Section 3

VOC Controls

Section 3.2

VOC Destruction Controls

Chapter 1

Flares

John L. Sorrels
Air Economics Group, OAQPS
U.S. Environmental Protection Agency
Research Triangle Park, NC 27711

Jeff Coburn
Kevin Bradley
David Randall
RTI International
Research Triangle Park, NC 27709

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1.1   Introduction

Flaring is a high-temperature oxidation process used to burn waste gases containing combustible components such as volatile organic compounds (VOCs), natural gas (or methane), carbon monoxide (CO), and hydrogen (H_2). The waste gases are piped to a remote, usually elevated location, and burned in an open flame in ambient air using a specially designed burner tip, auxiliary fuel, and, in some cases, assist gases like steam or air to promote mixing for nearly complete (e.g., $\geq$ 98%) destruction of the combustible components in the waste gas. Note that destruction efficiency is the percentage of a specific pollutant in the flare vent gas that is converted to a different compound (such as carbon dioxide [CO_2], carbon monoxide, or another hydrocarbon intermediate), while combustion efficiency is the percentage of hydrocarbon in the flare vent gas that is completely converted to CO_2 and water vapor. The destruction efficiency of the gases being combusted in a flare will always be greater than the combustion efficiency of these same gases in that same flare. It is generally estimated that a combustion efficiency of 96.5 percent is equivalent to a destruction efficiency of 98 percent (U.S. EPA, 2015). Gases flared from refineries, petroleum production, chemical industries, and to some extent, from coke ovens, are composed largely of inerts and low molecular weight hydrocarbons with high heating value. Blast furnace flare gases are largely composed of inert species and CO, with low heating value. Flares are also used for burning waste gases generated by sewage digesters, coal gasification, rocket engine testing, nuclear power plants with sodium/water heat exchangers, heavy water plants, and ammonia fertilizer plants. (U.S. EPA, 2015)

Combustion requires three ingredients: fuel, an oxidizing agent (typically oxygen in air), and heat (or ignition source). Flares typically operate with pilot flames to provide the ignition source, and they use ambient air as the oxidizing agent. The waste gases to be flared typically provide the fuel necessary for combustion. Combustible gases generally have an upper and lower flammability limit. The upper flammability limit (UFL) is the highest concentration of a gas in air that is capable of burning. Above this flammability limit, the fuel is too rich to burn. The lower flammability limit (LFL) is the lowest concentration of the gas in air that is capable of burning. Below the LFL, the fuel is too lean to burn. Between the LFL and UFL, combustion can occur. Completeness of combustion in a flare is governed by flame temperature, residence time and flammability of the gas in the combustion zone, turbulent mixing of the components to complete the oxidation reaction, and available oxygen for free radical formation. Combustion is complete if all hydrocarbons and CO are converted to CO_2 and water. Incomplete combustion results in some hydrocarbons or CO discharged to the flare being unaltered or converted to other organic compounds such as aldehydes or acids.

The flaring process can produce some undesirable by-products including noise, smoke, heat radiation, light, sulfur oxides (SO_x), nitrogen oxides (NO_x), CO, and can be an undesirable potential source of ignition. However, by proper design, these can be minimized.

To improve the clarity of this chapter, the following terms are defined:

Assist air means all air that intentionally is introduced prior to or at a flare tip through nozzles or other hardware conveyance for the purposes including, but not limited to, protecting the design of the flare tip, promoting turbulence for mixing or inducing air into the flame. *Assist air* does not include the surrounding ambient air.

Assist steam means all steam that intentionally is introduced prior to or at a flare tip through nozzles or other hardware conveyance for the purposes including, but not limited to, protecting the design of the flare tip, promoting turbulence for mixing or inducing air into the flame.

Auxiliary fuel means all gas introduced to the flare in order to improve the heat content of combustion zone gas.

Combustion zone gas means all gases and vapors found just after a flare tip. This gas includes all flare vent gas, all assist steam, and that portion of assist air, if any, that is intentionally introduced to the flare vent gas or center steam prior to the flare tip.

Flare purge gas means gas introduced between a flare header's water seal and the flare tip to prevent oxygen infiltration (backflow) into the flare tip. For a flare with no water seal, the function of flare purge gas is performed by flare sweep gas.

Flare sweep gas means the gas intentionally introduced into the flare header system to maintain a constant flow of gas through the flare header to prevent oxygen buildup in the flare header and, for a flare without a flare gas recovery system, to prevent oxygen infiltration (backflow) into the flare tip.

Flare vent gas means all gas found just prior to the flare tip. This gas includes all flare waste gas, that portion of flare sweep gas that is not recovered, flare purge gas and auxiliary fuel, but does not include pilot gas, assist steam or assist air.

Flare waste gas means the gas from facility operations that is directed to a flare for the purpose of disposing of the gas.

Pilot gas means gas introduced into a flare tip that provides a flame to ignite the flare vent gas.

1.1.1   Flare Types

Flares are generally categorized in two ways: (1) by the height of the flare tip (i.e., ground or elevated), and (2) by the method of enhancing mixing at the flare tip (i.e., steam-assisted, air-assisted, pressure-assisted, or non-assisted). Elevating the flare can prevent potentially dangerous conditions at ground level where the open flame (i.e., an ignition source) is located near a process unit. Further, the products of combustion can be dispersed above working areas to reduce the effects of noise, heat, smoke, and objectionable odors.

In most flares, combustion occurs by means of a diffusion flame. A diffusion flame is one in which air diffuses across the boundary of the fuel/combustion product stream toward the center of the fuel flow, forming the envelope of a combustible gas mixture around a core of fuel gas. This mixture, on ignition, establishes a stable flame zone around the gas core

above the burner tip. This inner gas core is heated by diffusion of hot combustion products from the flame zone[1].

Cracking can occur with the formation of small hot particles of carbon that give the flame its characteristic luminosity. If there is an oxygen deficiency and if the carbon particles are cooled to below their ignition temperature, smoking occurs. In large diffusion flames, combustion product vortices can form around burning portions of the gas and shut off the supply of oxygen. This localized instability causes flame flickering, which can be accompanied by soot formation.

As in all combustion processes, an adequate fuel and air supply and good mixing are required to achieve complete combustion and minimize smoke formation. The various flare designs differ primarily in their accomplishment of mixing.

Steam-Assisted Flares

Steam-assisted flares are typically single burner tips that are elevated above ground level for safety reasons and burn the vented gas in what is essentially a diffusion flame. They account for the majority of the flares installed and are the predominant flare type found in refineries and chemical plants (U.S. EPA, 2011; API/ANSI, 2014; Kalcevic, 1980). They are less common at oil production sites because such facilities generally do not install steam boilers.

To ensure an adequate air supply and good mixing, this type of flare system injects steam into the combustion zone to promote turbulence for mixing and to induce air into the flame. Steam-assisted flares tend to be more effective than air-assisted flares at achieving smokeless burning because high-pressure steam can supply more momentum, which enhances ambient air entrainment and air-fuel mixing (Bader, 2011). Steam-assist flares have a lower capital cost (for similarly-sized flares, where steam is available) and a wider operating range than air-assist flares. Steam-assisted flares are the focus of the chapter and will be discussed in greater detail in Sections 1.2 through 1.4.

Air-Assisted Flares

Some flares use forced air to provide the combustion air and the mixing required for smokeless operation. These flares are often built with a spider-shaped burner (with many small gas orifices) located inside but near the top of a steel cylinder that may be two or more feet in diameter. However, air-assisted flares are available as small as 2 to 3 inches in diameter and as large as 7 to 10 feet in diameter (Aereon, 2014; Zeeco, 2016). Assist air is provided by a fan in the bottom of the flare that directs air through an annulus or tubes within the flare stack to the flare tip to improve mixing and reduce soot (smoke) formation. The amount of combustion air can be varied by varying the fan speed. The principal advantage of the air-assisted flares is that they can be used where steam is not available. One disadvantage

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[1] Flares should not be confused with incinerators or oxidizers. An incinerator or oxidizer consists of a closed chamber in which the combustion takes place, providing more control over the combustion. For more information on incinerators and oxidizers, please review the Incinerators and Oxidizers chapter in the EPA Air Pollution Control Cost Manual.

of air-assisted flares is that they require electricity to power the blower/fan to provide the assist air.

Non-Assisted Flares

The non-assisted flare is just a flare tip without any auxiliary provision for enhancing the mixing of air into its flame. Its use is limited essentially to gas streams that have a low heat content and a low carbon/hydrogen ratio that burn readily without producing smoke (Shore, 1990). These streams require less air for complete combustion, have lower combustion temperatures that minimize cracking reactions, and are more resistant to cracking. Typically, high-pressure (15 pounds per square inch gauge (psig) or more) waste streams do not require any supplemental assist medium. (Bader, 2011)

Pressure-Assisted (and Multi-Point Pressure-Assisted) Flares

Pressure-assisted flares use the vent stream pressure to promote mixing at the burner tip. Several vendors now market proprietary, high pressure drop burner tip designs. If sufficient vent stream pressure is available, these flares can be applied to streams previously requiring steam or air assist for smokeless operation. Pressure-assisted flares may use burner arrangements that are either elevated or at ground level and typically use a "multi-point" design. Multi-point pressure-assisted flare designs have multiple burner heads which can be staged to operate based on the quantity of gas being released. The size, design, number, and group arrangement of the burner heads depend on the vent gas characteristics. Elevated multi-point flares are commonly used for off-shore oil and gas platforms. Ground-level multi-point flares are used in industrial applications, often as emergency release flares secondary to a steam- or air-assisted flare. Ground-level pressure-assisted flares are typically located in a remote area of the plant where there is plenty of space available and are surrounded by a radiant heat fence primarily for worker safety.

Other Flare Type Designations

In addition to designating flares by the method used to enhance mixing (assist type), flares may be classified by the height of the flare tip (i.e., ground or elevated), whether the flame is enclosed or not, whether there is a single or multi-point flare tip, and whether the flare is designed for permanent or temporary/portable installation. While each of these flare type designations will impact the design of the flare, these designations may be considered secondary to the assist type.

Enclosed ground flares have burner heads enclosed inside a shell that is internally insulated or shielded. This shell reduces noise, luminosity, and heat radiation and provides wind protection, which makes enclosed ground flares less susceptible to poor performance that can occur from open-flame flares during high winds. Enclosed ground flares are typically pressure-assisted or non-assisted flares. For some designs, the height of the shell must be adequate for creating enough draft to supply sufficient air for smokeless combustion and for dispersion of the thermal plume. A primary difference between an enclosed ground flare and a combustor is that an enclosed ground flare does not have a direct method to control the volume of air introduced in the combustion zone beyond the fixed stack height (i.e., no direct air supply or louvers to limit air supply within the flare enclosure). Enclosed ground flares always have the flare burners close to ground level and generally have less capacity than open

flares. They are commonly used to combust continuous, constant flow vent streams, although reliable and efficient operation can be attained over a wide range of design capacity. Enclosed ground flares are commonly found at landfills, anaerobic wastewater treatment plants and other remote facilities.

Temporary/mobile flares may be used in a variety of applications to control emissions from singular or limited events. Temporary flares are commonly used in the oil and gas industry during well completions and at industrial plants during specific maintenance activities or startup and shutdowns. Temporary flares are commonly trailer- or skid-mounted and may come with a knock-out drum as part of the mobile package.

1.1.2 Applicability

Flares can be used to control almost any VOC stream and can handle fluctuations in VOC concentration, flow rate, heating value, and inerts content. Flaring is appropriate for continuous, batch, and variable flow waste gas stream applications. The majority of chemical plants and refineries have existing flare systems designed to relieve emergency process upsets that require release of large volumes of gas. While these large diameter flares are designed to handle emergency releases, they can also be used to control vent streams from various process operations. Consideration of waste gas stream flow rate and available pressure must be given when considering tying in to an existing flare. Normally, emergency relief flare systems are operated at a small percentage of capacity and at negligible pressure. To consider the effect of controlling an additional vent stream, the maximum gas velocity, system pressure, and ground level heat radiation during an emergency release must be evaluated. If the vent stream pressure from the emission source is not sufficient to overcome the maximum flare system pressure during an emergency release event, then the safety implications of stopping the waste gas flow during an emergency event must be considered. If the pressure of the waste gas is sufficient to overcome the maximum pressure of the flare system, consideration must also be made of the impact of the ability of other vent streams to release to the flare when needed due to the added flow and pressure incurred by adding the new waste gas stream. If adding the waste gas stream causes the maximum velocity limits or ground level heat radiation limits to be exceeded or if it causes the flow of any vent stream discharging to the existing flare to be stopped during an emergency, then the addition of the waste gas stream to the existing flare system is not viable.

Many flare systems are currently operated in conjunction with baseload gas recovery systems. These systems recover and compress the waste VOC for use as a feedstock in other processes or as fuel. When baseload gas recovery systems are applied, the flare is used in a backup capacity and for emergency releases. Depending on the quantity of usable VOC that can be recovered, there can be a considerable economic advantage over operation of the flare system alone.

Streams containing high concentrations of halogenated or sulfur-containing compounds are not usually flared due to corrosion of the flare tip, formation of secondary pollutants (such as SO_2), and limitations on flaring these compounds in some EPA regulations. Some halogenated or sulfur-containing compounds can be removed from the waste gas stream using a halogen scrubber (to remove certain halogenated compounds) or amine scrubber (to remove hydrogen sulfide) prior to being sent to the flare. If these pollutants cannot be

removed prior to combustion, then thermal incineration followed by scrubbing to remove the acid gases would be the preferred method of control (U.S. EPA, 1993).

1.1.3   Performance

This section discusses the parameters that affect flare VOC destruction efficiency and presents the specifications that must be followed when flares are used to comply with EPA air emission standards.

1.1.3.1   <u>Factors Affecting Efficiency</u>

The major factors affecting flare combustion efficiency are the flammability, auto-ignition temperature, net heating value, density, and flame zone mixing of the gases in the combustion zone of the flare.

The flammability limits of the flared gases influence ignition stability and flame extinction. The flammability limits are defined as the stoichiometric composition limits (maximum and minimum) of an oxygen-fuel mixture that will burn indefinitely at given conditions of temperature and pressure without further ignition. In other words, gases must be within their flammability limits to burn. When flammability limits are narrow, the interior of the flame may have insufficient air for the mixture to burn. Fuels, such as hydrogen, with wide limits of flammability, are therefore easier to combust.

For most flare waste gas streams, the net heating value also affects flame stability, emissions, and flame structure. A lower net heating value produces a cooler flame that does not favor combustion kinetics and is also more easily extinguished. The lower flame temperature also reduces buoyant forces, which reduces mixing.

The density of the flare waste gas stream also affects the structure and stability of the flame through the effect on buoyancy and mixing. By design, the velocity in many flares is very low; therefore, most of the flame structure is developed through buoyant forces as a result of combustion. Lighter gases therefore tend to burn better. In addition to burner tip design, the density also directly affects the minimum purge gas required to prevent flashback, with lighter gases requiring more purge (Shore, 1990).

