as the flue gas flow rate, Q_{fi}, and is the gas rate on which the incinerator sizing and cost correlations are based. The flue gas flow rate measured at the standard conditions for temperature and pressure of 77°F and 1 atmosphere, where the increase in volumetric throughput due to an increase in the number of moles of gas as a result of combustion is neglected, is the sum of the inlet streams to the thermal oxidizer, as defined in the following equation:

$$\begin{aligned} Q_{fi} &= Q_{wo} + Q_a + Q_{af} \\ &= 20{,}000 + 0 + 40 \\ &= 20{,}040\ scfm \end{aligned}$$

Step 11c - Calculate the volume of catalyst in the catalyst bed. If the volumetric flow rate of gas through the catalyst bed, Q_{fi} and the nominal residence time (reciprocal space velocity) in the

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[31] At equilibrium, the temperature of the catalyst bed is maintained without requiring auxiliary fuel.

catalyst bed are known, then the volume of catalyst can be estimated. There exists a complex set of relationships between the catalyst volume and geometry, overall pressure drop across the catalyst, conversion of the oxidizable components in the gas, gas temperature, and the reaction rate. These relationships are dependent on the catalyst and the type of compound being oxidized. It is beyond the scope of this Manual to discuss these relationships, even in an approximate way. For the purposes of cost estimation, the space velocity, in reciprocal time units, necessary to achieve the required level of destruction can be used to approximate the catalyst volume requirement. The space velocity is defined as

$$\Phi = \frac{Q_{ft}}{V_{cat}} \tag{2.28}$$

where

V_{cat} = Overall bulk volume of the catalyst bed, including interparticle voids (ft^3)

By petro-chemical industry convention, the space velocity is computed at the conditions of 60°F (not 77°F) and 1 atm. The volumetric flow rate, Q_{fi}, must be corrected to these conditions. The proper space velocity to achieve a desired level of conversion is based on experimental data for the system involved. For precious metal monolithic catalysts, the space velocity generally lies between 10,000 h^{-1} and 60,000 h^{-1}. Base metal catalysts operate at lower space velocities, ranging from 5,000 to 15,000 h^{-1} (Yarrington, 1989b).

There are a number of catalyst bed parameters, such as catalyst configuration and bed design that are not significant for study type cost estimates. Accordingly, design of these factors is not discussed here.

2.5 Cost Analysis for Thermal and Catalytic Oxidizers

This section presents procedures and data for estimating capital and annual costs for four types of incinerators:(1) thermal-recuperative, (2) thermal regenerative, (3) fixed-bed catalytic, and (4) fluid-bed catalytic.

2.5.1 Estimating Total Capital Investment

Total capital investment, (TCI), includes the equipment cost, EC, for the incinerator itself, the cost of auxiliary equipment (e.g., ductwork), all direct and indirect installation costs, and costs for buildings, site preparation, offsite facilities, land, and working capital. However, the last five capital cost items usually do not apply to incinerators. (See Section 1, Chapter 2 (Cost Estimation: Concepts and Methodology) of this Manual for a detailed description of the elements comprising the TCI.) In the previous edition of this chapter, although industry representatives were reluctant to provide updated costs in 1999 dollars, they did indicate costs had not significantly changed since 1988 (Masonick, 1999, Raemhild, 1999, and Wilke, 1999) and 11 quotes from vendors for a specific configuration for three types of incinerator systems (recuperative, regenerative, fixed-bed catalytic) compared favorably to those generated using the cost equations.

2.5.1.1 Equipment Costs, EC

As discussed in Section 2.2.3, the EC given in this chapter apply to packaged incinerators, except for regenerative oxidizers. For regenerative oxidizers, the costs apply to field-erected units. The EC typically includes all flange-to-flange equipment needed to oxidize the waste gas, including the auxiliary burners, combustion chamber, catalyst, primary heat exchanger (except for the “zero heat recovery” cases), weathertight housing and insulation, fan, flow and temperature control systems, a short stack, and structural supports. Smaller units, e.g., typically less than 20,000 scfm, are typically preassembled skid-mounted (Githens, 1978). The various available incineration systems are presented in four groups delineated according to their similarity of design. These groups are outlined in Table 2.9. With the exception of regenerative thermal and fluid-bed catalytic oxidizers, the maximum size for which costs are given is 50,000 scfm. Although larger units of each technology can be built, applications are rare at flow rates above 50,000 scfm. Regenerative thermal oxidizer costs are provided for flow rates from 10,000 to 100,000 scfm. Fluid-bed catalytic oxidizer costs are provided for flow rates from 2,000 to 25,000 scfm.

Table 2.9: Scope of Cost Correlations

Incinerator Type	Total (Flue) Gas Flowrate, scfm	Figure Number
Thermal – Recuperative	500^a – 50,000	2.4
Thermal – Regenerative	10,000 – 100,000	2.5
Fixed-Bed Catalytic	2,000 – 50,000	2.6
Fluid-Bed Catalytic	2,000 – 25,000	2.7

[a] Although Figure 2.4 covers the 1,000 to 50,000 scfm range, the correlation is valid for the 500 to 50,000 scfm range.

The cost curves are least-squares regressions of cost data provided by different vendors. It must be kept in mind that even for a given incineration technology, design and manufacturing procedures vary from vendor to vendor, so that costs may vary. As always, once the study estimate is completed, it is recommended that more than one vendor be solicited for a more detailed cost estimate.

The additional expense of acid gas clean-up or particulate control is not treated in this section. The equipment cost of a gas absorber to remove any acid gases formed in the incinerator can be quite large, sometimes exceeding the equipment cost of the incinerator itself even for simple packed tower scrubbers (Jones, 1988). For more complex absorbers that include venturi scrubbers instead of, or in addition to, packed beds, the cost of the scrubber alone may be up to 4 times that of the incinerator (Sheffer, 1988). These more complex absorbers are sometimes necessary when particulates, in addition to acid gases, must be removed from the flue gas. Costs for acid gas clean-up or particulate control are covered in other chapters of this Cost Manual.