Poor mixing at the flare tip is the primary cause of flare smoking when burning a given material. Streams with high carbon-to-hydrogen mole ratio (greater than 0.33), branched chain paraffins rather than normal isomers, and unsaturated hydrocarbons have a greater tendency to smoke and require better mixing for smokeless flaring (U.S. EPA, 2015). For this reason, one generic steam-to-vent gas ratio is not necessarily appropriate for all flare waste gas streams. The required steam rate is dependent on the carbon to hydrogen ratio of the flare waste gas stream. A high ratio requires more steam to prevent a smoking flare. Using too high of a steam (or air) rate can also cause poor flare performance. If the steam (or air) rate is too high, it will lower the net heating value and reduce the flammability of the gases in the combustion zone of the flare and potentially quench the flame (U.S. EPA, 2012).

The flare tip velocity can also impact flare performance. At too high an exit velocity, the flame can lift off the tip and flame out, while at too low a velocity, it can burn back into

the tip or down the sides of the stack. Also, at low flare tip velocities, the flame is more susceptible to wind shear and flame out due to crosswinds.

1.1.3.2   Flare Specifications

The current EPA requirements for flares used to comply with EPA air emission standards are specified in the General Provisions at 40 CFR 60.18 and 40 CFR 63.11 and, for flares subject to Subpart CC - NESHAP From Petroleum Refineries, in 40 CFR 63.670. The requirements include flare tip velocity limits, visible emission limits, flame presence and net heating value limits. There are some differences in the requirements for steam-assisted, air-assisted, and non-assisted flares.

Steam-Assisted Flares

The requirements for steam-assisted flares include the following:

- The flare must have an exit velocity at the flare tip of less than 60 feet per second (ft/sec) or, alternatively, less than 400 ft/sec if the flare vent gas has a net heat content $\geq$1,000 British thermal units per standard cubic feet (Btu/scf) or less than the maximum allowable flare tip velocity (V_{max}, in ft/sec) determined by the following equation if the flare vent gas stream has a net heat content >300 Btu/scf and <1,000 Btu/scf:

$$\log_{10}(V_{max}) = \frac{B_v + 1{,}212}{850} \tag{1.1}$$

where:

V_{max} = maximum allowable flare tip velocity (ft/sec)
B_v = net heating value of the vent gas (Btu/scf)

- The flare must have no visible emissions except for periods not to exceed a total of five-minutes during any two consecutive hours.

- The flare must have a flame present at all times when emissions may be vented. The presence of a pilot flame shall be monitored using a thermocouple or equivalent device.

- The net heating value of the gas being combusted in the flare must be:

  o For 40 CFR 60.18 and 40 CFR 63.11, 300 Btu/scf or greater evaluated on the flare vent gas.

  o For 40 CFR 63.670, 270 Btu/scf or greater evaluated on the combustion zone gas with an allowance to use 1,212 Btu/scf for H_2 .

In addition to these requirements, owners or operators must monitor to ensure that flares are operated and maintained in conformance with their design.

<u>Air-Assisted Flares</u>

The requirements for air-assisted flares include all of the requirements outlined above for steam-assisted flares. In addition, air-assisted flares subject to 40 CFR 63.670 must also comply with a net heating value dilution parameter of 22 Btu per square foot (Btu/ft^2), which limits the amount of assist air that can be used for a given flare vent gas flow rate, combustion zone net heating value, and flare tip diameter.

<u>Non-Assisted Flares</u>

The requirements for non-assisted flares is similar to those for steam-assisted flares except for the requirements for the net heating value of the gas being combusted. For non-assisted flares, the net heating value of the flare vent gas must be:

- o For 40 CFR 60.18 and 40 CFR 63.11, 200 Btu/scf or greater evaluated on the flare vent gas.

- o For 40 CFR 63.670, 270 Btu/scf or greater evaluated on the flare vent gas (which is the same as the combustion zone gas for a non-assisted flare) with an allowance to use 1,212 Btu/scf for hydrogen.

<u>Pressure-Assisted (and Multi-Point Pressure-Assisted) Flares</u>

Pressure-assisted flares typically require site-specific monitoring plans, or under certain circumstances, may be required to comply with site-specific alternative standards. These plans/standards generally include requirements for no visible emissions and continuous pilot flames similar to those for steam assisted flares, but do not typically have a maximum velocity limit. They also typically have a minimum flare vent gas net heating value limit: however, this limit is typically much higher than 200 or 300 Btu/scf given the operating envelope needed to achieve a stable flare flame and high destruction efficiency for these flare types.

<u>Other Flare Types</u>

Enclosed ground flares and portable flares must meet the requirements above based on the flare burner design. For example, an enclosed ground flare that uses multi-point, pressure-assisted flare burners would need to meet the requirements for pressure-assisted flares, and an enclosed ground flare that uses non-assisted flare burners would have to meet the requirements for non-assisted flares. Similarly, a non-assisted portable or temporary flare would have to meet the requirements for non-assisted flares.

1.2   Process Description

The remainder of this chapter provides flare descriptions, design and cost information focused primarily on elevated steam-assisted flares used as a VOC control device as these are the most common for industrial facilities. Flares used at remote facilities, like upstream oil and gas facilities or landfills, are often non-assisted or air-assisted. As such, some of the descriptions, design and cost information provided in this chapter may not apply to these other flare types. Additionally, the information provided in this chapter generally does not consider

the design complexity and costs associated with utility safety flares that service hundreds of sources.

The elements of an elevated steam-assisted flare generally consist of gas transport piping (also referred to as flare header or gas collection header), utilities (fuel, steam, and air), piping from the base up, knock-out drum, liquid seal, flare stack, gas seal, burner tip, pilot burners, steam jets, ignition system, and controls. Figure 1.1 is a diagram of a steam-assisted elevated smokeless flare system showing the usual components that are included.
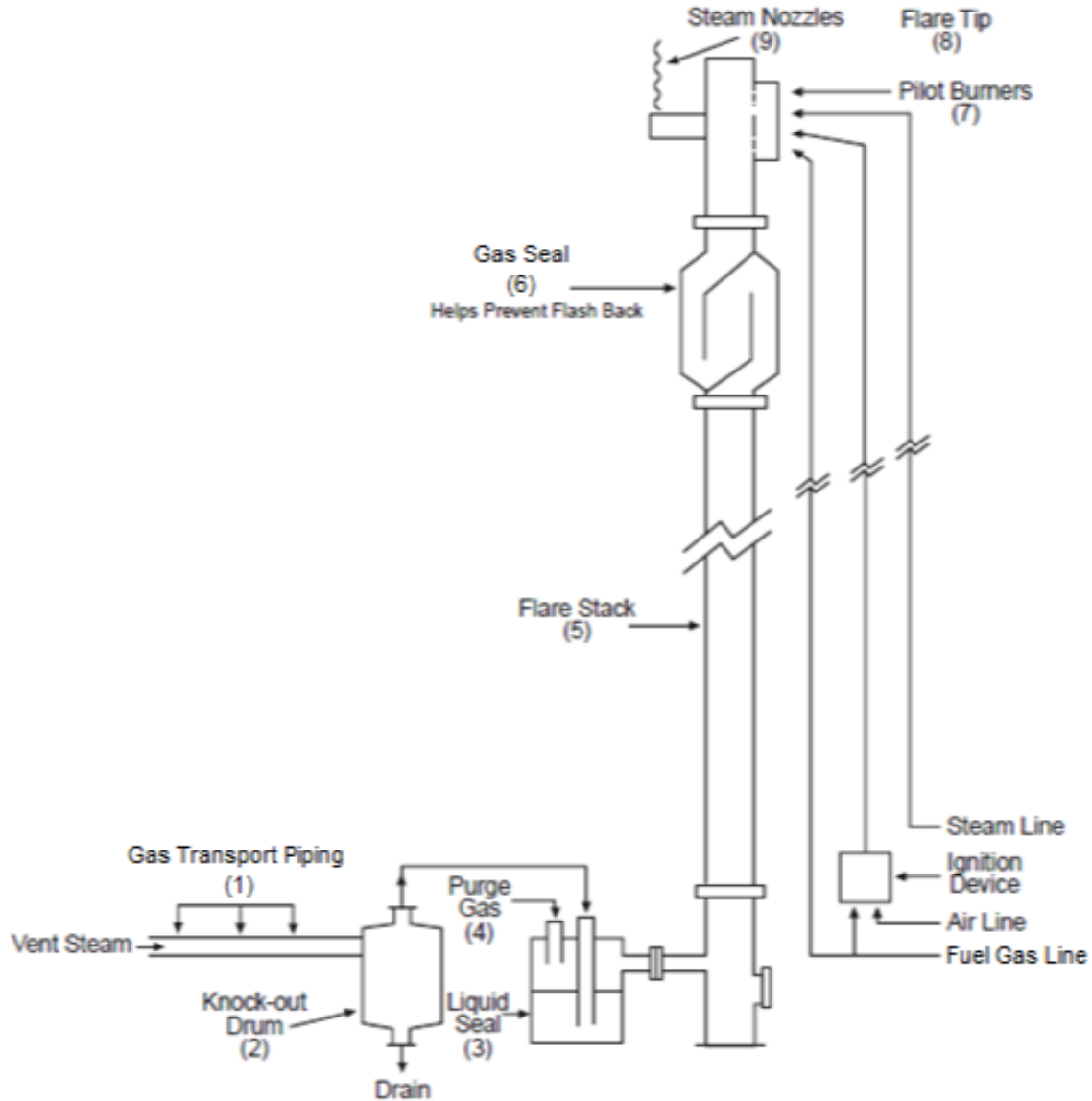
**Figure 1.1:** Steam-assisted Elevated Flare System

1.2.1 Gas Transport Piping

Process vent streams are sent from the facility release point(s) to the flare location through the gas transport piping, also referred to as the gas collection header when multiple release points are directed to the flare. The piping (generally schedule 40 carbon steel) is designed to minimize pressure drop. Ducting is not used as it is more prone to air leaks. Valving should be kept to an absolute minimum and should be "car sealed" (sealed) open. Pipe layout is designed to avoid any potential dead legs and liquid traps. The piping is equipped for sweep gas so that explosive mixtures do not occur in the flare system either on start-up or during operation.

1.2.2 Knock-out Drum

Liquids that may be in the vent stream gas or that may condense out in the collection header and transfer lines are removed by a knock-out drum. (See Figure 1.2) The knock-out, or disentrainment, drum is typically either a horizontal or vertical vessel located at or close to the base of the flare, or a vertical vessel located inside the base of the flare stack. Liquid in the vent stream can extinguish the flame or cause irregular combustion and smoking. In addition, flaring liquids can generate a spray of burning chemicals that could reach ground level and create a safety hazard. For a flare system designed to handle emergency process upsets, this drum must be sized for worst-case conditions (e.g., loss of cooling water or total unit depressuring) and is usually quite large. For a flare system devoted only to vent stream VOC control, the sizing of the drum is based primarily on the flare waste gas flow rate with consideration given to liquid entrainment.
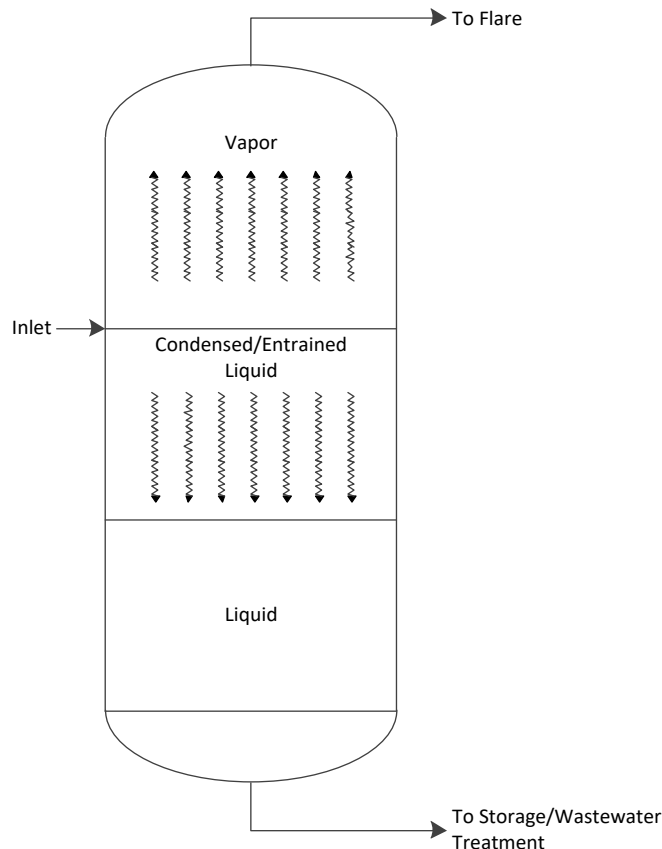
Figure 1.2: Typical Vertical Knock-out Drum

1.2.3   Liquid Seal or Flame Arrestor

Process vent streams are usually passed through a liquid seal before going to the flare stack. The liquid seal can be downstream of the knockout drum or incorporated into the same vessel. This seal prevents possible flame flashbacks, caused when air is inadvertently introduced into the flare system and the flame front pulls down into the stack. The liquid seal also serves to maintain a positive pressure on the upstream system and acts as a mechanical damper on any explosive shock wave in the flare stack. Other devices, such as flame arresters, detonation arrestors, and check valves, may sometimes replace a liquid seal or be used in conjunction with it. Sweep or purge gas (as discussed in Section 1.2.4) also helps to prevent flashback in the flare stack caused by low vent gas flow.

There are two types of flame arrestors: deflagration flame arrestors and detonation flame arrestors. Deflagration flame arrestors are designed to prevent the propagation of a subsonic combustion front. Detonation flame arrestors are designed to prevent the propagation of a sonic or supersonic combustion fronts as well as shock (pressure) fronts. Both types of arrestors dissipate heat via a metal matrix to extinguish the flame front as it attempts to pass through the device.

1.2.4 Flare Sweep or Purge Gas

The total volumetric flow to the flame must be carefully controlled to prevent low flow flashback problems or oxygen ingress into the flare or flare header system, which could lead to the formation of an explosive mixture in the flare system. This is accomplished through the use of flare sweep or purge gas. Although these two terms are often used interchangeably, flare sweep gas generally refers to the gas intentionally introduced into the flare header system to maintain a constant flow of gas through the flare header in order to prevent oxygen buildup in the flare header, and, for flares without a liquid seal, the flare sweep gas also serves to prevent oxygen infiltration (backflow) into the flare tip. For flares that have a liquid seal, the flare sweep gas is typically recovered and reused. To ensure a positive flow through all flare components, flare sweep gas injection should be at the farthest upstream point in the flare transport piping.

Flare purge gas, on the other hand, refers to gas introduced between a flare header's water seal and the flare tip to prevent oxygen infiltration (backflow) into the flare tip. Thus, flare purge gas is specific to flares that operate most of the time with a liquid seal in-place (e.g., emergency flares or flares with a flare gas recovery system). For a flare with no liquid seal, the function of flare purge gas is performed by flare sweep gas. Flare sweep or purge gas is typically natural gas or process gas, but inert gases, such as nitrogen (N_2) or CO_2, may be used to maintain a minimum required positive flow through the system. Flare sweep or purge gas is common with industrial flares.

1.2.5 Flare Stack

For safety reasons a stack is used to elevate the flare. The flare must be located so that it does not present a hazard to surrounding personnel and facilities. Elevated flares can be self-supported (free-standing), guyed, or structurally supported by a derrick. Examples of these three types of elevated flares are shown in Figures 1.3, 1.4, and 1.5 for self-supported, derrick supported, and guy-supported flares, respectively. Self-supporting flares are generally used for lower flare tower heights (30-100 feet) but can be designed for up to 250 feet. Guy towers are designed for flare stacks between 50 and 450 feet, while derrick towers are designed for flare stacks above 200 feet (McCartney, 1990; Tkatschenko, 1990; Tyler, 1990; Bozai, 1990; Parker, 1990; Sanderson, 1990).

Self-supported (free-standing) flares provide ideal structural support. However, for very high units, the costs increase rapidly. In addition, the foundation required and nature of the soil must be considered.

The guy-supported flare is the simplest of all the support methods. However, a considerable amount of land is required since the guy wires are widely spread apart. A rule of thumb for space required to erect a guy-supported flare is a circle on the ground with a radius equal to the height of the flare stack (Tkatschenko, 1990).

Derrick-supported flares can be built as high as required since the system load is spread over the derrick structure. This design provides for differential expansion between the stack, piping, and derrick. Derrick-supported flares are the most expensive design for a given flare height.

1.2.6　Gas Seal

Air may tend to flow back into a flare stack due to wind or the thermal contraction of stack gases and create an explosion potential. To limit the amount of flare sweep or purge gas required to prevent air infiltration, a gas seal is typically installed in the flare stack. There are many different designs of gas seals. These include seals referred to as a flare seal, stack seal, molecular seal, labyrinth seal, or gas barrier. The gas seal is located below the flare tip to impede the flow of air back into the flare stack or flare gas network. Some "seals" act as orifices in the top of the stack to reduce the purge gas volume needed to achieve the desired protective velocity in the flare stack when there is no waste gas flow. These are known by the names "internal gas seal", "fluidic seal", and "arrestor seal" (Shore, 1990). These gas and fluidic seals are usually proprietary in design, and their presence reduces the operating sweep or purge gas requirements.

1.2.7　Pilot Burners

A source of ignition is needed to assure the safe destruction of any hydrocarbons sent to the flare, including hydrocarbons in the purge or sweep gas. Reliable ignition is obtained by continuous pilot burners designed for stability and positioned around the outer perimeter of the flare tip. The pilot burners are ignited by an ignition source system, which can be designed for either manual or automatic actuation. Automatic systems are generally activated by a flame detection device using either a thermocouple, an infrared sensor or, more rarely, (for ground flare applications) an ultraviolet sensor (McCartney, 1990). The most common flame-detection method involves measuring the temperature at the end of a pilot tip with a standard thermocouple. The thermocouple is connected to a temperature switch or a control system that indicates pilot failure if the temperature drops below the setpoint (Bader, 2011). The EPA regulations require the presence of a continuous pilot flame.
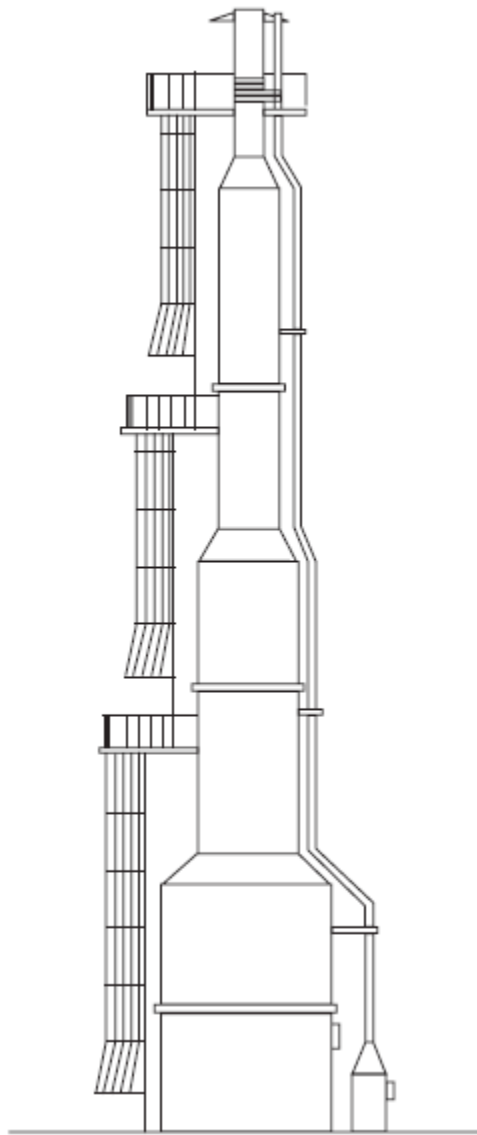
Figure 1.3: Self-supported Elevated Flare
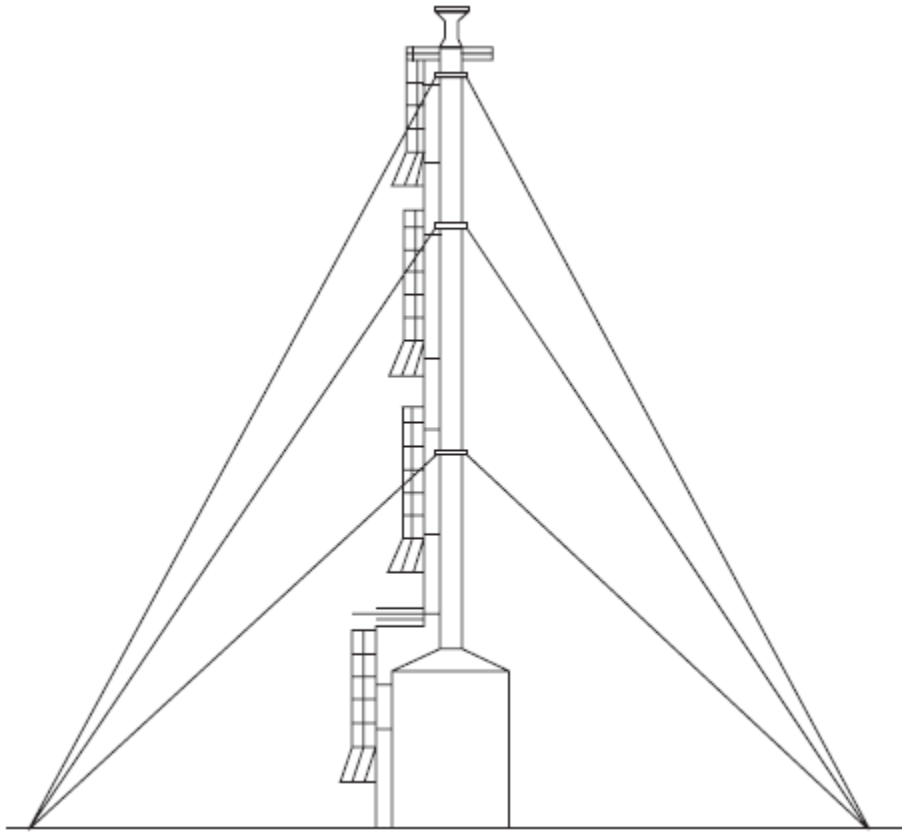
Figure 1.4: Guy-supported Elevated Flare
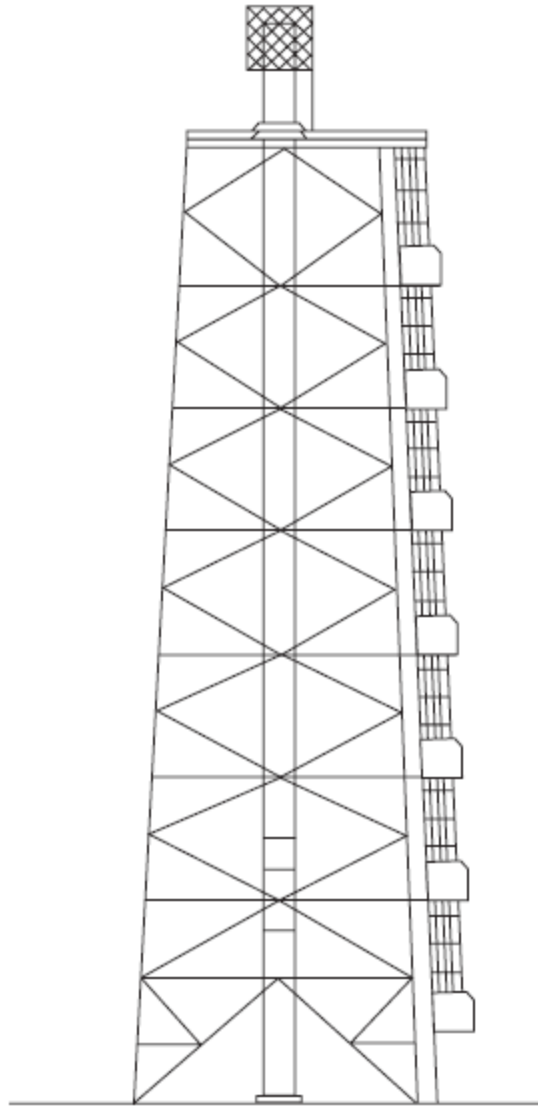
Figure 1.5: Derrick-supported Elevated Flare

1.2.8   Flare Tip

The flare tip (or burner tip), is designed to give environmentally acceptable combustion of the vent gas over the flare system's capacity range. The burner tips are normally proprietary in design. Consideration is given to flame stability, ignition reliability, and noise suppression. The maximum and minimum capacity of a flare to burn a flared gas with a stable flame (not necessarily smokeless) is a function of tip design. Flame stability can be enhanced by flame holder retention devices incorporated in the flare tip inner circumference. Burner tips with modern flame holder designs can have a stable flame over a flare gas exit velocity range of 1 to 600 ft/sec (Kalcevic, 1980). The actual maximum capacity of a flare tip is usually limited by the vent stream pressure available to overcome the system pressure drop. Elevated flare diameters are normally sized to provide vapor velocities at maximum throughput of about 50 percent of the sonic velocity of the gas subject to the constraints of 40 CFR 60.18, 40 CFR 63.11, or 40 CFR 63.670, as applicable (API/ANSI, 2014).

1.2.9   Steam Nozzles

A diffusion flame receives its combustion oxygen by diffusion of air into the flame from the surrounding atmosphere. The high volume of fuel flow in a flare may require more combustion air at a faster rate than simple gas diffusion can supply. High velocity steam injection nozzles, positioned around the outer perimeter of the flare tip, increase gas turbulence in the flame boundary zones, drawing in more combustion air and improving mixing. For the larger flares, steam can also be injected concentrically into the flare tip.

The injection of steam into a flare flame can produce other results in addition to air entrainment and turbulence. Three mechanisms in which steam reduces smoke formation have been presented (API/ANSI, 2014). Briefly, one theory suggests that steam separates the hydrocarbon molecule, thereby minimizing polymerization, and forms oxygen compounds that burn at a reduced rate and temperature not conducive to cracking and polymerization. Another theory claims that water vapor reacts with the carbon particles to form CO, CO_2, and H_2 thereby removing the carbon before it cools and forms smoke. An additional effect of the steam is to reduce the temperature in the core of the flame and suppress thermal cracking (Shore, 1990). The physical limitation on the quantity of steam that can be delivered and injected into the flare flame determines the smokeless capacity of the flare. Smokeless capacity refers to the volume of gas that can be combusted in a flare without smoke generation. The smokeless capacity is usually less than the stable flame capacity of the burner tip.

Significant disadvantages of steam usage are the increased noise and cost. Steam aggravates the flare noise problem by producing high-frequency jet noise. The jet noise can be reduced by the use of small multiple steam jets and, if necessary, by acoustical shrouding. Over-use of steam can increase the cost of operation and may lead to poor flare performance and deteriorated flare combustion efficiencies. Steam injection can be controlled manually with the operator observing the flare (either directly or on a television monitor) and adding steam as required to maintain smokeless operation. To optimize steam usage, infrared sensors are available that sense flare flame characteristics and adjust the steam flow rate automatically to maintain smokeless operation. Automatic control, based on flare gas flow and flame radiation or net heating value monitoring, gives a faster response to the need for steam and a

better adjustment of the quantity required. If a manual system is used, steam metering should be installed to significantly increase operator awareness and reduce steam consumption.

1.2.10 Controls

Flare system control can be automated or manual. Components of a flare system which can be controlled automatically include the auxiliary gas, steam injection, and the ignition system. Fuel gas consumption can be minimized by continuously measuring the vent gas flow rate and net heat content (Btu/scf) and automatically adjusting the amount of steam and auxiliary fuel to maintain the required minimum of 300 Btu/scf in the flare vent gas or 270 Btu/scf in the combustion zone gas, as applicable, for steam-assisted flares. Steam consumption should be minimized by controlling flow based on minimum amounts needed for equipment integrity (e.g., flare tip cooling) as well as flare vent gas flow rate to prevent over-steaming the flare. Steam flow can also be controlled using visual smoke and flame monitors.

Some flares are equipped with automatic pilot ignition panels that sense the presence of a flame with either visual or thermal sensors and reignite the pilots when flameouts occur. Other flares use manually triggered spark ignition based on readings from the thermocouple or other device used to monitor for the presence of a flame.

1.2.11 Flare Gas Recovery

Flare gas recovery may be used to recover flare waste and sweep gas that enters the flare header system prior to destruction in the flare so that these gases can be used as fuel for process heaters, recovered as product, or for other useful purposes. Flare gas recovery systems use one or a series of compressors to withdraw flare waste and sweep gas from the transport piping, increase the pressure of the flare vent gas stream, and transport the vent gas stream from the flare header to the facility's fuel gas system, gas treatment plant, or process heater or boiler. If the recovered flare vent gas can be recovered as product or used to offset fuel purchases, the value of the recovered flare gas can offset the costs of the flare gas recovery system, depending on the quantity and consistency of the flare vent gas flow. To apply flare gas recovery, a water seal must be used to allow recovery of the flare gas without the ingress of air at the flare tip. Typically, these water seals are located at the base of the flare stack and are an integral part of the flare equipment. For retrofit applications, a separate water seal vessel can be installed in the flare transport piping downstream of the knockout drum and near the flare.

1.3   Design Procedures

Flare design is influenced by several factors, including the availability of space, the characteristics of the flare gas (namely composition, quantity, and pressure level) and occupational concerns that limit ground-level thermal radiation intensity, luminosity, and noise. The sizing of flares requires determination of the required flare tip diameter and height. The emphasis of this section will be to size a steam-assisted elevated flare for a given application, but many of the design procedures described in this section are also applicable to non-assisted and air-assisted flares.

1.3.1 Auxiliary Fuel Requirement

The flare tip diameter is a function of the flare waste gas flow rate plus the auxiliary fuel and flare sweep or purge gas flow rate (i.e., the flare vent gas flow rate). The flare sweep or purge gas flow rate is typically small relative to the waste gas and auxiliary fuel flow rates during waste gas flow events, so it may be ignored when determining the tip diameter. The flow rate of the auxiliary fuel, if required, can be significant, and must be calculated before the tip diameter can be computed.