Thermal oxidizers. Among the thermal units, the direct flame (0% energy recovery) and recuperative systems are treated together because the various levels of energy recovery are achieved simply by adding heat exchanger surface area. Costs for these units were provided by several vendors (Bumford, 1988, Martinson, 1988, and Stettenbenz, 1988). The EC of these units are given as a function of total volumetric throughput, Q_{tot}, in scfm. “Q_{tot}”, is the total volume of the gaseous compounds exiting the combustion chamber; it is identical to the term, "Q_{fi}," used in Figures 2.1 and 2.2. This includes the combustion products, nitrogen, unburned fuel and organics, and other constituents. (See Figure 2.4). Note that costs are given free on board (F.O.B.) in April 1988 dollars[32]. Based on a least-squares regression analysis, a log-log relationship between throughput and EC was found for a given level of energy recovery (*HR*) over the flow rate range from 500 to 40,000 scfm. These relationships are as follows:

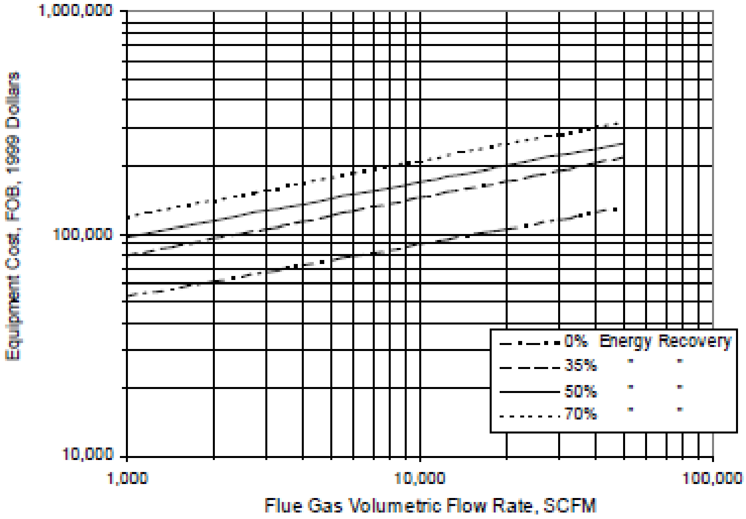

Figure 2.4: Equipment Costs of Thermal Oxidizers, Recuperative

$$EC = 10{,}294\ \ Q_{tot}^{0.2355}\ \ \text{HR} = 0\% \tag{2.29}$$

$$EC = 13{,}149\ \ Q_{tot}^{0.2609}\ \ \text{HR} = 35\% \tag{2.30}$$

$$EC = 1{,}7056\ \ Q_{tot}^{0.2502}\ \ \text{HR} = 50\% \tag{2.31}$$

$$EC = 21{,}342\ \ Q_{tot}^{0.2500}\ \ \text{HR} = 70\% \tag{2.32}$$

The regenerative (or excess enthalpy) systems provide up to 95 percent heat recovery at the expense of higher capital costs. Their unique design (Renko, 1988, and Mueller, 1988), which

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[32] For escalating these and the other incinerator and oxidizer prices to more current dollars, EPA suggests that the Chemical Engineering Plant Cost Index (CEPCI) could be useful. This cost index covers construction and equipment costs for the chemical process industries, a likely sector for installation of these control devices. The CEPCI is available at http://www.chemengonline.com/pci. Mention of the CEPCI in this Manual is not meant to serve as endorsement for commercial purposes.

combines the heat exchanger and reactor, is substantially different from traditional thermal units and is therefore treated separately in Figure 2.5. The ECs of these systems are given as an approximately linear function of total flow rate over a 10,000 to 100,000 scfm range by the following equation based on data for units installed between 2015 and 2016 (ICAC, 2016):

$$EC = 2.664 \times 10^5 + 13.98\, Q_{tot} \tag{2.33}$$

Again, the higher capital costs of these units can be substantially offset by the substantial savings in auxiliary fuel costs.

Catalytic oxidizer. The EC for a catalytic oxidizer is a function of the type of catalyst contacting pattern used and the total gas flow rate, Q_{tot}, for a given level of energy recovery. There are three types of contacting configurations used in catalytic systems: fixed-bed, catalytic monolith, and fluid-bed. The EC for the first two are generally comparable and are given in Figure 2.6. The data provided by several vendors (Bumford, 1988, Hablewitz, 1988, Martinson, 1988, and Stettenbenz, 1988) exhibited curvilinear relationships with Q_{tot} for each of the energy recovery rates. Least squares regressions of the data yielded the following correlations for total flow rates between 2,000 and 50,000 scfm:

$$EC = 1105 \quad Q_{tot}^{0.5471} \quad \text{HR = 0\%} \tag{2.34}$$

$$EC = 3623 \quad Q_{tot}^{0.4189} \quad \text{HR = 35\%} \tag{2.35}$$

$$EC = 1215 \quad Q_{tot}^{0.5575} \quad \text{HR = 50\%} \tag{2.36}$$

$$EC = 1443 \quad Q_{tot}^{0.5527} \quad \text{HR = 70\%} \tag{2.37}$$

Fluid-bed catalytic oxidizers afford certain advantages over fixed-bed catalyst units in that they tolerate waste streams with (1) higher heating values, (2) particulate contents, and (3) chlorinated species. For this enhanced flexibility of feed streams, a higher capital cost is incurred, as indicated by the EC shown in Figure 2.7.

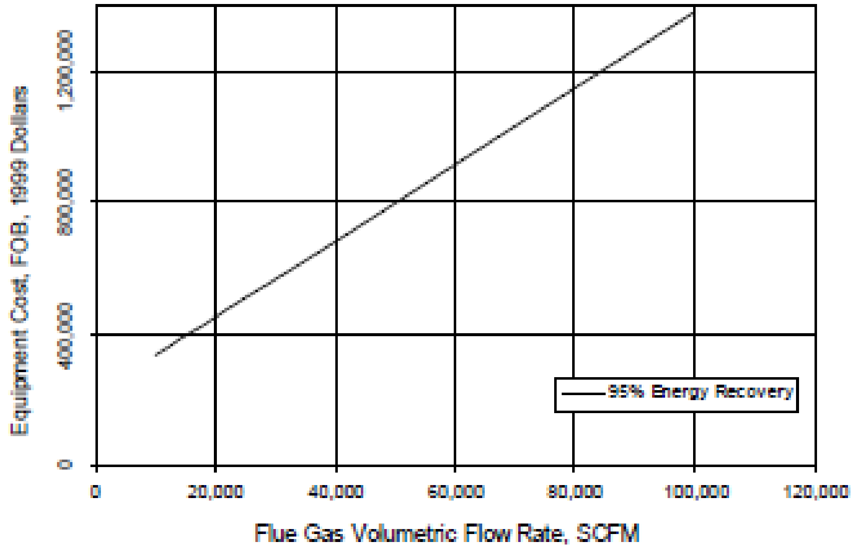

Figure 2.5: Equipment Costs of Thermal Oxidizers, Regenerative

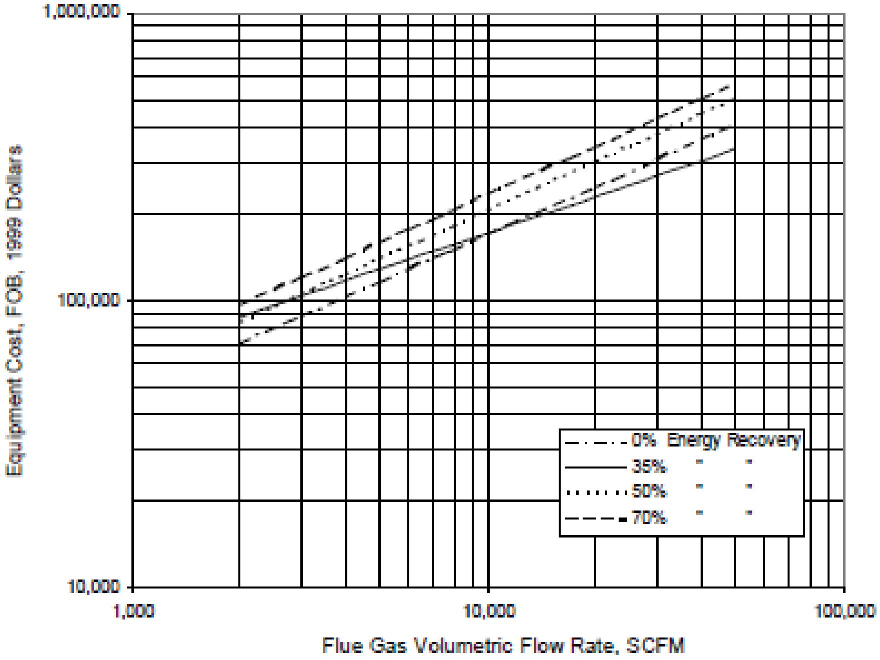

Figure 2.6: Equipment Cost of Catalytic Oxidizers, Fixed-Bed

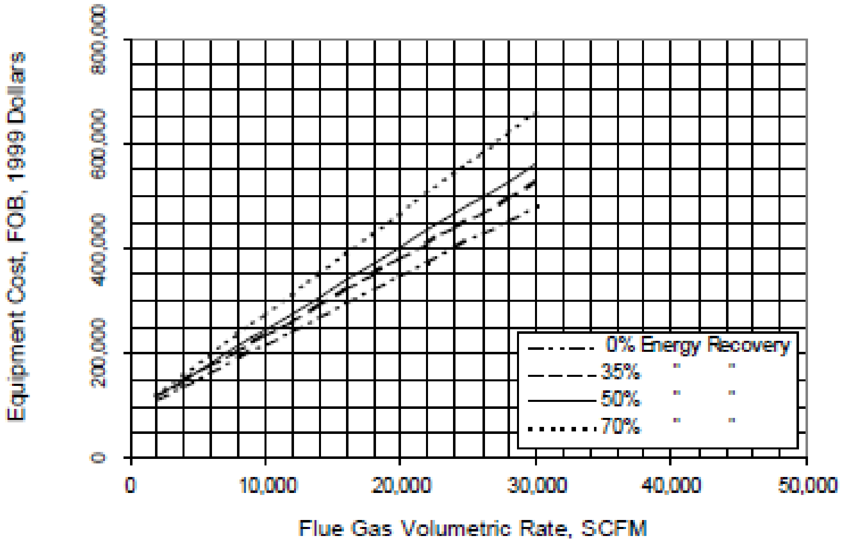

Figure 2.7: Equipment Costs of Catalytic Oxidizers, Fluid-Bed

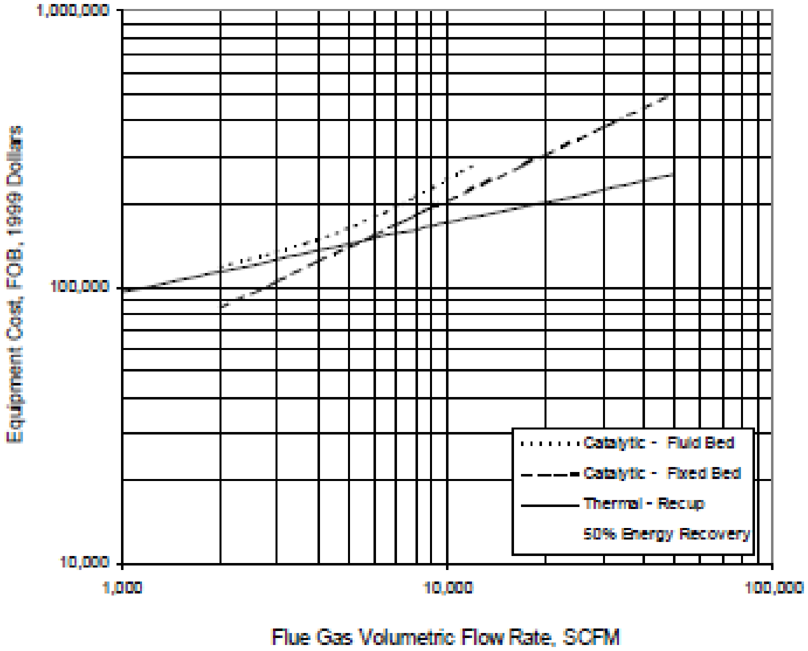

Figure 2.8: Equipment Costs Comparison of Incinerator Types

The data shown were provided by vendors (Sheffer, 1998, and Jones, 1988) and exhibited a linear relationship over the range of flow rates from 2,000 to 25,000 scfm. They can be approximated by the following equations:

$$EC = 8.48 \times 10^4 + 13.2\, Q_{tot} \quad \text{HR} = 0\% \qquad (2.38)$$

$$EC = 8.84 \times 10^4 + 14.6\, Q_{tot} \quad \text{HR} = 35\% \qquad (2.39)$$

$$EC = 8.66 \times 10^4 + 15.8\, Q_{tot} \quad \text{HR} = 50\% \qquad (2.40)$$

$$EC = 8.39 \times 10^4 + 19.2\, Q_{tot} \quad \text{HR} = 70\% \qquad (2.41)$$

A comparison of the thermal, catalytic fixed-bed, and catalytic fluid-bed systems with 50 percent energy recovery is shown in Figure 2.8.

For the example problem in Table 2.6, the equipment costs can be calculated using the equations for 70 percent heat recovery and 20,000 scfm. For this example, the auxiliary equipment costs are assumed to be negligible. Hence, the equipment costs for the regenerative thermal oxidizer are calculated using equation 2.33, as follows:

$$EC = 21{,}342 Q_{tot}^{0.2500} = (21{,}342)(20{,}167)^{0.2500} = \$254{,}000$$

For the fluid-bed catalytic oxidizer in our example, the equipment costs are calculated using equation 2.42, as follows:

$$EC = 8.39 \text{ x } 10^4 + 19.2\ Q_{tot} = (8.39 \text{ x } 10^4) + (19.2)(20{,}040) = \$469{,}000$$

2.5.1.2 Installation Costs

As explained in Section 1, the purchased equipment cost, PEC, is calculated by taking the sum of the EC and the cost of auxiliary equipment (e.g., ductwork), taxes, freight, and instrumentation. Average values of direct and indirect installation factors (Vatavuk, 1980) to be applied to the PEC are given in Table 2.10 for both recuperative thermal and fixed- and fluid-bed catalytic oxidizers.