Some flares are provided with auxiliary fuel to combust hydrocarbon vapors when a lean flare gas stream falls below the flammability range or heating value necessary to sustain a stable flame. The amount of fuel required is calculated based on maintaining the flare vent gas net heating value (NHV) at the minimum of 300 Btu/scf or maintaining the combustion zone gas NHV at a minimum of 270 Btu/scf as required by 40 CFR 60.18, 40 CFR 63.11, or 40 CFR 63.670, as applicable. A generalized heat balance follows:

$$Q B_v + F B_f = \left(NHV_{t \arg et} \right)\left(Q + F + K_1 S \right) \tag{1.2}$$

where:

Q = flow rate of the waste gas stream (scfm)
F = flow rate of the auxiliary fuel (scfm)
B_v = heat content of the waste gas stream (Btu/scf)
B_f = heat content of the auxiliary fuel (Btu/scf); can assume B_f = 920 Btu/scf when natural gas is used as the auxiliary fuel.
NHV_{target} = net heating value operational target (Btu/scf)
K_1 = combustion zone factor (dimensionless); K_1=1 if the net heating value target is to be assessed in the combustion zone; K_1=0 if the net heating value target is to be assessed based on the flared gas (e.g., complying with 40 CFR 60.18 or 63.11).
S = flow rate of the assist steam (scfm)

To ensure compliance with the minimum NHV limits, it is reasonable to assume NHV_{target} is 10 percent higher than the minimum NHV required. When the NHV must be determined based on the combustion zone properties, the steam flow rate should first be adjusted to see if NHV_{target} can be met simply by reducing the assist steam rate. For many steam assisted flares, there is a minimum steam assist rate to prevent condensation (and potential freezing) of water in the steam lines and maintain a minimum cooling rate at the flare tip for equipment integrity purposes. If reducing the assist steam rate to its minimum value does not achieve NHV_{target}, Equation 1.3 can be used to calculate the minimum auxiliary fuel flow rate:

$$F_{\min} = \frac{Q\left(NHV_{t \arg et} - B_v \right) + NHV_{t \arg et} K_1 S_{\min}}{B_f - NHV_{t \arg et}} \tag{1.3}$$

where:

S_{min} = minimum flow rate of the assist gas needed for steam lines (scfm).
F_{min} = minimum flow rate of auxiliary fuel needed (scfm).

It is important to note that S_{min} in Equation 1.3 may also be constrained by the smokeless operation of the flare for certain vent gas flows, vent gas compositions, and flare tip diameter sizes, and that this should be considered in addition to the flare's minimum assist steam rate. Therefore, Equation 1.3 should be evaluated for a variety of expected flow conditions and the flare sized to handle the reasonable worst-case flow conditions. Note that the design procedures described in this *Control Cost Manual* are for the control of routine waste gas streams and not facility-wide flares designed for emergency releases caused be a severe malfunctions or abrupt shutdowns.

The contribution to the minimum annual auxiliary fuel requirement for each waste gas flow scenario is calculated by converting F_{min} from scfm to thousand scf (Mscf)/yr as follows:

$$F_{a,\min} = F_{\min}\left(\frac{60\,t_{af}}{1,000}\right) \tag{1.4}$$

where:

$F_{a,min}$ = contribution to the minimum annual auxiliary fuel requirement for a given waste gas flow scenario (Mscf/yr)
t_{af} = total time attributable to a given waste gas flow scenario in the year (hr/yr)
60 = conversion from minutes to hours (min/hr)
1,000 = conversion from scf to Mscf (scf/Mscf)

Most flares will use either natural gas or fuel gas from a fuel gas system as auxiliary fuel if it is needed. Automatic control of the auxiliary fuel is ideal for processes with large fluctuations in VOC compositions. These flares are used for the disposal of such streams as sulfur tail gases and ammonia waste gases, as well as any low Btu vent streams (Kalcevic, 1980).

1.3.2   Flare Tip Diameter

Flare tip diameter is generally sized on a velocity basis, although pressure drop must also be checked. Flare tip sizing for steam assisted flares used to comply with the EPA air emission standards is governed by rules defined in the Code of Federal Regulations (see 40 CFR 60.18, 40 CFR 63.11, or 40 CFR 63.670, as applicable).[2] To comply with these requirements, the maximum velocity of a steam-assisted elevated flare can either be maintained below 60 ft/sec regardless of the NHV of the flare vent gas (while meeting the minimum requirements for flare vent gas NHV) (Option 1 in Table 1.1) or below specific flow limits based on the flare vent gas NHV. For flares burning flare vent gas with a NHV between 300 Btu/scf and 1000 Btu/scf, the maximum permitted flare tip velocity can be calculated using Equation 1.1 (Option 2a in Table 1.1); for flares burning flare vent gas with a NHV of 1,000 Btu/scf or more, the flare tip velocity must remain less than 400 ft/sec (Option 2b in

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[2] The requirements in 40 CFR 60.18 also are recommended in the model rule language in the 2016 Control Techniques Guidelines (CTG) for the Oil and Natural Gas Industry (U.S. EPA, 2016).

Table 1.1). The maximum velocity limits in 40 CFR 60.18, 40 CFR 63.11, or 40 CFR 63.670 may be exceeded for certain malfunction events. However, as noted previously, this *Control Cost Manual* considers flare applications used as a control device rather than safety and malfunction applications. When applied as a control device, it is good practice to size the flare for the largest flow event that is required to meet these velocity limits, which may include startup and planned shutdown events.

Table 1.1: Maximum Permitted Velocity for a Steam-assisted Flare

Option	Net Heating Value of Vent Stream, B_v (Btu/scf)	Maximum Permitted Velocity, V_{max} (ft/sec)
1	≥ 300	< 60
2a	≥ 300 & $< 1{,}000$	$\log_{10}(V_{max}) = \dfrac{(B_v + 1{,}212)}{850}$
2b	$\geq 1{,}000$	< 400

By determining the maximum permitted flare tip velocity, V_{max} (ft/sec), and knowing the maximum total volumetric flow rate, Q_{tot} (acfm), including flare waste gas and auxiliary fuel gas, a minimum flare tip diameter, D_{min} (in), can be calculated for each flow scenario. It is standard practice to size the flare so that the design velocity of flow rate Q_{tot}, is 80 percent of the allowable V_{max}, i.e.:

$$D_{min} = 12 \sqrt{\frac{\dfrac{4}{\pi} \times \dfrac{Q_{tot}}{60}}{0.8\, V_{max}}} = 1.95 \sqrt{\frac{Q_{tot}}{V_{max}}} \tag{1.5}$$

where:

Q_{tot} = maximum total volumetric flow = maximum expected Q_{max} + F (acfm)
Q_{max} = maximum expected flow rate of the waste gas stream (acfm)
  12 = conversion from feet to inches (in/ft)
 $4/\pi$ = part of the equation for calculating the area of a circle
  60 = conversion from seconds to minutes (sec/min)
 0.8 = sizing factor

Note that the maximum D_{min} may not occur at the highest flow rate scenario. A high flow scenario of low heat content gas may yield a larger D_{min} than a higher flow rate of high heat content gas. The flare tip diameter, D, is the calculated diameter, D = maximum(D_{min}), rounded up to the next commercially available size. The minimum flare size is 1 inch; larger sizes are available in 2-inch increments from 2 to 24 inches and in 6-inch increments above 24 inches. The maximum size commercially available is 120 inches (Zeeco, 2017). Custom-

designed flare tips can be made for a specific size need, but the costs presented in this Manual are specific to these commercially available incremental flare tip sizes.

A pressure drop calculation is required at this point to ensure that the vent stream has sufficient pressure to overcome the pressure drop occurring through the flare system at maximum flow conditions. The pressure drop calculation is site specific but must take into account losses through the collection header and piping, the knock-out drum, the liquid seal or flame arrestor, the flare stack, the gas seal, and finally the flare tip. Piping size should be assumed equal to the flare tip diameter. Schedule 40 carbon steel pipe is typically used. If sufficient pressure is not available, the economics of either a larger flare system (pressure drop is inversely proportional to the pipe diameter) or a mover such as a fan or compressor (typically only applicable for non-emergency flare systems) must be weighed (refer to Section 1.3.8 for typical pressure drop relationships).

1.3.3   Flare Height

The height of a flare is determined based on the ground level limitations of thermal radiation intensity, luminosity, noise, height of surrounding structures, and the dispersion of the exhaust gases. In addition, consideration must also be given for plume dispersion in case of possible emission ignition failure. Industrial flares are normally sized for a maximum heat intensity of 1,500-2,000 Btu/hr-ft^2 when flaring at their maximum design rates (API/ANSI, 2014; Kalcevic, 1980). At this heat intensity level, workers can remain in the area of the flare for a limited period only. If, however, operating personnel are required to remain in the unit area performing their duties, the recommended design flare radiation level excluding solar radiation is 500 Btu/hr-ft^2 (API/ANSI, 2014). The intensity of solar radiation is in the range of 250-330 Btu/hr-ft^2 (API/ANSI, 2014). Flare height may also be determined by the need to safely disperse the vent gas in case of flameout. The height in these cases would be based on dispersion modeling for the particular installation conditions and is not addressed here. The minimum elevated flare height normally used is 30 feet (Shore, 1990). Equation 1.6 by Hajek and Ludwig may be used to determine the minimum distance, H, required from the center of the flare flame and a point of exposure where thermal radiation must be limited (API/ANSI, 2014).

$$H^2 = \frac{\tau \, f \, R}{4 \, \pi \, K} \tag{1.6}$$

where:

H = minimum distance (height) from flare flame and where radiation must be limited (ft)
τ = fraction of heat intensity transmitted
f = fraction of heat radiated
R = net heat release (Btu/hr)
K = maximum allowable radiation (500 Btu/hr-ft^2)

The conservative design approach used here ignores wind effects and calculates the distance assuming the center of radiation is at the base of the flame (at the flare tip), not in the

center. It is also assumed that the location where thermal radiation must be limited is at the base of the flare. Therefore, the distance, H, is equal to the required flare stack height (which is a minimum of 30 feet). The *f* factor allows for the fact that not all the heat released in a flame can be released as radiation. Heat transfer is propagated through three mechanisms: conduction, convection, and radiation. Thermal radiation may be either absorbed, reflected, or transmitted. Since the atmosphere is not a perfect vacuum, a fraction of the heat radiated is not transmitted due to atmospheric absorption (humidity, particulate matter). Table 1.2 is a summary of heat radiated from various gaseous diffusion flames (API/ANSI, 2014):

Table 1.2: Heat from Various Gaseous Diffusion Flames

Gas	Flare Tip Diameter (in)	Fraction of Heat Radiated (f)
Hydrogen		
	<1	0.10
	1.6	0.11
	3.3	0.16
	8.0	0.15
	16.0	0.17
Butane		
	<1	0.29
	1.6	0.29
	3.3	0.29
	8.0	0.28
	16.0	0.30
Methane		
	<1	0.16
	1.6	0.16
	3.3	0.15
Natural Gas		
	8.0	0.19
	16.0	0.23

In general, the fraction of heat radiated increases as the stack diameter increases. If stream-specific data are not available, a design basis of *f* = 0.3 will give a conservatively high estimate of the fraction of heat radiated for use in Equation 1.6 (resulting in a larger stack height).

The maximum expected heat release, *R*, is calculated from the combination of the maximum flare waste gas flow rate, Q_{max}, and the maximum net heating value, $B_{v,max}$, of the flare waste gas and, if needed, the auxiliary fuel flow rate and net heat content as follows:

$$R = \left(Q_{max} \times B_{v,max} + F \times B_f\right) \times 60 \qquad (1.7)$$

where:

R = maximum heat release (Btu/hr)
Q_{max} = maximum expected flow rate of the flare waste gas stream (scfm)

$$
\begin{array}{rcl}
F & = & \text{flow rate of the auxiliary fuel (scfm)} \\
B_{v,max} & = & \text{maximum expected heat content of the flare waste gas stream (Btu/scf)} \\
B_f & = & \text{heat content of the auxiliary fuel (Btu/scf)} \\
60 & = & \text{conversion of minutes to hours (min/hr)}
\end{array}
$$

In some cases, flare height may be limited based on applicable height restrictions from aviation concerns, extreme wind considerations, or other factors. If the minimum height exceeds applicable height restrictions for a given application, it may be necessary to use two flares to limit the maximum flow and thereby the maximum heat release expected from the flare.

1.3.4 Sweep or Purge Gas Requirement

The total volumetric flow to the flame must be carefully controlled to prevent low flow flashback problems and to avoid flame instability. Sweep or purge gas, typically natural gas, fuel gas, N_2, or CO_2, is used to maintain a minimum required positive flow through the system. If there is a possibility of air in the flare manifold, N_2, another inert gas, or a flammable gas must be used to prevent the formation of an explosive mixture in the flare system. To ensure a positive flow through all flare components, sweep gas injection should be at the farthest upstream point in the flare transport piping. The amount of sweep gas required is dependent on the complexity of the flare collection header system, with more sweep gas required for larger, more complex collection headers. For flares with flare gas recovery, the sweep gas is recovered so the quantity of sweep gas does not impact the annual operating costs. For flares without flare gas recovery, the sweep gas also acts as purge gas to prevent air ingress and the flare tip. For simple flare collection headers where a single emission source is controlled by a flare, the amount of sweep gas required can be estimated by the gas flow rates needed to prevent oxygen ingress at the flare tip.

The minimum continuous purge or sweep gas flow rates required to prevent oxygen ingress at the flare tip is determined by the design of the stack seals, which are usually proprietary devices. Modern labyrinth and internal gas seals are stated to require a gas velocity of 0.001 to 0.04 ft/sec (at standard conditions) (Tkatschenko, 1990; Tyler, 1990; Bozai, 1990; Parker, 1990; Sanderson, 1990). Using the conservative (or higher-end) value of 0.04 ft/sec and knowing the flare diameter (in), the annual purge gas volume, F_{pu}, can be calculated:

$$
F_{pu} = (0.04)\left(\frac{\left(\frac{\pi D^2}{4}\right)}{144}\right)\left(\frac{3{,}600}{1{,}000}\right)t_{op} = \left(7.85\times10^{-4}\right)\times t_{op}\times D^2 \tag{1.8}
$$

where:

$$
\begin{array}{rcl}
F_{pu} & = & \text{annual flare purge gas volume (Mscf/yr)} \\
D & = & \text{flare diameter (in)} \\
t_{op} & = & \text{total time flare was in operation (i.e., capable of receiving waste gas) in the} \\
& & \text{year (hr/yr); typically assume } t_{op} = 8{,}760 \text{ hr/yr.} \\
0.04 & = & \text{conservative minimum gas velocity requirement (ft/sec)} \\
\pi/4 & = & \text{part of the equation for calculating the area of a circle}
\end{array}
$$

144	=	conversion from square feet to square inches (in^2/ft^2)
3,600	=	conversion from seconds to hours (sec/hr)
1,000	=	conversion from scf to Mscf (scf/Mscf)

There is another minimum flare tip velocity for operation without burn lock or instability. This minimum velocity is dependent on both gas composition and diameter and can range from insignificant amounts on small flares to 0.5 ft/sec on greater than 60-inch diameter units (Shore, 1990).

Sweep and purge gas is also required to clear the system of air before startup, and to prevent a vacuum from pulling air back into the system after a hot gas discharge is flared (the cooling of gases within the flare system can create a vacuum). The sweep or purge gas volumes associated with these short-term uses are assumed to be minor.

1.3.5 Pilot Gas Requirement

The number of pilot burners required depends on flare size and, possibly, on flare gas composition and wind conditions. Pilot gas usage is a function of the number of pilot burners required to ensure positive ignition of the flared gas, of the design of the pilots, and of the mode of operation. The average pilot gas consumption based on an energy-efficient model is 70 scf/hr (of typical 1,000 Btu per scf gas) per pilot burner (Tkatschenko, 1990; Tyler, 1990; Bozai, 1990; Parker, 1990; Sanderson, 1990). The recommended number of pilot burners, N, based on flare size is provided in Table 1.3 (Tkatschenko, 1990; Tyler, 1990; Bozai, 1990; Parker, 1990; Sanderson, 1990).