Table 2.11 shows the itemized installation costs that are obtained when these installation factors are applied to the PECs for the example incinerators. Depending on the site conditions, the installation costs for a given incinerator could deviate significantly from costs generated by these average factors. Oxidizers installed on existing emissions units may have higher installation costs. For example, retrofits may require longer ductwork to transfer the waste gas from the emissions unit to the control device. Vatavuk and Neveril (Vatavuk, 1980) provide some guidelines for adjusting the average installation factors to account for other-than-average installation conditions. For units handling total gas flow rates lower than 20,000 scfm the installation costs are minimal, amounting normally to only utility tie-ins (electrical and, if necessary, combustion or dilution air). The installation costs for these smaller incinerators would be 20 to 25% of the PEC. Smaller units may be installed on the roofs of manufacturing buildings rather than at ground level. In such cases the installation factors could be as high as (or higher than) the factors shown in Table 2.10, even though the units would be “packaged.”

Table 2.10: Capital Cost Factors for Thermal and Catalytic oxidizers (Vatavuk, 1980)

Cost Item	Factor
Direct Costs	
Purchased equipment costs	
Incinerator (EC) + auxiliary equipment[a]	As estimated, A
Instrumentation[b]	0.10A
Sales taxes	0.03A
Freight	0.05A
Purchased equipment cost, PEC	B=1.18A
Direct installation costs	
Foundations & supports	0.08B
Handling & erection	0.14B
Electrical	0.04B
Piping	0.02B
Insulation for ductwork[c]	0.01B
Painting	0.01B
Direct installation costs	0.3B
Site preparation	As required, SP
Buildings	As required, Bldg.
Total Direct Costs, DC	1.30 B + SP + Bldg.

Cost Item	Factor
Indirect Costs (installation)	
Engineering	0.10B
Construction and field expenses	0.05B
Contractor fees	0.10B
Start-up	0.02B
Performance test	0.01B
Total Indirect Costs, IC	0.28B
Contingency Costs, C[d]	CF(DC + IC)
Total Capital Investment=DC+IC+C	1.58 B + SP + Bldg.+ C

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

[b] Instrumentation and controls often furnished with the incinerator, and those often included in the EC.

[c] If ductwork dimensions have been established, cost may be estimated based on \$10 to \$12/ft^2 of surface for fluid application. (Alternatively, refer to Section 1.2 of this Manual. Fan housing and stacks may also be insulated.)

[d] The default value for the contingency factor, CF, is 0.10. However, values of between 0.05 and 0.15 may be included to account for unexpected costs associated with the fabrication and installation of the control system. More information can be found on contingency in the cost estimation chapter of this Manual.

Table 2.11: Capital Cost Factors for Thermal and Catalytic oxidizer (Vatavuk, 1980) Example Problems

Cost Item	Thermal Recuperative	Fluid-Bed Catalytic
Direct Costs		
Purchased equipment costs		
Incinerator (EC)	\$254,000	\$469,000
Auxiliary equipment	–	–
Sum = A	254,000	469,000
Instrumentation, 0.1A[a]	25,400	46,900
Sales taxes, 0.03A	7,620	14,000
Freight 0.05A	12,700	23,400
Purchased equipment cost, B	\$300,000	\$553,000
Direct installation costs		
Foundations & supports, 0.08 B	24,000	44,200
Handling & erection, 0.14 B	42,000	77,400
Electrical, 0.04 B	12,000	22,100
Piping, 0.02 B	6,000	11,000
Insulation for ductwork, 0.01B	3,000	5,500
Painting, 0.01 B	3,000	5,500
Direct installation costs	\$90,000	\$166,000

Site preparation[a]	–	–
Buildings[a]	–	
Total Direct	**$390,000**	**$719,000**
Indirect Costs (installation)		
Engineering, 0.10 B	30,000	55,300
Construction and field expenses, 0.05 B	15,000	27,600
Contractor fees, 0.10 B	30,000	55,300
Start-up, 0.02 B	6,000	11,000
Performance test, 0.01 B	3,000	5,500
Total Indirect Costs	$84,000	$155,000
Total Direct and Indirect Costs	$474,000	$874,000
Contingency Cost, 0.10(DC + IC)	$47,000	$87,000
Total Capital Investment (rounded)	$521,000	$961,000

[a] These items are not required for this example problem.

2.5.2 Estimating Total Annual Cost

The total annual cost (TAC) is the sum of the direct and indirect annual costs. The TAC for both example systems is given in Table 2.12, along with suggested factors for calculating them.

2.5.2.1 Direct Annual Costs

Direct annual costs for incinerators include labor (operating and supervisory), maintenance (labor and materials), fuel, electricity, and (in catalytic units) replacement catalyst. For thermal and catalytic units, the fuel usage rate is calculated as shown in Sections 2.3.2 and 2.3.3, respectively where natural gas (methane) is assumed to be the fuel. (Other fuels could be used for thermal units.)

The electricity costs are primarily associated with the fan needed to move the gas through the incinerator. The power (in kilowatts) needed to move a given inlet volumetric flow rate of air (Q_{wi} per Sections 2.2.3 and 2.3.3) at a total flange-to-flange pressure drop of ΔP inches of water and combined motor/fan efficiency, $\mathcal{E}$, is adapted from Equation 2.7, as follows:

$$Power_{fan} = \frac{1.17 \times 10^{-4}\ Q_{w_i}\, \Delta P}{\varepsilon}$$

Fan efficiencies vary from 40 to 80 percent (Green, 1999) while motor efficiencies are usually 90 percent. The total pressure drop across an incinerator system depends on the number and types of equipment elements included in the system and on design considerations. The estimation of actual pressure drop requirements involves complex calculations based on the specific system's waste gas and flue gas conditions and equipment used. For the purposes of this section, however, the approximate values shown in Table 2.13 can be used.

For the example cases, we will assume 8,000 hours per year operation and a 60% efficiency for the fan and motor together. Using pressure drops of 4 and 8 inches of water, respectively, for the thermal and fluid-bed catalytic oxidizers,[33] and adding the pressure drop of 15 inches of water for 70% heat recovery, the fan power requirements can be calculated as follows:

Thermal oxidizer[34]

$$Power_{fan} = \frac{1.17 \times 10^{-4}\ (20{,}900\ acfm)(19\ inches\ water)}{0.60} = 77.4kW$$

Catalytic oxidizer

$$Power_{fan} = \frac{1.17 \times 10^{-4}\ (20{,}900\ acfm)(23\ inches\ water)}{0.60} = 93.7kW$$

The annual electricity costs would be the products of these usages, the annual operating hours, and the electricity cost ($/kWh), or:

$$\begin{aligned}\text{Electricity Cost (Thermal)} &= 77.4\text{ kW} \times 8{,}000\text{ hours/yr} \times \$0.0689/\text{kWh}\\ &= \$42{,}700\text{ per yr}\\ \text{Electricity Cost (Catalytic)} &= 93.7\text{ kW} \times 8{,}000\text{ hours/yr} \times \$0.0689/\text{kWh}\\ &= \$51{,}600\text{ per yr}\end{aligned}$$

The catalyst replacement costs and scheduling are highly variable and depend on the nature of the catalyst, the amount of “poisons” and particulates in the gas stream (including the auxiliary fuel), the temperature history of the catalyst, and the design of the unit. It is impossible to predict the costs in a general sense. However, noble metal monolith catalysts operating on pure hydrocarbon gases in air will last longer than fluid-bed base metal catalysts operating on chlorinated hydrocarbons in air. Noble metal catalysts are also more expensive than base metal oxide catalysts. The catalyst life for many field units is from 1 to 4 years. The cost, in 2014 dollars, of the replacement catalyst must be obtained from the vendor, but it may be estimated at $3,000/ft^3 for noble metal catalysts and $650/ft^3 for base metal oxide catalysts. For the example case, the catalyst is a base metal oxide because the waste gas contains a chlorinated compound. We will assume a four-year catalyst life. Knowing that the catalyst volume is 39 ft^3 and using a

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[33] A fluid-bed catalytic oxidizer is used because the waste gas contains a chlorinated compound which would poison the catalyst in a fixed-bed incinerator.

[34] Computed from inlet waste-gas flow rate (20,000 scfm) at preheater inlet temperature (100°F).

catalyst cost of \$650/ft^3 and the future worth factor (FWF). The FWF is used because the annual catalyst replacement cost is accrued starting in the first year of operation, while catalyst replacement purchases occur every few years. To account for the time value of money, the FWF amortizes the catalyst cost over the years preceding the actual catalyst purchase. Because the money is allocated in advance of the purchase, the sum of the annual catalyst replacement costs is less than the purchase price of the catalyst. The FWF for the catalyst replacement is calculated as follows:

$$FWF = i\left[\frac{1}{(1+i)^{y}-1}\right] = 0.0425 \times \frac{1}{(1+0.0425)^{4}-1} = 0.2346$$

Where i is the interest rate and y is the life of the catalyst. The annual expense for catalyst replacement is

$$Catalyst\ Replacement = 1.08 \ \times 650\$/cf \times 39\ cf \ \times 0.2346 = \$6{,}423$$

(The “1.08” factor covers the freight and sales tax for the replacement catalyst.)

To calculate the fuel or electricity annual cost, multiply the fuel usage rate (scfm) or the electricity usage rate (kW) by the total hours per year of operation (e.g., 333 d/yr x 24 h/d = 8,000 h/yr) and by the appropriate unit cost (e.g., \$/scfm for fuel and \$/kWh for electricity).

For the example cases, the fuel costs can be calculated from the fuel usage rates and the natural gas unit cost of \$0.00384/scf. For the thermal oxidizer example, the annual fuel cost is calculated as follows:

$$Annual\ Fuel = 0.00384\ \$/scf \times 167\ scf/min \times 60 min/hr \times 8{,}000 hr/yea$$

$$Cost, Thermal = \$308{,}000\ per\ year$$

For the catalytic oxidizer example. The annual fuel cost is found similarly:

$$Annual\ Fuel\ Cost, Catalytic = \$73{,}400\ per\ year$$

Operating and maintenance labor are estimated as 0.5 hours per 8-hour shift each, supervisory labor at 15% of operating labor, and maintenance material as 100% of maintenance labor.

Table 2.12: Annual Costs for Thermal and Catalytic oxidizers - Example Problem

Cost Item	Suggested Factor	Unit Cost[a]	Thermal	Fluid-Bed Catalyst
Direct Annual Costs[b], DC				
Operating Labor				
Operator[d]	0.5 hr/shift	$26.70/hr	13,350	13,350
Supervisor	15% of operator	—	2,000	2,000
Operating Materials	—			
Maintenance				
Labor[d]	0.5 hr/shift	$27.25/hr.	13,625	13,625
Materials	100% of maintenance labor	—	13,625	13,625
Catalyst replacement	100% of catalyst replaced	\$650/ft^{3} for metal oxide	0	6,423
Utilities				
Natural Gas[e]	—	\$3.84/kft^{3}	308,000	73,400
Electricity[f]	—	$0.0689/kWh	42,700	51,600
Total DC			$393,000	$174,000
Indirect Annual Cost, IC				
Overhead	60% of sum of operating, supervisor, & maintenance labor & maintenance materials	—	25,600	29,300
Administrative Charges	2% TCI	—	10,400	19,200
Property Taxes	1% TCI	—	5,210	9,610
Insurance	1% TCI	—	5,210	9,610
Capital recovery[c]	CRF [TCI - 1.08 (cat. Cost)]	—	49,200	90,100
			$85,580	**$139,000**
Total Annual Cost (rounded)			**$479,000**	**$313,000**

[a] 2015 dollars

[b] Assumes 8,000 hr/yr

[c] The capital recovery cost factor, CRF, is a function of the equipment life (typically, 20 years) and the opportunity cost of the capital (i.e., interest rate). For a 20-year equipment life and a 4.25% interest rate, CRF = 0.0752.

[d] Bureau of Labor Statistics, *May 2015 National Occupational Employment and Wage Estimates – United States*, May 2015 (*http://www.bls.gov/oes/current/oes_nat.htm*). Hourly rates for operators based on data for plant and System Operators – other (51-8099). Hourly rates for maintenance workers based on electrical and electronics commercial and industrial equipment repairers (49-2094).

[e] Annual average price paid for natural gas by industrial facilities in 2015 from the U.S. Energy Information Administration. Available at http://www.eia.gov/dnav/ng/hist/n3035us3A.htm.

[f] Average annual electricity cost for industrial plants is based on 2015 price data compiled by the U.S. Energy Information Administration from data reported on Form EIA-861 and 861S, (http://www.eia.gov/electricity/data.cfm#sales).

Table 2.13: Typical Pressure Drop Across Selected Equipment

Equipment Type	Energy	ΔP, in. H_2O
Thermal Oxidizers	0	4
Catalytic Oxidizer (Fixed-bed)	0	6
Catalytic Oxidizer (Fluid-bed)	0	6-10
Heat Exchangers	35	4
Heat Exchangers	50	8
Heat Exchangers	70	15

2.5.2.2 Indirect Annual Costs

The indirect (fixed) annual costs include capital recovery, overhead, and property taxes, insurance, and administrative (G&A) charges. The last three of these can be estimated at 1%, 1%, and 2% of the total capital investment, respectively. The system capital recovery cost is based on an estimated 20-year equipment life. This estimate of oxidizer equipment life is consistent with information available to EPA and is consistent with a statement by Anguil Environmental, Inc., a large vendor for incinerators and oxidizers (Chemical Engineering, 2015). (See Section 1.2 for the discussion of the capital recovery cost and the variables that determine it.) The system capital recovery cost is the product of the system capital recovery factor (CRF) and the total capital investment (TCI) less the purchased cost of the catalyst (C_{cat} x 1.08 where the 1.08 is for freight and sales tax). These values calculated for the example cases are given in Table 2.12.