Table 1.3: Number of Burners by Flame Tip Diameter

Flare Tip Diameter (in)	Number of Pilot Burners (N)
1 - 10	1
12 - 24	2
30 - 60	3
> 60	4

The annual pilot gas consumption, F_{pi} is calculated by:

$$F_{pi} = Q_{pi} \times N \times t_{op} \times 10^{-3} \qquad (1.9)$$

where:

F_{pi}	=	annual pilot gas consumption (Mscf/yr)
N	=	number of pilot burners
t_{op}	=	total time flare was in operation (i.e., capable of receiving waste gas) in the year (hr/yr); typically assume t_{op} = 8,760 hr/yr.

Q_{pi} = average pilot gas consumption based on pilot burner design (scf/hr); can use default of 70 scf/hr for an energy-efficient pilot burner if actual pilot gas consumption is not available from the manufacturer

10^{-3} = conversion factor to Mscf (Mscf/scf)

1.3.6   Steam Requirement

The steam requirement depends on the composition of the vent gas being flared, the steam velocity from the injection nozzle, and the flare tip diameter. Although some gases can be flared smokelessly without any steam, typically 0.01 to 0.6 pound of steam per pound of flare gas is required (Tkatschenko, 1990; Tyler, 1990; Bozai, 1990; Parker, 1990; Sanderson, 1990). The ratio is usually estimated from the molecular weight of the gas, the carbon-to-hydrogen ratio of the gas, or whether the gas is saturated or unsaturated. For example, olefins, such as propylene, require higher steam ratios than would paraffin hydrocarbons to burn smokelessly (Kalcevic, 1980).

In any event, if a proprietary smokeless flare is purchased, the manufacturer should be consulted about the minimum necessary steam rate. A small diameter flare tip (less than 24 inches) can use steam more effectively than a large diameter tip to mix air into the flame and promote turbulence (Kalcevic, 1980). For a typical refinery, the average steam requirement is approximately 0.25 pounds of steam per pound of flared gas (lb/lb), with this number increasing to 0.5 lb/lb in chemical plants where large quantities of unsaturated hydrocarbons are flared (Sanderson, 1990).

For general consideration, the quantity of steam required, M_{steam}, can be assumed to be 0.4 pounds of steam per pound of flare gas, Q. Using a 0.4 ratio, the amount of steam required is:

$$M_{steam} = 0.4 \times Q \times 60 \times \left(\frac{MWt}{MVC} \right) \qquad (1.10)$$

where:

M_{steam} = quantity of steam required (lb/hr)
0.4 = assumed ratio of steam to flare gas (lb/lb)
Q = volumetric flow rate of the vent gas stream (scfm)
60 = conversion of minutes to hours (min/hr)
MWt = molecular weight of the vent gas stream (lbs/lb-mol)
MVC = molar volume conversion factor (scf/lb-mol) = 385.3 scf/lb-mol

Operating a flare at too high a steam-to-gas ratio is not only costly, but also results in a lower combustion efficiency and a noise nuisance. The capacity of a steam-assisted flare to burn smokelessly may be limited by the quantity of steam that is available.

Depending on the flare design and location, flares may have a minimum steam flow rate. Based on data for 31 industrial flares ranging in size from 14 to 60 inches in diameter,

minimum steam rates averaged 46 lbs/hr steam per inch of flare diameter (API/NPRA/ACC, 2011).

1.3.7 Knock-out Drum

As explained previously, knock-out drums are used to remove any liquids that may be in the vent stream. Two types of drums are used: horizontal and vertical. The economics of vessel design influences the choice between a horizontal and a vertical drum. When a large liquid storage vessel is required and the vapor flow is high, a horizontal drum is usually more economical. Vertical separators are used when there is small liquid load, limited plot space, or where ease of level control is desired. Flares designed to control continuous vent streams generally have vertical knockout drums, whereas emergency flares typically have horizontal vessels. As this chapter focuses on cost for flares used as an emissions control device rather than an emergency safety device, the procedure described below for estimating the size of a knockout drum applies to vertical drums exclusively. A typical vertical knock-out drum is presented in Figure 1.2. Vendor quotes should be obtained to size and cost horizontal knock-out drums for emergency flare systems.

Liquid particles will separate when the residence time of the vapor is greater than the time required to travel the available vertical height at the dropout velocity of the liquid particles, *i.e.*, the velocity is less than the dropout velocity. In addition, the vertical gas velocity must be sufficiently low to permit the liquid droplets to fall. Since flares are designed to handle small-sized liquid droplets, the allowable vertical velocity is based on separating droplets from 300 to 600 micrometers in diameter (API/ANSI, 2014). The dropout velocity, *U*, of a particle in a stream, or the maximum design vapor velocity, is calculated as follows (Wu, 1984):

$$U = G \sqrt{\frac{\rho_l - \rho_v}{\rho_v}} \qquad (1.11)$$

where:

U = dropout velocity, or maximum design vapor velocity (ft/sec)
G = design vapor velocity factor (ft/sec)
ρ_l = liquid density (lb/ft^3)
ρ_v = vapor density (lb/ft^3)

Note that in most cases,

$$\frac{\rho_l - \rho_v}{\rho_v} \approx \frac{\rho_l}{\rho_v} \qquad (1.12)$$

When considered, Equation 1.11 becomes:

$$U = G \sqrt{\frac{\rho_l}{\rho_v}} \qquad (1.13)$$

The design vapor velocity factor, G, ranges from 0.15 to 0.25 for vertical gravity separators at 85% of flooding (Wu, 1984):

Once the maximum design vapor velocity has been determined the minimum vessel cross-sectional area, A, can be calculated by:

$$A = \frac{Q_{actual}}{60 \times U} \tag{1.14}$$

where:

$$
\begin{aligned}
A &= \text{minimum vessel cross-sectional area (ft}^2) \\
Q_{actual} &= \text{vent stream flow in actual conditions (acfm)} \\
60 &= \text{conversion from seconds to minutes (sec/min)}
\end{aligned}
$$

The vessel diameter, d_{min}, is then calculated by:

$$d_{min} = 12\sqrt{\frac{4}{\pi}A} = 13.5\sqrt{A} \tag{1.15}$$

In accordance with standard head sizes, drum diameters in 6-inch increments are assumed so:

$$d = d_{min} \quad \text{(rounded to the next largest size)} \tag{1.16}$$

Some vertical knock-out drums are sized as cyclones and utilize a tangential inlet to generate horizontal separating velocities. Vertical vessels sized exclusively on settling velocity (as in the paragraph above) will be larger than those sized as cyclones (Shore, 1990).

The thickness of the knock-out drum vessel wall is dependent on the size of the vessel and the maximum expected pressure of the system. For the purposes of developing initial costs, the vessel thickness, t, is determined from the diameter as shown in Table 1.4 (Piedmont HUB, 1990). The American Society of Mechanical Engineers (ASME) Boiler and Pressure Vessel Code, Section VIII, Division 1 should be reviewed to determine whether the vessel thickness estimated using Table 1.4 is sufficient for the maximum pressure at the knockout drum. Proper vessel height, h, is usually determined based on required liquid surge volume. The calculated height is then checked to verify that the height-to-diameter ratio is within the economic range of 3 to 5 (Wu, 1984). For small volumes of liquid, as in the case of continuous VOC vent control, it is not necessary to provide more liquid surge than is necessary to satisfy the h/d > 3 condition. So for purposes of flare knock-out drum sizing:

$$h = 3d \tag{1.17}$$

Table 1.4: Vessel Thickness based on Diameter[3]

Diameter, d (inches)	Thickness, t (inches)
$d < 36$	0.25
$36 < d < 72$	0.37
$72 < d < 108$	0.55
$108 < d < 144$	0.75
$d > 144$	1.0

Each knock-out drum will need equipment to control the liquid level and handle the collected water. If systems are not available for handle the collected water, there may be significant additional costs associated with the knock-out drum. For more complex flare collection systems, multiple knock-out drums may be required.

1.3.8   Liquid Seal

As noted in Section 1.2.3, a liquid seal can be used downstream of the knockout drum or incorporated into the same vessel. Liquid seals are common for flares that may be on standby for significant periods or flares for which flare gas recovery systems are used. For flares used for VOC control of continuously operated sources, a liquid seal may not be required. However, if a liquid seal is not used, a flame arrestor or detonation arrestor must be used to prevent flame flashbacks. Detonation flame arrestors come in pipe connection sizes from 1 to 24 inches (Emerson, 2017).

1.3.9   Gas Mover System and Flare Gas Recovery

Generally, flare systems are designed for limited pressure drop so waste gases can be discharged to the flare without mechanically induced flow. The total system pressure drop is a function of the available pressure of the waste gas stream, the flow rates of waste gas, auxiliary fuel and assist gas, and the design of the various flare system components. The estimation of actual pressure drop requirements involves complex calculations based on the specific system's vent gas properties and equipment used. For the purposes of this section, however, approximate values can be used. The design pressure drop through the flare tip can range from 0.1 to 2 psi with the approximate pressure drop relationships provided in Table 1.5. The total system pressure drop ranges from about 1 to 25 psi (Shore, 1990). A gas mover system (fan, blower, or compressor) can be used if the vent stream pressure is less than the pressure drop of the system to enable the stream to be discharged to the flare. Some gas mover systems may increase the pressure in the flare gas collection header, which may impact the ability for other streams to vent to the flare.

One specific type of gas mover system is a flare gas recovery system. Flare gas recovery systems are generally sized to recover routine gas flow with some added capacity to

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[3] Preliminary thickness estimates; ASME Boiler and Pressure Vessel Code, Section VIII, Division 1 should be reviewed to ensure the vessel thickness is sufficient for the maximum vessel pressure expected.

reduce flaring during high flow events. Based on a review of flare gas recovery system projects reported in the literature (U.S. DOE, 2005; John Zink, 2006; Envirocomb, 2006; U.S. EPA, 2008; Zadakbar et al., 2008) flare gas recovery systems can operate at 30 to 90 percent of capacity, with an average capacity utilization rate of 75 percent. The amount of additional flare gas recovery capacity to be installed is dependent on the magnitude of the normal flow, the fluctuations expected in the flare gas flow, the need to recover high flow events (due to local requirements), and the need ensure recovery during compressor maintenance. If a single compressor is used, the compressor can be sized to recovery all normal flow rates at 70 to 80 percent of capacity. However, no recovery will be available during maintenance of the compressor and the compressor will have limited capacity to recover gas during high flow events. If a two-compressor system is used, each compressor can be designed to recover 120 percent of the routine gas flow. This will allow full recovery of normal gas flow while one compressor is being serviced for maintenance and provides excess capacity for high flow events. In a three-compressor system, one can design each compressor to recovery 60 percent of the normal flow, such that two compressors have adequate capacity to recovery normal flow events while providing excess capacity to limit flaring during maintenance or high flow events.

Table 1.5: Design Pressure Losses through the Flare Tip

Equipment	Approximate Pressure Loss
Gas seal:	1 to 3 times flare tip pressure drop
Stack:	0.25 to 2 times flare tip pressure drop
Liquid seal and knock-out drum:	1 to 1.5 times flare tip pressure drop *plus* pressure drop due to liquid depth in the seal, which is normally 0.2 to 1.5 psi.
Gas collection system:	Calculated based on diameter, length, and flow. System is sized by designer to utilize the pressure drop available and still leave a pressure at the stack base of between 2 and 10 psi.

Sizing of flare gas recovery compressors is dependent on the inlet pressure, the outlet pressure, and flare gas recovery system capacity, Q_{cap}. The total horsepower (Hp) of the system typically ranges for 60 to 100 Hp per million cubic feet per day of flow capacity with 75 Hp as a common midpoint (JW Power, 2017).[4] As electricity costs are commonly provided in kilowatt hours (kW-hr), it is convenient to estimate the total compressor power of the flare gas recovery system in terms of kilowatts (kW) rather than horsepower as follows.

$$E_{FGR} = 75 \times \left(0.00144\, Q_{cap}\right) \times 0.746 = 0.0806\ Q_{cap} \qquad (1.18)$$

where:

E_{FGR} = total compressor power for flare gas recovery system (kW)
75 = default Hp per million scf/day of flare gas recovery capacity (Hp/MMscfd)

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[4] Based on inlet suction pressure of 3 psig and outlet pressure of 40 psig.

$$
\begin{array}{rll}
Q_{cap} & = & \text{flare gas recovery system capacity (scfm)} \\
0.00144 & = & \text{conversion from scfm to MMscfd (MMscfd/scfm)} \\
0.746 & = & \text{conversion from Hp to kW (kW/Hp)}
\end{array}
$$

For a multi-compressor flare gas recovery system is used, it is useful to estimate the compressor power (electricity need) per compressor.

Flare gas recovery is generally used when the recovered gas can be used for beneficial purposes (e.g., used in process heaters or boilers). In these cases, the recovered gas can offset natural gas purchases. The total heat value of the recovered gas can be used to determine the amount of natural gas purchases that can be offset.

$$ F_{offset} = (V_{rec} \times B_v)/B_f \tag{1.19} $$

where:

$$
\begin{array}{rll}
F_{offset} & = & \text{annual volume of natural gas purchases offset (Mscf)} \\
Vrec & = & \text{annual volume of waste gas recovered (Mscf)} \\
B_v & = & \text{heat content of the waste gas stream (Btu/scf)} \\
B_f & = & \text{heat content of the natural gas (Btu/scf); can assume } B_f = 920 \text{ Btu/scf when} \\
& & \text{heat content of waste gas is determined on a net or lower heating value basis} \\
& & \text{or assume } B_f = 1{,}030 \text{ Btu/scf when heat content of waste gas is determined on} \\
& & \text{a higher heating value basis.}
\end{array}
$$

1.4 Estimating Total Capital Investment

The capital costs of an elevated, steam-assisted flare system are presented in this section and are based on the design/sizing procedures discussed in Section 1.3. The costs presented are in 2017 dollars.[5] Most of the capital costs for this chapter were updated through vendor contacts in the summer of 2000 (i.e., from EPA Air Pollution Control Cost Manual, Sixth Edition, January 2002) and escalated to 2017 dollars using the Chemical Engineering Plant Cost Index (CEPCI). Costs for some auxiliary equipment are based on recent vendor data.

As discussed in sections 1.4.1 – 1.4.2 below, total capital investment (*TCI*) includes the equipment costs (*EC*) for the flare itself, the cost of auxiliary equipment including monitoring equipment costs, the cost of taxes, freight, and instrumentation, and all direct and indirect installation costs.

The capital cost of flares depends on the degree of sophistication desired (i.e., manual versus automatic control) and the number of appurtenances selected, such as knock-out drums,

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[5] For cost escalation or de-escalation, one suggested index is the Chemical Engineering Plant Cost Index (CEPCI). More information on CEPCI values and the indexing procedure can be found at http://www.chemengonline.com/pci-home. Other indexes are also available. For more information on cost escalation or de-escalation, please refer to the cost methodology chapter in the Control Cost Manual (Section 1, Chapter 2).

seals, controls, ladders, and platforms. The basic support structure of the flare, the size and height, and the auxiliary equipment are the controlling factors in the cost of the flare. The capital investment will also depend on the availability of utilities such as steam, natural gas, and instrument air.

The total capital investment is a battery limit cost estimate and does not include the provisions for bringing utilities, services, or roads to the site, the backup facilities, the land, the research and development required, or the process piping and instrumentation interconnections that may be required in the process generating the waste gas. These costs are based on a new plant installation; no retrofit cost considerations such as demolition, crowded construction working conditions, scheduling construction with production activities, and long interconnecting piping are included. These factors are so site-specific that no attempt has been made to provide their costs. In some cases, the additional costs may be limited to additional transport piping; in other cases, significant investment in utilities and site preparation may be required. The additional retrofit costs should be assessed on a case-by-case basis.

1.4.1   Equipment Costs

As was done for a previous version of this chapter, flare vendors were asked to provide budget estimates for the spectrum of commercial flare sizes. Quotes from NAO, Inc., Kaldair, Inc., Peabody Engineering Corp., John Zink Hamworthy Co., and Flaregas Corp. were used to develop the equipment cost correlations for flare units (Tkatschenko,1990; Tyler, 1990; Bozai, 1990; Parker, 1990; Sanderson, 1990), while the cost equations for knock-out drums were based on estimations from *Chemical Engineering* and *Process Plant Construction Estimating Standards* (Mulet, 1981; Richardson Engineering, 1988) and the cost equations for flare piping were based on estimations from *Plant Design and Economics for Chemical Engineers* and a quote from Piedmont HUB, Inc. (Peters and Timmerhaus, 1980; Piedmont HUB, 1990). The expected accuracy of these costs is ± 30% (i.e., "study" estimates), as are the cost estimates for other chapters in the Manual. Keeping in mind the height restrictions discussed in Section 1.2.5, these cost correlations apply to flare tip diameters ranging from 1 to 60 inches and stack heights ranging from 30 to 500 feet. The standard construction material is carbon steel except when it is standard practice to use other materials, as is the case with burner tips.

The flare costs are presented in Equations 1.20 through 1.22 in 2017$ and are based on support type as follows:

Self-supported Group:

$$C_F = (93.6 + 10.97D + 0.899H)^2 \qquad\qquad (1.20)$$

Guy-supported Group:

$$C_F = (124 + 10.42D + 0.564H)^2 \qquad\qquad (1.21)$$

Derrick-supported Group:

$$C_F = (91.7 + 3.26D + 1.968H)^2 \qquad\qquad (1.22)$$

where:

$$C_F \quad = \quad \text{flare cost (2017 \$)}$$
$$D \quad = \quad \text{diameter of the flare tip (in)}$$
$$H \quad = \quad \text{flare stack height (ft); 30 ft minimum}$$

The equations are least-squares regression of cost data provided by different vendors. It must be kept in mind that even for a given flare technology (i.e., elevated, steam-assisted), design and manufacturing procedures vary from vendor to vendor, so that costs may vary. Once a study estimate is completed, it is recommended that several vendors be solicited for more detailed cost estimates.

Each of these costs includes the flare tower (stack) and support, burner tip, pilots, utility (steam, natural gas) piping from base, utility metering and control, liquid seal, gas seal, and galvanized caged ladders and platforms as required. Costs are based on carbon steel construction, except for the upper four feet and burner tip, which are based on a 310 stainless steel construction. The updated costs (costs escalated to 2017$) are presented in Tables 1.6 to 1.8 and Figures 1.6 to 1.8 (Kalcevic, 1980; Tyler, 1990).

While the costs for utility metering and control are included in the flare equipment costs estimates, the utility metering and controls included are generally only able to make coarse adjustments. These coarse adjustments are acceptable when complying with the flare requirements in 40 CFR 60.18 and 40 CFR 63.11, but finer adjustments may be needed to meet the combustion zone NHV requirements in 40 CFR 63.670. Additional metering and control systems for assist air or assist steam may be needed.

The equipment costs for the flare do not include monitoring equipment costs that may be needed to demonstrate compliance with applicable operating limits for flares. The key monitoring systems that may be needed include presence of a flame monitor, flare vent gas flow rate meter, and net heat content monitor. Thermocouples or ultraviolet monitors are commonly used to monitor for the presence of a flame. Ultrasonic flow meters are commonly used for flares because of the wide flow range these instruments can measure and the ability to measure flow without an additional pressure drop in the flare vent line. Alternatively, flow rates can be estimated using pressure monitoring in the vent gas line and engineering calculations. Net heat content is typically measured using a calorimeter or calculated based on flare gas composition. Other monitoring systems may also be required depending on the flare application. For example, new flares at petroleum refineries may also be required to install a total reduced sulfur monitor for the flare vent gas. Facilities may also elect to install a H_2 analyzer to account for the special H_2 heat content value provided in 40 CFR 63.670. Consequently, flare monitoring equipment costs may include costs for a variety of additional monitoring equipment.

Table 1.6: Self-supporting Flare Costs

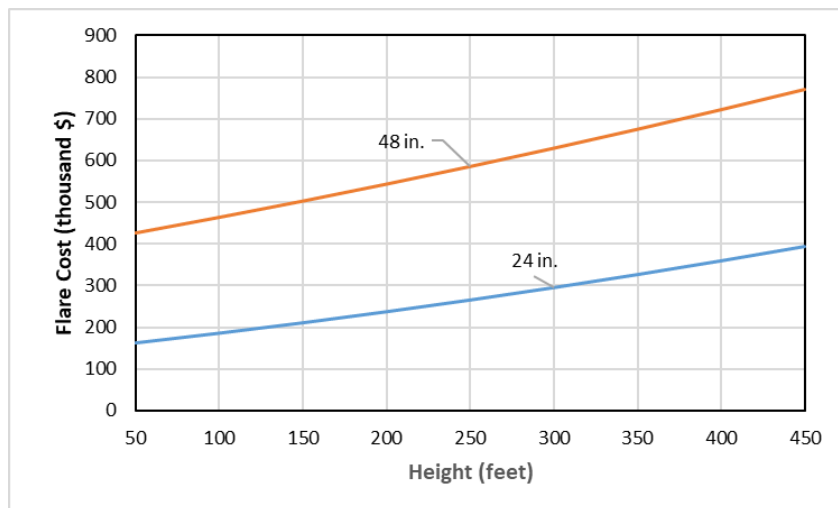| D_f (Diameter in Inches)	H_f (Height in Feet)	C_F (in 2017 Dollars)
12	30	$63,600
12	40	$68,200
12	50	$73,000
12	60	$77,900
12	70	$83,000
12	80	$88,300
12	90	$93,700
12	100	$99,300
24	30	$147,300
24	40	$154,300
24	50	$161,500
24	60	$168,800
24	70	$176,200
24	80	$183,900
24	90	$191,700
24	100	$199,600
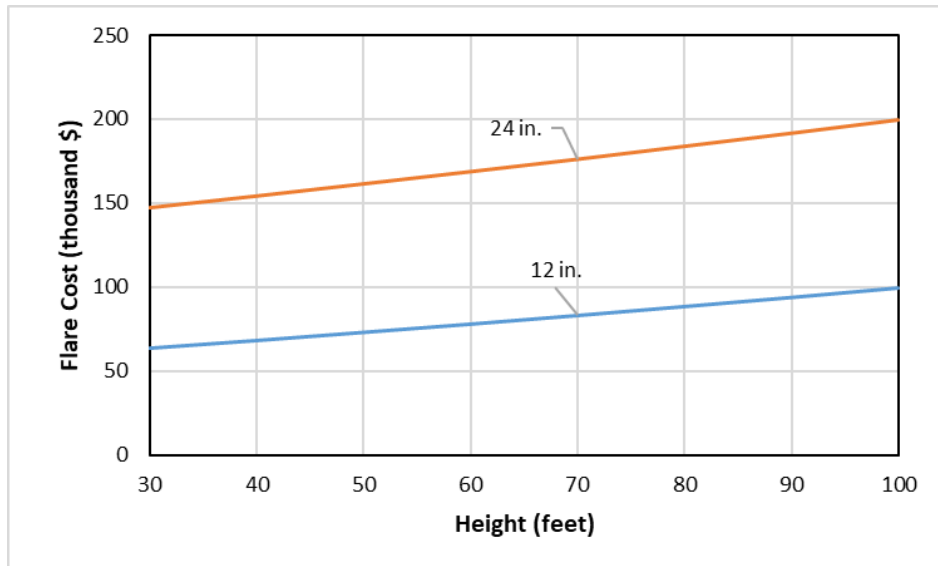
Figure 1.6: Capital Costs of Self-supporting Flares for 12 in. and 24 in. Diameters

Table 1.7: Guy-supported Flare Costs

D$_f$ (Diameter in Inches)	H$_f$ (Height in Feet)	C$_F$ (in 2017 Dollars)
24	50	$161,800
24	100	$185,300
24	150	$210,400
24	200	$237,100
24	250	$265,300
24	300	$295,200
24	350	$326,600
24	400	$359,600
24	450	$394,200
48	50	$425,600
48	100	$463,200
48	150	$502,300
48	200	$543,100
48	250	$585,500
48	300	$629,400
48	350	$675,000
48	400	$722,100
48	450	$770,800

Figure 1.7: Capital Costs of Guy-supported Flares for 24 in. and 48 in. Diameters

Table 1.8: Derrick-supported Flare Costs

D$_f$ (Diameter in Inches)	H$_f$ (Height in Feet)	C$_F$ (in 2017 Dollars)
36	200	$363,200
36	250	$491,500
36	300	$639,100
36	350	$806,200
36	400	$993,000
36	450	$1,198,000
36	500	$1,423,000
36	550	$1,668,000
36	600	$1,932,000
54	200	$437,400
54	250	$577,200
54	300	$736,400
54	350	$915,000
54	400	$1,113,000
54	450	$1,330,000
54	500	$1,567,000
54	550	$1,823,000
54	600	$2,098,000
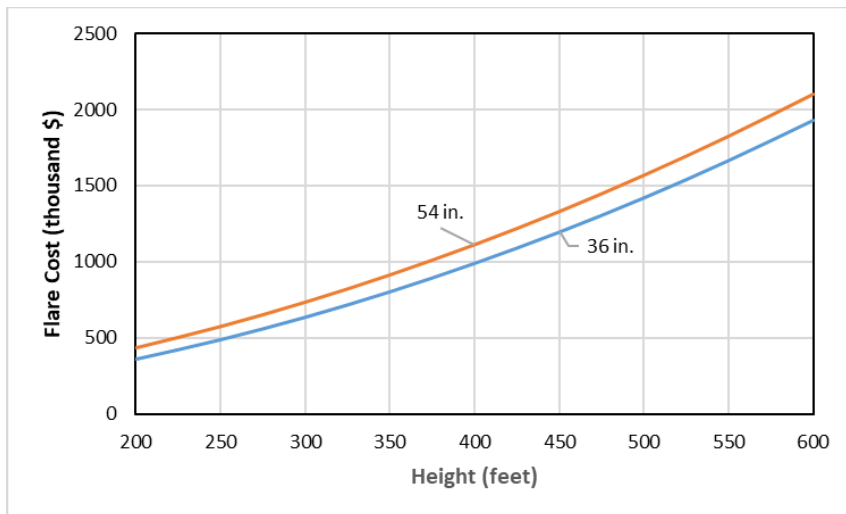
Figure 1.8: Capital Costs of Derrick-supported Flares for 36 in. and 54 in. Diameters

Equipment costs for monitors that detect the presence of a flame are generally low relative to other monitoring systems. Equipment costs for flow monitoring systems and steam controls were estimated based on data submitted by Marathon Petroleum Company for 5 facilities with a total of 20 flares (Coburn, 2014). Calorimeter costs were estimated based on vendor quotes from four vendors (Coburn, 2014). Gas chromatography costs were estimated based on vendor quotes from five vendors (Coburn, 2014). These monitoring system costs were representative of costs in 2010; these costs were escalated to 2017 costs using CEPCI.[6] Hydrogen analyzer costs were estimated based on a single vendor quote in 2015 (Coburn, 2015); these costs were escalated to 2017 costs using CEPCI. Most monitoring system costs are not significantly dependent on the size of the flare, so the flare monitoring system costs provided in Table 1.9 are generally applicable for all flare applications, as needed.

Table 1.9: Equipment Costs for Flare Monitoring Systems[6]

Monitoring System Type	Equipment Costs (2017$)
Pilot Flame Monitor[a]	4,100
Gas Chromatograph (GC)	131,000
H_2 Analyzer (supplemental)	36,900
Calorimeter	77,300
Flare Gas Flow Monitor	58,000
Steam Fine Controls/Metering	58,700
Air Fine Controls/ Metering	49,600

[a]The pilot system monitoring costs are based on thermocouples monitoring 3 pilot flames; if more than 3 pilots are to be monitored, then add $500 for each pilot flame over 3 that will be monitored.

Assuming one flare vent gas monitoring location is needed for a given flare, and only a pilot flame monitor, flow monitor and calorimeter is needed, the flare monitoring equipment cost (C_M) is estimated to be $139,400 (4,100 + 58,000 + 77,300) per flare. If a fine steam control/metering system is also required, the C_M is estimated to be $198,100 ($139,400 + 58,700) per flare. It is important to note that applicable rule subparts should provide provisions stating how owners or operators using flares shall monitor these control devices and that these requirements should be considered before estimating flare monitoring equipment costs for a given application. The equipment costs for all required monitoring systems for a given flare application should be included in the C_M term.

The gas collection header and transfer line requirements are very site specific and depend on the process facility where the emission is generated and on where the flare is located. For the purposes of estimating capital cost, it is assumed that the transfer line will be

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[6] Note that the costs presented in the body of the memorandum of Coburn (2014) are total capital investment costs and includes costs of ancillary materials, installation, and supervision. The costs presented in Table 1.9 are costs for the purchased equipment only based on data presented in Attachment 3 of Coburn (2014). Although recent rules have allowed mass spectrometers as an alternative to GCs and calorimeters, the cost for such monitoring equipment have not been included in this chapter because they are not included in the Coburn (2014) memorandum.

the same diameter as the flare tip (Tkatschenko, 1990) and will be 100 feet long. Most installations will require much more extensive piping, so 100 feet is considered a minimum.

The costs for vent stream piping are presented separately in Equation 1.23 or 1.24 based on the diameter of the flare tip (Piedmont HUB, 1990).

$$C_P = 183 \left(\frac{L}{100}\right) D^{1.21} \quad (where \quad 1" < D < 24") \tag{1.23}$$

$$C_P = 200 \left(\frac{L}{100}\right) D^{1.07} \quad (where \quad 30" < D < 60") \tag{1.24}$$

where:

C_P = vent stream piping cost (2017 $)
L = length of pipe run (ft); 100 ft minimum
D = diameter of the pipe/flare tip (in)

The costs for piping (C_p) as calculated using Equation 1.23 or 1.24 include costs for straight, Schedule 40, carbon steel pipe only.