2.5.3 Cost Comparison for Example Case

The example VOC stream defined in Section 2.4.1 serves to illustrate some typical characteristics of thermal and catalytic systems. First, the auxiliary fuel costs for natural gas are a significant part of the total annual operating costs for both the catalytic and thermal oxidizers. As expected, the total annual costs for the catalytic system’s auxiliary fuel costs are significantly lower than those of the thermal unit. The auxiliary fuel costs account for approximately 64 percent of the total annual cost for the thermal oxidizer and less than 24 percent of the total annual cost for the catalytic oxidizer. In this example, the disparity of $234,600 in the auxiliary fuel cost is enough to offset the higher capital costs of the catalytic oxidizer over the assumed 20-year lifetime of the units. However, the direct annual costs are highly sensitive to changes in the price of natural gas. Lower natural gas prices, such as those experienced in 2016, can significantly reduce the total annual costs for both types of oxidizers and narrow the disparity between the annual costs for operating the two units. Conversely, higher natural gas prices increase the total annual costs and widen the disparity in the total annual costs such that the higher capital costs of the catalytic oxidizer are easily offset by the higher annual operating costs of the thermal oxidizer. For example, if we use the comparatively low natural gas price of $2.96/Mscf instead of the $3.84/Mscf used in the example above, then the difference in the annual auxiliary fuel costs narrows such that the total annual costs for the catalytic oxidizer is only about $180,000 less than the total annual costs for the thermal oxidizer. However, if we use the 10-year high of

$9.65/Mscf[35] for the natural gas price at industrial facilities, the total annual costs for the catalytic oxidizer would be $589,000 less than the total annual costs for the thermal oxidizer, which easily offsets the $440,000 additional capital investment required for the catalytic oxidizer.

Two other factors that should be noted in the comparison of these two systems are (1) the 99 percent level of destruction met by the thermal oxidizer may be difficult to reach by the catalytic system (this may be important in some cases), and (2) the example waste stream is of particularly low heating value (4 Btu/scf) which favors the catalytic system due to the lower auxiliary fuel requirements.

2.6 Cost Analysis for Incinerators

Solid waste incineration systems are designed to handle a specific volume and type of waste. Information needed to select/design a solid waste incinerator include:

- The total amount of solids to be processed, including the average and peak amounts.
- Characteristics of the waste material, including moisture content, percent volatile solids, heat value, and concentration of inorganics.

Capital costs of incinerators depend on the size and type of unit. In addition to the cost of the incinerator, other components of an incinerator system include:

- Solid waste storage and conveyance system to transfer waste to the incinerator;
- Ash handling and storage system;
- Air pollution controls; and
- Auxiliary fuel system (e.g., natural gas pipelines, fuel oil storage tanks).

Some incinerators operate using ambient air, while others preheat the air before it enters the incinerator. Heating the air reduces fuel costs by about 60 percent. However, the additional equipment needed to preheat the air may increase the capital costs by as much as 15 percent (EPA 2003).

Sewage Sludge Incinerators

Annual operating costs depend on fuel costs. The amount of auxiliary fuel combusted depends on the heat value of the waste. The heat of combustion for sewage sludge ranges from 18,624 to 30,364 kj/dry Kg of solids (8,000 to 13,000 Btu/dry lb) for primary wastewater solids and 11,640 to 23,280 kJ/dry kg of solids (5,000 to 10,000 Btu/dry lb) for mixtures of primary waste and activated solids. Incinerators handling anaerobically digested solids are generally more expensive to operate because of the low heat value of the solids (12,804 kJ/dry kg of solids (5,500 Btu/dry lb). The amount of auxiliary fuel consumed is also dependent on the moisture content of

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[35] Average annual natural gas price for industrial consumers in 2008 based on U.S. Energy Information Administration data available at https://www.eia.gov/dnav/ng/hist/n3035us3A.htm.

the sludge. In general, incineration of sewage sludge is most economical when the sludge is dewatered to at least 25 percent solids. Typical annual O&M costs range from $83 to $269/dry Mg ($76 to $245/dry ton (adjusted to 2002 ENR values). The adjusted 2002 O&M costs for a multiple hearth facility retrofitted with additional air pollution control equipment to comply with federal regulations (40 CFR Part 503, 40 CFR Part 62, Subpart LLL, 40 CFR Part 60, Subpart LLLL, and 40 CFR Part 60, Subpart MMMM) (EPA 2003).

Capital costs for sludge waste incinerators is difficult to obtain because few new incinerators have been constructed in recent years due to the relatively low cost of land application (estimated at $15.40/dry Mg ($14/dry ton)) (EPA 2003).

Two new sewage sludge incinerators were installed at the R.L Sutton Wastewater Reclamation facility in 2009. The system consists of 2 new fluidized bed incinerators and a dewatering system consisting of four centrifuges, cake pumps and screw feeders, sludge silos, polymer system, and controls. The total capital cost of the system was $56,890,000 (EPA 2003).

Three new fluidized bed incinerators were installed at the Mill Creek in 2010. They cost $75 million to permit and construct, and are anticipated to save about $1 million in fuel costs a year. The new incinerators burn undigested sludge, which means the incinerators can be self-fired by the thermal energy contained in the waste being burned. The old incinerator building will be demolished, and the Mill Creek digesters will be decommissioned.[36]

The rated capacity of the three new incinerators is 4 dry tons per hour each, for a total of 12 dry tons per hour. When operating continuously, they can be autogenous, or self-fueling. This feature will dramatically reduce fuel consumption and air emissions, including greenhouse gas (GHG) emissions. Due to the lower operating costs, greater operational flexibility, and lower emissions, the EPA believes that only fluidized bed incinerators are likely to be constructed in the future (ERG 2011).[37]

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[36] See http://projectgroundwork.org/sustainability/op_enviro/incinerator.html.

[37] Memorandum from Eastern Research Group, Inc. to U.S. EPA, *Revised Estimation of Impacts for New Units Constructed Within Five Years After Promulgation of the SSI NSPS,* January 2011.