The costs for a knock-out drum are presented separately in Equation 1.25 (Mulet, 1981; Richardson Engineering, 1988).

$$C_K = 20.5 \left[d \times t \times (h + 0.812\, d)\right]^{0.737} \tag{1.25}$$

where:

C_K = knock-out drum cost (2017 $)
d = drum diameter (in)
t = vessel thickness (in), based on drum diameter
h = height (in)

The costs for a liquid seal are included in the equipment costs for the flare. For flares with routine flow, a flame arrestor may be required rather than a liquid seal. The costs for a flame detonator arrestor (applicable for flares with a diameter of 24 inches or less) are presented separately in Equation 1.26 (based on data from Harmon, 2018).

$$C_S = 39.15D^2 + 3{,}592 \tag{1.26}$$

where:

C_S = flame arrestor cost (2017 $)
D = diameter of the pipe/flare tip (in)

The costs for flare gas recovery systems (C_{FGR}), if applicable, are estimated based on the design capacity of the system. The design capacity of the system can be estimated based on the largest flow required to be collected or based on the routine flow to the flare and a design capacity utilization factor. A capacity utilization factor of 0.7 to 0.8 is typical.

Equipment costs for a flare gas recovery system are presented separately in Equation 1.27 based on the data presented in Table 1.10.

$$C_{FGR} = 731.3 Q_{cap} \qquad (1.27)$$

where:

C_{FGR} = flare gas recovery system costs (2017 $)
Q_{cap} = design flare gas recovery capacity of system (scfm).

Table 1.10: Equipment Costs for Flare Gas Recovery Systems

Project No.	Installed Capital Cost[a]	Year[b]	CEPCI	Installed Cost 2017$[c]	Estimated Equip/Cost 2017$[d]	Capacity (MMscfd)[e]	Reference
1	$4,300,000	2006	499.6	$4,884,400	$1,991,000	4.9	John Zinc (2006)
2	$19,000,000	2006	499.6	$21,582,300	$8,797,500	17.3	John Zinc (2006)
3	$900,000	2006	499.6	$1,022,300	$416,700	1.0	Arcadis (2008)
4	$1,000,000	2005	468.2	$1,212,100	$494,100	0.73	U.S. DOE (2005)
5	$700,000	2008	575.4	$690,400	$281,400	0.54	Zadakbar et al. (2008)
6	$1,400,000	2008	575.4	$1,380,800	$562,800	21.2	Zadakbar et al. (2008)
7	$7,425,000	2006	499.6	$8,434,100	$3,438,000	6.0	Envirocomb (2006)

[a] The capital costs for Projects 3-7 were reported to be installed costs. Projects 1 and 2 did not directly report capital costs. The installed capital costs for these projects were estimated using the reported recovery credits and the payback period.
[b] Projects 3, 5, 6, and 7 did not reference year for cost estimates; publication year of the study was assumed.
[c] The CEPCI for 2017 used to escalate capital costs to 2017$ was 567.5.
[d] Installed equipment costs were estimated using the capital cost equipment in Section 3.2, Chapter 1, Table 1-10 of the EPA Air Pollution Control Cost Manual.
[e] Capacity was not provided for Project 4; estimated capacity assuming actual recovery was 70 percent of capacity. Reported capacity for Project 6 appears to be mis-reported or is an outlier. Consequently, data for Projects 4 and 6 were not used in developing cost correlation equation.

The costs for the necessary utilities (C_U) to supply the flare with steam and auxiliary fuel as well as electricity to power the monitoring equipment should be considered on a case-by-case basis.

Flare system equipment cost (EC) is the total of the calculated flare, monitoring system, knock-out drum, piping, and seal costs plus, if applicable, flare gas recovery and utilities costs.

$$EC = C_F + C_M + C_K + C_P + C_S + C_{FGR} + C_U \qquad (1.28)$$

Purchased equipment costs (*PEC*) is equal to *EC* plus factors for ancillary instrumentation (i.e., control room instruments) (0.10), sales taxes (0.03), and freight (0.05) or,

$$PEC = EC(1 + 0.10 + 0.03 + 0.05) = 1.18EC \qquad (1.29)$$

1.4.2 Installation Costs

Direct installation costs cover foundations and supports, equipment handling and erection, piping, insulation, painting, and electrical. Indirect installation costs cover engineering, construction and field expenses, contractor fees, start-up, and performance testing. These direct and indirect installation costs are generally estimated as a factor of the PEC; default factors for the direct and indirect installation costs are provided in Table 1.11. Other direct costs include site preparation and building installation costs, as needed. Depending on the site conditions, the installation costs for a given flare could deviate significantly from costs generated by these average factors. Vatavuk and Neveril provide some guidelines for adjusting the average installation factors to account for other-than-average installation conditions (Vatavuk and Neveril, 1980).

In general, the costs presented here assume that steam is available at the facility either as purchased steam or steam produced from an existing facility boiler. If additional steam generation facilities were required to supply steam for a new steam-assisted flare, these costs can be accounted for under the site preparation term. If buildings are needed to house controls or monitoring equipment, these costs can be accounted for under the building cost term. We do not attempt to provide cost factors or cost algorithms for these costs, because they may not be required and because they are highly variable based on site-specific conditions when they are applicable.

1.4.3 Total Capital Investment (TCI) Costs

The *TCI* is obtained by summing the purchased equipment cost (*PEC*), the direct installation costs, and the indirect installation costs and adding costs for contingencies. Contingency costs are typically estimated as a factor of the total direct costs and installation costs. Contingency costs are typically between 5 and 15 percent of the total direct and indirect costs; a contingency factor (*CF*) of 0.1 can be used as a practical default value. The *TCI* can be calculated as follows.

$$TCI = (1.89PEC + SP + Bldg)(1 + CF) \qquad (1.30)$$

where:

TCI = total capital investment (\$)
PEC = purchased equipment cost (\$)
SP = site preparation costs (\$)
$Bldg$ = building costs (\$)
CF = contingency factor

Table 1.11: Capital Cost Factors for Flare Systems

Cost Item	Factor
Direct Costs	
Purchased equipment costs	
Flare system, EC	As estimated, A
Instrumentation	0.10 A
Sales taxes	0.03 A
Freight	0.05 A
Purchased equipment cost, PEC	B = 1.18 A
Direct installation costs	
Foundations & supports	0.12B
Handling & erection	0.40 B
Electrical	0.01 B
Piping	0.02 B
Insulation	0.01 B
Painting	0.01 B
Direct installation costs	0.57 B
Site preparation	
As required, SP	
Buildings	
As required, Bldg	
Total Direct Costs, DC	1.57 B + SP + Bldg
Indirect installation costs	
Engineering	0.10 B
Construction and Field expenses	0.10 B
Contractor fees	0.10 B
Start-up	0.01 B
Performance test	0.01 B
Total Indirect Costs, IC	0.32 B
Contingencies, C[a]	$CF \times (DC + IC)$
Total Capital Investment = DC + IC + C	$(1.89\ B + SP + Bldg) \times (1 + CF)$

[a] Where "CF" is the contingency factor. Typical values for CF for mature technologies such as flares range from 5 to 15 percent. A value of 0.1 (10 percent) serves as a useful midpoint estimate for the CF.

1.5 Estimating Total Annual Costs

The total annual cost (*TAC*) is the sum of the direct and indirect annual costs. The bases for calculating annual cost factors are given in Table 1.12.

Table 1.12: Suggested Annual Cost Factors for Flare Systems

Cost Item	Factor
Direct Annual Costs, DC	
Operating labor	
Operator	630 to 1,260 man-hours/year[a]
Supervisor	15% of operator
Operating materials	–
Maintenance	
Labor	½ to 1 hour per shift
Material	100% of maintenance labor[a]
Utilities	
Electricity	All utilities equal to:
Purge gas	(Consumption rate) x
Pilot gas	(Hours/yr) x (unit cost)
Auxiliary fuel (use and offset)	
Steam	
Indirect Annual Costs, IC	
Overhead	60% of total labor and material costs
Administrative charges	2% of Total Capital Investment
Property tax	1% of Total Capital Investment
Insurance	1% of Total Capital Investment
Capital recovery[b]	0.0963 x Total Capital Investment
Total Annual Cost	Sum of Direct and Indirect Annual Costs

[a] Use the lower value for flares without flare gas recovery systems; use the higher value for flares with flare gas recovery systems.

[b] See Section 1, Chapter 2 of this Manual.

1.5.1 Direct Annual Costs

Direct annual costs include labor (operating and supervisory), maintenance (labor and materials), natural gas, steam, and electricity. Unless the flare is to be dedicated to one vent stream and specific on-line operating factors are known, costs should be calculated based on a continuous operation of 8,760 hr/yr and expressed on an annual basis. Flares serving multiple process units typically run continuously for several years between maintenance shutdowns.

Operating labor for most industrial applications is estimated at 630 hours annually (U.S. EPA, 1993). A completely manual system could easily require 1,000 hours. Conversely, 130 hours has been estimated for flares to control vented storage tank emissions at unmanned oil production sites (U.S. EPA, 2016). A standard supervision ratio of 0.15 should be assumed. Maintenance labor is estimated at 0.5 hours per 8-hour shift for applications without a flare gas recovery system, or the same as operating labor for unmanned sites. Maintenance labor is estimated at 1 hour per 8-hour shift for applications with a flare gas recovery system. Maintenance materials costs are assumed to equal maintenance labor costs.

Flare gas recovery systems may be treated as a separate industrial application, effectively doubling the operating, supervisory, and maintenance labor and maintenance costs.

Flare utility costs include natural gas, steam, and electricity. Flare systems can use natural gas in three ways: in pilot burners that fire natural gas, in combusting low Btu vent streams that require natural gas as auxiliary fuel, and as purge gas. The total natural gas cost, C_f, to operate a flare system includes pilot, C_{pi}, auxiliary fuel, C_a, and purge costs, C_{pu}, less the offset credits, C_{offset}, associated with the recovered flare gas (when using a flare gas recovery system):

$$C_f = C_{pi} + C_a + C_{pu} - C_{offset} \qquad (1.31)$$

Where, C_{pi} is equal to the annual volume of pilot gas, F_{pi}, multiplied by the cost per scf, $Cost_{fuel}$:

$$C_{pi} = F_{pi} \times Cost_{fuel} \qquad (1.32)$$

C_a, C_{pu}, and C_{offset} are similarly calculated.

For systems with flare gas recovery, the recovered gas can be used to offset natural gas purchase. The total heat content of the gas recovered annually can be calculated based on the volume of gas recovered and the heat content of the recovered gas and the quantity of natural gas determined.

Steam cost (C_s) to eliminate smoking is equal to the annual steam consumption, multiplied by the cost per lb, $Cost_{Steam}$:

$$C_s = t_{op} \times M_{Steam} \times Cost_{Steam} \qquad (1.33)$$

The price (cost per pound) of steam is dependent on the pressure of the steam required. Most steam-assisted flares require high pressure steam (100 to 150 psig). For costing purposes, use steam prices assuming 150 psig steam is required.

Electricity is needed for gas mover or flare gas recovery systems. Electricity may also be needed for pumps to transport liquids collected in the knock-out drum. Electricity cost (C_e) is equal to electricity consumption, multiplied by the cost per kW-hr, $Cost_{electricity}$:

$$C_e = t_{FGR} \times E_{FGR} \times Cost_{electricity} \qquad (1.34)$$

where:

t_{FGR} = time the flare gas recovery system is operated (hr).

1.5.2 Indirect Annual Costs

The indirect (fixed) annual costs include overhead, capital recovery, general and administrative (G & A) charges, property taxes, and insurance. Suggested indirect annual cost factors are presented in Table 1.12.

Overhead is calculated as 60% of the total labor (operating, maintenance, and supervisory) and maintenance material costs. Overhead cost is discussed in Section 1, Chapter 2 of this Manual.

The system capital recovery cost (*CRC*) is based on an estimated 15-year equipment life. (See Section 1, Chapter 2 of this Manual for a thorough discussion of the capital recovery cost and the variables that determine it.) For a 15-year life and an interest rate of 5%, the capital recovery factor is 0.0963. The system capital recovery cost is the product of the system capital recovery factor (*CRF*) and the *TCI*, or:

$$CRC = CRF \times TCI = 0.0963 \times TCI \qquad (1.35)$$

As shown in Table 1.12, G & A, taxes, and insurance can be estimated at 2%, 1%, and 1% of the *TCI*, respectively.

1.5.3 Example Flare Costs

Annual operating costs for pilot fuel and maintenance were estimated to be $2,965. Capital recovery costs in the Colorado Department of Public Health and Environment (CDPHE) analysis were estimated using a discount rate of 5 percent and an assumed equipment life of 15 years.

1.6 Example Problem 1 (Flare without Flare Gas Recovery)

The example problem described in this section shows how to apply the elevated, steam-assisted flare sizing and costing procedures to the control of a vent stream associated with the distillation manufacturing of methanol.

1.6.1 Required Information for Design

The first step in the design procedure is to determine the specifications of the vent gas to be processed. The minimum information required to size a flare system for estimating costs are the vent stream:

- Volumetric or mass flow rate
- Heating value or chemical composition temperature
- System pressure
- Vapor and liquid densities

In addition, the following are needed to calculate direct annual costs.

- Labor costs
- Fuel costs
- Steam costs
- Electricity costs

Vent stream parameters and cost data to be used in this example problem are listed in Table 1.13.

Table 1.13: Example Problem Data

Vent Stream Parameters

Condition 1: High waste gas flow rate[a]	3,200 scfm
Condition 2: Typical waste gas flow rate[a]	600 scfm
Condition 3: Low waste gas flow rate[a]	50 scfm
Condition 4: No waste gas flow rate[a]	0 scfm
High flow heat content	750 Btu/scf
Typical and low flow heat content	450 Btu/scf
Average gas stream molecular weight	31 lb/lb-mol
System pressure[b]	10 psig
Temperature	110 °F
Liquid density [c, d]	49.60 lb/ft^3
Vapor density [c]	0.0845 lb/scf
Maximum flow rate at flare tip[e]	3,120 acfm

Cost Data (2014)

Condition 1: Time with high flow	260 hrs/yr
Condition 2: Time with typical flow	7,000 hrs/yr
Condition 3: Time with low flow	1,000 hrs/yr
Condition 4: Time with no flow	500 hrs/yr
Natural gas[f]	$ 4.14 per 1000 scf
Steam[g]	$ 7.70 per 1000 lbs
Electricity[h]	$0.0688 per kW-hr
Operating labor[i]	$ 29.63 per hr
Maintenance labor[i]	$ 25.12 per hr

[a] Standard conditions: 68 °F, 1 atmosphere.
[b] Pressure at source (gas collection point). Pressure at flare tip is lower: 1 psig.
[c] *Handbook of Chemistry and Physics*, Haynes, ed., 2015.
[d] Measured at standard conditions.
[e] Actual conditions at flare tip: 90 °F, 15.7 psia.
[f] Annual average industrial price for natural gas published by the U.S. Department of Energy, Energy Information Administration (see https://www.eia.gov/dnav/ng/ng_pri_sum_dcu_nus_a.htm).
[g] Calculated based on industrial natural gas price, 85.7% overall boiler efficiency for delivered steam, NHV value for natural gas of 920 Btu/scf, steam pressure of 150 psig and inlet water temperature of 100 deg. F (resulting in an energy need of 1,128 Btu/lb steam; see U.S. DOE, 2012), and multiplying the fuel cost by 1.3 to get the total cost of steam (see U.S. DOE, 2003).
[h] Average industrial price for electricity in 2017 (U.S. EIA, 2018).
[i] Bureau of Labor Statistics, median labor costs for occupation codes 51-8091 and 49-9040, 2017.

1.6.2 Capital Equipment

The first objective is to properly size a steam-assisted flare system to effectively destroy 98% of the VOC (methanol) in the vent gas stream. Using the vent stream parameters and the design procedures outlined in Section 1.3, flare and knock-out drum heights and diameters can be determined. Once equipment has been specified, the capital costs can be determined from equations presented in Section 1.4.1.