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Appendix A

Properties of Selected Compounds

Table 2.14: Limits of Flammability of Combustible Organic Compounds in Air at Atmospheric Pressure, Room Temperature (Lide, 2005 and Vatavuk, 1990)

Compound	Molecular Weight (g)	LEL[a] (volume %)	UEL[b] (volume %)	Net Heat Content (Btu/scf)
Methane	16.04	5.0	15.0	44.3
Ethane	30.07	3.0	12.5	47.6
Propane	44.09	2.1	9.5	47.7
Butane	58.12	1.9	8.5	53.2
Pentane	72.15	1.4	8.0	51.0
Hexane	86.17	1.1	7.5	51.9
Octane	114.23	1.0	6.5	54.0
Nonane	128.25	0.8	2.9	54.1
Decane	142.28	0.8	5.4	52.9
Ethylene	28.05	2.7	36	39.7
Propylene	42.08	2.0	11.1	51.4
Cyclohexane	84.16	1.3	8	53.3
Benzene	78.11	1.2	7.8	45.8
Toluene	92.13	1.1	7.1	50.3

[a] Lower Explosive Limit

[b] Upper Explosive Limit

Table 2.15: Molar Heat Capabilities of Gases at Zero Pressure (Kobe, 1954)

$$C_p = a + bT + cT^2 + dT^3\ ; T\ in\ °K$$

$$C_{pm} = \frac{\int_{T_1}^{T_2} C_p dT}{(T_2 - T_1)}$$

$$C_p\ \ in\ calories/g - moles\ °K\ btu/lb - mole\ °R$$

Compound	**a**	**b x 10^2**	**c x 10^6**	**d x 10^{10}**	**Temperature Range (°K)**
Methane	4.750	1.200	0.3030	-2.630	273-1500
Ethane	1.648	4.124	-1.530	1.740	273-1500
Propane	-0.966	7.279	-3.755	7.580	273-1500
Butane	0.945	8.873	-4.380	8.360	273-1500
Pentane	1.618	10.85	-5.365	10.10	273-1500
Hexane	1.657	13.19	-6.844	13.78	273-1500
Cyclopentane	-12.957	13.087	-7.447	16.41	273-1500
Cyclohexane	-15.935	16.454	-9.203	19.27	273-1500
Benzene	-8.650	11.578	-7.540	18.54	273-1500
Toluene	-8.213	13.357	-8.230	19.20	273-1500
Nitrogen	6.903	-0.037553	0.1930	-0.6861	273-1500
Oxygen	6.085	0.3631	-0.1709	0.3133	273-1500
Air	6.713	0.04697	0.1147	-0.4696	273-1500
Carbon Dioxide	5.316	1.4285	-0.8362	1.784	273-1500

Table 2.16: Heats of Combustion of Selected Gaseous Organic Compounds (Green, 1999)

Compound	Molecular Weight	Heat of Combustion	
		(cal/g)	(Btu per lb)
Methane	16.043	11,957	21,523
Ethane	30.07	11,355	20,439
Propane	44.097	11,073	19,932
Butane	58.123	10,927	19,669
Pentane	72.15	10,749	19,348
Hexane	86.177	10,692	19,245
Octane	114.231	10,617	19,110
Nonane	128.258	10,593	19,068
Decane	142.285	10,573	19,031
Ethylene	28.054	11,273	20,291
Propylene	42.081	10,937	19,688
Acetylene	26.038	11,537	20,767
Cyclohexane	84.161	10,382	18,689
Benzene	78.114	9,595	17,271
Toluene	92.141	9,685	17,434

Appendix B

Design Procedure for Non-Recuperative Thermal oxidizers

Not all thermal oxidizers are equipped with recuperative heat exchangers to transfer energy from the flue gas stream to the incoming waste gas stream. These non-recuperative units use other mechanisms to recovery flue gas energy. One of these types is the *regenerative oxidizer*. As discussed in a subsection of Section 2.1 entitled Regenerative oxidizers, a regenerative oxidizer accomplishes energy recovery by conveying the flue gas through a ceramic bed which captures a portion of the stream's enthalpy. After a switching mechanism is engaged, the incoming waste gas passes through this hot bed and is warmed to its ignition temperature. This process is illustrated in Figure 2.2.

While we can determine the stream inlet and outlet temperatures for a recuperative heat exchanger fairly accurately, we cannot always do so for a regenerative oxidizer bed. For one thing, these beds do not behave like typical heat exchangers. The bed temperature profiles are often difficult to predict. More importantly, because regenerative oxidizers do not operate at steady state conditions, the temperatures within the beds and many other parts of the unit vary with time. For that reason, it is more convenient to view the entire regenerative oxidizer as a "black box" into which waste gas and auxiliary fuel flow and from which flue gas emanates. Around this black box we may make mass and energy balances. In this way, we need not make any assumptions about what occurs inside the incinerator regarding temperatures, flow rates, or other stream parameters.

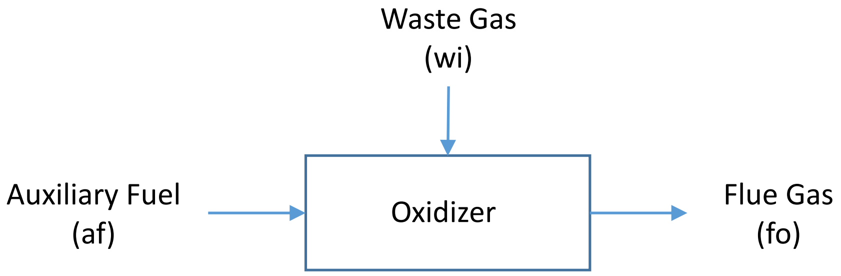
However, to determine the auxiliary fuel requirement for regenerative oxidizers via the procedure shown in this appendix we have to make two key assumptions, viz.: (1) the temperatures and flow rates of all streams entering and leaving the incinerator are at steady state and (2) the combustion temperature (and by inference, the heat loss fraction) are constant as well. The other assumptions will be addressed in the following design steps:

Steps 1 to 4: These are the same as those for thermal recuperative and catalytic oxidizers. (See Section 2.4.1.)

Step 5t - Establish the incinerator operating temperature: Because their designs are more resistant to thermal stresses and because they can achieve very high heat recoveries, regenerative oxidizers are usually operated at higher temperatures than recuperative units. Consequently, higher VOC destruction efficiencies are achieved. Operating temperatures of 1800 to 2000°F are typical.

Step 6t - Calculate the waste gas temperature at the exit of the preheater: As explained above, regenerative oxidizers do not employ preheaters. The preheating is done by and within the ceramic beds. Moreover, because the mass and energy balances are made around the entire unit, we do not need to know the temperature of the preheated waste gas to calculate the auxiliary fuel requirement.

Step 7t - Calculate the auxiliary fuel requirement, Q_{af}: Because a regenerative oxidizer recovers nearly all of the energy from the combustion (flue) gas, its auxiliary fuel requirement is usually lower than that for a recuperative oxidizer. However, as discussed above, this fuel requirement is determined via mass and energy balances taken around the entire unit, not just the combustion chamber. Consider the following diagram:



Taking mass and energy balances around the incinerator, we obtain:

Mass balance:

$$Mass\ in = Mass\ out$$

$$Mass\ fuel + Mass\ waste\ gas = Mass\ flue\ gas$$

$$\rho_{af}Q_{af} + \rho_{wi}Q_{wi} = \rho_{fo}Q_{fo} \tag{2.43}$$

Energy balance:

Next, we take an energy balance around the incinerator unit:

$$Energy\ in - Energy\ out + Energy\ generated = 0$$

The terms of the energy balance equation are the inlet waste gas and outlet flue gas enthalpies (H_{wi} and H_{fo}, respectively), the energy loss (H_L), and the waste gas VOC and fuel (natural gas) heat contents (H_{cwi} and H_{caf}, in turn):

$$H_{wi} - \left(H_{fo} + H_L\right) + \left(H_{cwi} + H_{caf}\right) = 0 \tag{2.44}$$

The variables comprising each of the terms in this energy balance equation are listed in Table 2.7. They are:

$$H_{wi} = \rho_{wi}Q_{wi}C_{pmwi}\left(T_{wi} - T_{ref}\right)$$

$$H_{fo} = \rho_{fo}Q_{fo}C_{pmfo}\left(T_{fo} - T_{ref}\right)$$

$$H_L = \eta\rho_{fi}Q_{fi}C_{pmfi}\left(T_{fi} - T_{ref}\right)$$

$$H_{cwi} = \rho_{wi}Q_{wi}(-\Delta h_{cwi})$$

$$H_{caf} = \rho_{af}Q_{af}\left(-\Delta h_{caf}\right)$$

where

η = energy loss from combustion chamber (fractional)

T_{fi} = combustion temperature (°F)

We next substitute these variables into Equation 2.44 and solve for the fuel mass rate ($\rho_{af}Q_{af}$). When doing so, we make the following assumptions:

- The streams flowing to and from the incinerator are at steady state conditions.
- The auxiliary air requirements are zero.
- The ambient, reference, and fuel inlet temperatures are equal (77°F). (This assumption results in the inlet fuel stream having a zero enthalpy.)
- The heat capacities of the gas streams to and from the unit are approximately the same, regardless of composition.
- The mean heat capacities of the streams above the reference temperature (77°F) are approximately equal, regardless of temperature. Further, the mean heat capacity of the waste gas/flue gas stream entering/leaving the incinerator is evaluated at the average of the inlet (T_{wi}) and combustion (T_{fi}) temperatures. That is, $C_{pmwi} = C_{pmfi} = C_{pmfo} = C_{pm}$.

When we do all this, we get the following expression:

$$\rho_{wi}Q_{wi}C_{pm}\left(T_{wi}-T_{ref}\right)-\left[\rho_{fo}Q_{fo}C_{pm}\left(T_{fo}-T_{ref}\right)+\eta\rho_{fi}Q_{fi}C_{pm}\left(T_{fi}-T_{ref}\right)\right]+$$

$$(Energy\ in) \qquad (Energy\ out)$$

$$\left[\rho_{wi}Q_{wi}\left(-\Delta h_{cwi}\right)+\rho_{af}Q_{af}\left(-\Delta h_{caf}\right)\right]=0$$

$$(Energy\ generated)$$

Substitution for $\rho_{fo}Q_{fo}$ per Equation 2.43 above yields:

$$\left[\rho_{wi}Q_{wi}C_{pm}\left(T_{wi}-T_{ref}\right)\right]-\left[\eta C_{pm}\left(\rho_{af}Q_{af}+\rho_{wi}Q_{wi}\right)\left(T_{fi}-T_{ref}\right)+\right.$$
$$\left.C_{pm}\left(\rho_{af}Q_{af}+\rho_{wi}Q_{wi}\right)\left(T_{fo}-T_{ref}\right)+\rho_{wi}Q_{wi}\left(-\Delta h_{cwi}\right)+\rho_{af}Q_{af}\left(-\Delta h_{caf}\right)\right]=0$$

Finally, solving for $P_{af}Q_{af}$ the auxiliary fuel mass rate (lb/min) yields:

$$\rho_{wi}Q_{wi}\left\{C_{pm}\left[\eta\left(T_{fi}-T_{ref}\right)+\left(T_{fo}-T_{wi}\right)\right]-\left(-\Delta h_{cwi}\right)\right\}\rho_{af}\ Q_{af}$$
$$=\left\{\left(-\Delta h_{caf}\right)-C_{pm}\left[\eta\left(T_{fi}-T_{ref}\right)+\left(T_{fo}-T_{ref}\right)\right]\right\} \tag{2.45}$$

Equation 2.45 provides the auxiliary fuel requirement for any type of thermal oxidizer, as it is independent of any intermediate variables, such as the temperature of the preheated waste gas. Clearly, this equation can be used with regenerative oxidizers, as long as the above-stated assumptions hold.

The heat loss fraction (η) will vary according to the incinerator type, how the incinerator components are configured in the unit, the construction materials, the type and amount of insulation, and other factors. For instance, for recuperative oxidizers, η is approximately 0.10. The heat loss fraction for regenerative oxidizers is considerably lower, however. There are two reasons for this. First, the components of a regenerative oxidizer—combustion chamber, ceramic beds, etc.—are housed in a single enclosure, while in a recuperative oxidizer the combustion chamber, heat exchanger, and interconnecting ductwork are housed separately, thus offering more heat transfer area. Second, because regenerative units are lined with ceramic, they are better insulated than recuperative oxidizers.

To gain an estimate of this heat loss fraction, we contacted two regenerative oxidizer vendors. [Vatavuk 1992a and 1992b] Based on the heat loss data that they supplied, we calculated values ranging from 0.002 to 0.015 (0.2 to 1.5%). These values varied according to the incinerator configuration (vertical or horizontal), the waste gas flow rate, the ambient temperature, and the wind speed.

Step 8t - Verify that the auxiliary fuel requirement is sufficient to stabilize the burner flame: As explained in Section 2.4.2, only a small amount (< 5% of the total energy input) is needed to stabilize the burner flame. With recuperative oxidizers, the auxiliary fuel requirement is usually much larger than the burner stabilization requirement, so that this constraint rarely comes into play. With regenerative oxidizers, however, the auxiliary fuel requirement may be as low as or lower than the fuel needed to stabilize the burner. Therefore, it is important to compare these two requirements. This comparison is made via Equations 2.23 and 2.24. If the auxiliary fuel is less, the minimum fuel requirement would be set at 5% of the total energy input.

Step 9t - Calculate the flue gas volumetric flow rate, Q_{fi}: As with thermal recuperative oxidizers, the regenerative oxidizer flue gas flow rate is the rate used to size and cost the unit. Measured at standard conditions (1 atmosphere and 77°F), is the sum of the inlet waste gas (Q_{wi}) and fuel (Q_{af}) flow rates. But since Q_{af} for regenerative units is small compared to Q_{wi}, the waste gas and flows should be virtually identical.