1.6.2.1 Equipment Design

The first step in flare sizing is determining the appropriate flare tip diameter. Knowing the net (lower) heating value of the vent stream, the maximum allowed velocity can be calculated from the requirements listed in the Code of Federal Regulations for flares mentioned earlier in this chapter. Since the heating value is in the range of 300 to 1,000 Btu/scf, the maximum velocity, V_{max}, is determined by Equation 1.1. A quick review of the flow conditions suggests that either Condition 1 (high flow) or Condition 2 (typical flow) will provide the maximum flare tip diameter. The allowable flare tip velocity is calculated for both of these conditions.

Parameter	Condition 1	Condition 2
$\log_{10} V_{max}$ (Eqn 1.1)	$(750+1,212)/850 = 2.308$	$(450+1,212)/850 = 1.955$
V_{max} (ft/s)	203	90.2

Because the design vent stream parameters for Conditions 1 and 2 have a net heating value well above 300 Btu/scf, no auxiliary fuel is expected to be required for these flow conditions. Hence, Q_{tot} can be assumed to equal the maximum vent stream flow rate, converted to actual conditions based on the temperature and pressure at the flare stack (110 °F and 1 psig). Based on Q_{tot} and V_{max}, the flare tip diameter can be calculated using Equation 1.5.

Parameter	Condition 1	Condition 2
Q_{tot} (acfm)	$3200[(460+110)/528][14.7/(14.7+1)]$ $= 3,235$	$600[(460+110)/528][14.7/(14.7+1)]$ $= 606.5$
D_{min} (Eqn 1.5; in)	$1.95(3235/203)^{0.5} = 7.8$	$1.95(606.5/90.2)^{0.5} = 5.1$

Flow Condition 1 yields the largest value of D_{min}, so the flare stack diameter, D, is selected from the next largest commercially available standard size, which is 8 inches.

The next parameter to determine is the required height of the flare stack. The heat release from the flare is calculated using Equation 1.7. Condition 1 has the highest flow rate and heat content, so this condition will have the largest heat release. Again, no auxiliary fuel is expected to be required for Condition 1, so Equation 1.7 simplifies to:

$$R = \left(Q_{max} \times B_{v,max}\right) \times 60$$

So,

$$R = 3200 \times 750 \times 60 = 144,000,000 \; \frac{Btu}{hr}$$

Substituting R and appropriate values for other variables into Equation 1.6:

$$H^2 = \frac{(1)(0.3)(144,000,000)}{4\pi(500)} = 6,875 \; ft^2$$

Resulting in:
$$H = 82.9 \; ft$$

This is the minimum distance from the flare flame to the point where radiation must be limited. The point where radiation must be limited should include an allowance for personnel height (7 feet), so the flare height is set to this value, $H = 90$ ft (rounded up from 82.9+7=89.9 ft).

Next the knock-out drum must be sized. Assuming a design vapor velocity factor, G, of 0.20, and substituting the vapor and liquid densities of methanol into Equation 1.11 yields a maximum velocity of:

$$U = 0.20\sqrt{\frac{49.60 - 0.0845}{0.0845}} = 4.84 \; \frac{ft}{sec}$$

Given a maximum vent gas flow rate of 3,200 scfm, the actual flow rate is estimated using the ideal gas law and the system temperature (°R) and pressure (psia) at the knock out drum:

$$Q_{actual} = Q_{max}\left(\frac{T_{actual}}{T_s}\right)\left(\frac{P_s}{P_{actual}}\right)$$

The pressure in the knock-out pot will be somewhere between the system pressure and the flare tip pressure, so 5 psig is used as the pressure in the knock-out pot.

$$Q_{actual} = 3200\frac{(460 + 110)}{(460 + 68)}\left(\frac{14.7}{14.7 + 5}\right) = 2,578 \; acfm$$

The minimum vessel cross-sectional diameter is calculated by Equation 1.14:

$$A = \frac{2,578}{(60)(4.84)} = 8.87 \; ft^2$$

This results in a minimum vessel diameter, calculated using Equation 1.15, of:

$$d_{min} = 13.5\sqrt{8.87} = 40.2 \; in$$

The selected diameter, d, rounded to the next largest 6-inch increment according to Equation 1.16, is 42 inches. Using Equation 1.17, the height to diameter ratio of three gives a vessel height of 126 inches, or 10.5 feet.

1.6.2.2 Equipment Costs

Once the required flare tip diameter and stack height have been determined, the equipment costs can be calculated. Since the height is 90 feet, the flare can be either self-supporting or guy-supported. The costs for an 8-inch diameter, 90-feet tall, self-supported flare are determined from Equation 1.20.

$$C_F = [93.6 + 10.97(8) + 0.899(90)]^2 = \$68{,}786$$

The costs for an 8-inch diameter, 90-feet tall, guy-supported flare are determined using Equation 1.21.

$$C_F = [124 + 10.42(8) + 0.564(90)]^2 = \$66{,}626$$

In this example, the cost for a guy-supported flare are slightly less than for a self-supported flare. Provided that there is adequate room for the guy support wires, selecting a guy-supported flare will reduce the equipment costs slightly.

Knock-out drum costs are determined using Equation 1.25, where t is determined from the ranges presented in Section 1.3.7. Substituting 0.37 for t for a 42-inch diameter, 126-inch tall knockout drum:

$$C_K = 20.5\big[(42)(0.37)\big(126 + 0.812(42)\big)\big]^{0.737} = \$6{,}524$$

Since the flare tip is 8 inches in diameter, the diameter of the transport piping is also 8 inches. Therefore, transport piping costs are determined using Equation 1.23. For this example, we estimate 200 feet of transport piping are needed so the transport piping costs are:

$$C_P = 183 \left(\frac{200}{100}\right)(8)^{1.21} = \$4{,}531$$

Because the flare receives waste gas during most time periods, a flame arrestor rather than a water seal will be used. Flame arrestor costs are estimated using Equation 1.26 and the transport piping/flare tip diameter of 8 inches.

$$C_S = 39.15(8)^2 + 3592 = \$6{,}098$$

Monitoring system costs are dependent on the applicable standard. Pilot flame monitoring is generally always required. For flares complying with requirements in 40 CFR 60.18 or 40 CFR 63.11, a heat content monitor and flow rate monitor may also be needed. For flares complying with 40 CFR 63.670, heat content monitor, flow rate monitor, and enhanced steam controls will typically be needed. In this example, we assume the flare has to meet combustion limits based on the combustion zone NHV, so we assume monitoring systems will be needed for the pilot flame, flare gas flow, heat content (calorimeter), and steam controls.

The fixed monitoring system costs, based on the values provided in Table 1.9, are $4,100 + $77,300 + $58,000+ $58,700 = $198,100.

The total auxiliary equipment cost is the sum of the costs for the knock-out drum, transport piping, seal/flame arrestor, and monitoring system, or: $6,524 + $4,531+ $6,098 + $198,100 = $215,253. Therefore, the flare system equipment cost is $215,253 + $66,626 = $281,879.

PEC and *TCI* are calculated using the factors given in Table 1.11. In this example, we assume no site preparation or buildings are needed and a contingency factor of 10 percent (the midpoint of the recommended range) is appropriate. The calculations are shown in Table 1.14.

$$PEC = 1.18 \times (\$281,879) = \$332,617$$

$$TCI = (1.1) \times 1.89 \times (\$332,617) = \$691,511$$

1.6.3   Operating Requirements

Operating labor is estimated at 630 hours annually with supervisory labor at 15% of this amount. Maintenance labor is estimated at 1/2 hour per shift, which translates to 547.5 hours annually. Maintenance material costs are assumed to be equal to maintenance labor costs.

Since the heat content of the example stream is above 300 Btu/scf, as indicated in Table 1.13, no auxiliary fuel is needed to comply with the flare requirements in 40 CFR 60.18 or 40 CFR 63.11. However, if the flare is required to meet the requirements in 40 CFR 63.670, then one must estimate the steam flow rates for the various flow scenarios using Equation 1.10 and consider the minimum design steam flow rate for the flare.

$$M_{steam} = 0.4 \times Q \times 60 \times \left(\frac{MWt}{MVC} \right)$$

At high flow conditions, $M_{steam,1}$ = 0.4×3200×60×31/385.3 = 6,179 lb/hr.

At typical flow conditions, $M_{steam,2}$ = 0.4×600×60×31/385.3 = 1,159 lb/hr

At low flow conditions, $M_{steam,3}$ = 0.4×50×60×31/385.3 = 96.5 lb/hr

Using the average minimum steam supply rate of 46 lbs/hr of steam per inch of flare diameter, an 8-inch diameter flare could have an expected minimum steam flow rate of 46×8 = 368 lbs/hr steam flow requirement. Thus, if the flare requires a minimum steam flow rate (based on the flare design or location of the flare), more steam will need to be supplied to the flare during the low flow conditions. For this example, we assume the flare has a minimum design steam flow rate so the steam rate at low and no flow conditions, $M_{steam,3}$ and $M_{steam,4}$, equals 368 lbs/hr. Because the design steam flow rates Conditions 1 and 2 (i.e., $M_{steam,1}$ and $M_{steam,2}$), exceed 368 lbs/hr, they stay as calculated above for the time being.

Table 1.14: Capital Cost for Flare Systems - Example Problem 1

Cost Item	Cost[a]
Direct Costs	
Purchased Equipment Costs	
Flare system (Self-supporting)	$66,626
Knock-out drum	$6,524
Vent stream transfer line	$4,531
Flame arrestor	$6,098
Monitoring system	$198,100
Equipment Cost (Sum = A)	$281,879
Instrumentation (0.1A)	$28,188
Sales taxes (0.03A)	$8,456
Freight (0.05A)	$14,094
Purchased equipment cost, PEC	$332,617
Direct Installation Costs	
Foundations & Supports (0.12 PEC)	$39,914
Handling & Erection (0.40 PEC)	$133,047
Electrical (0.01 PEC)	$3,326
Piping (0.02 PEC)	$6,652
Insulation (0.01 PEC)	$3,326
Painting (0.01 PEC)	$3,326
Direct Installation Costs	$189,591
Site Preparation	$0
Facilities and Buildings	$0
Total Direct Costs, DC	**$522,208**
Indirect Installation Costs	
Engineering (0.1 PEC)	$33,262
Construction and Field Expenses (0.1 PEC)	$33,262
Contractor Fees (0.1 PEC)	$33,262
Start-Up (0.01 PEC)	$3,326
Performance Test (0.01 PEC)	$3,326
Total Indirect Costs, IC	**$106,438**
Contingencies, C (0.1 x (DC + IC)	$62,865
Total Capital Investment = 1.1 × (DC + IC) (rounded)	**$691,500**

[a] Costs are in 2017 dollars.

Mass steam flow rates can be converted to a volumetric steam flow rate, S (scfm), using 18 lb/lb-mol for the molecular weight of water, MWt (lb/lb-mol) and ideal gas law molar volume correction factor, MVC, of 385.3 scf/lb-mol as follows:

$$S = \left(\frac{M_{steam}}{60}\right)\left(\frac{MVC}{MWt}\right)$$

Thus, the volumetric steam flow rates are:

At high flow conditions, $S_1 = (6,179/60)(385.3/18) = 2,204$ scfm.

At typical flow conditions, $S_2 = (1,159/60)(385.3/18) = 413$ scfm.

At low and no flow conditions, $S_3 = S_4 = (368/60)(385.3/18) = 131.3$ scfm.

For flares required to meet the requirements in 40 CFR 63.670, the projected combustion zone NHV must be assessed to determine if the steam rates projected above allow compliance with the 270 Btu/scf combustion zone NHV operating limit. One can simply input the gas flow rates, heat content and steam rates into Equation 1.3 and if the calculated auxiliary fuel rate is negative, then no auxiliary fuel is required. Alternatively, one can calculate the combustion zone NHV assuming no auxiliary fuel is used to see if it meets the required limit, and then calculate the amount of auxiliary fuel needed using Equation 1.3 only for those scenarios where the target is not met. The combustion zone net heating value, NHV_{cz}, when there is no auxiliary fuel is calculated as follows:

$$NHV_{cz} = NHV_{vg}\left(\frac{Q}{Q+S}\right)$$

The combustion zone NHV with no auxiliary fuel for the three flow conditions are:

At high flow conditions, $NHV_{cz,1} = (750)(3200)/(3200+2204) = 444$ Btu/scf.

At typical flow conditions, $NHV_{cz,2} = (450)(600)/(600+413) = 267$ Btu/scf.

At low flow conditions, $NHV_{cz,3} = (450)(50)/(50+197) = 91$ Btu/scf.

To ensure continuous compliance, the target NHV_{cz} (or NHV_{target}) will generally be set slightly above the regulatory limit. Using a 10 percent margin, we set NHV_{target} to 297 Btu/scf. At the high flow conditions, the calculated NHV_{cz} easily exceeds NHV_{target}. At the typical flow conditions, the combustion zone NHV almost meets the target. Rather than adding auxiliary fuel under this condition, the steam rate can be reduced to achieve NHV_{target} without the addition of auxiliary fuel. One can re-arrange the above equation to determine the maximum amount of steam that can be used while meeting NHV_{target}.

$$S = Q\left(\frac{NHV_{vg}}{NHV_{target}} - 1\right)$$

Therefore, for the typical flow conditions, the steam rate, S_2, is set to 309 scfm ($600 \times [(450/297)-1]$) so that $M_{steam, 2} = 309 \times 60 \times 18/385.3 = 866$ lbs/hr. For the low flow cases, the steam rate is already set at its minimum, so auxiliary fuel must be added in order to achieve the target combustion zone net heating value. Equation 1.3 is used to determine the amount of auxiliary fuel needed for this flow condition.

$$F_{min} = \frac{Q\left(NHV_{t \, arg \, et} - B_v\right) + NHV_{t \, arg \, et} K_1 S_{min}}{B_f - NHV_{t \, arg \, et}}$$

Assuming natural gas is used as the auxiliary fuel, $B_f = 920$ Btu/scf and the amount of auxiliary fuel needed is:

$$F_{min,3} = \frac{50 \, (297 - 450) + 297 \times 1 \times 131.3}{920 - 297} = 50.31 \; scfm$$

Under 40 CFR 63.670, the NHV_{cz} limit does not apply when no "regulated material" is discharged to the flare. Therefore, under Condition 4 (no waste gas flow), no auxiliary fuel is expected to be required. This "no flow" condition is included because this is time that the flare is still operational (on standby ready to receive waste gas) and will be supplied by purge and sweep gas during this time (so this time must be accounted for under the total operational time for purge and pilot gas use). There may be times when an operator may need or elect to meet the NHV_{cz} limit during this "no waste gas flow" condition. If fuel gas that may contain "regulated material" is used for the purge gas, then the NHV_{cz} limit would apply and the operator would need to meet the limit during this "no waste gas flow" period. If the time period when no "regulated material" is discharged to the flare, it may be easier for the operator (or automated system) to maintain a consistent NHV_{target} value. In this example, the flare is not commonly on standby (no waste gas flow), so the operator may elect to consistently meet a consistent NHV_{target} value, event during times of no waste gas flow. The amount of auxiliary fuel needed to meet the NHV_{target} value during Condition 4 (no waste gas flow) would be:

$$F_{min,4} = \frac{0 + 297 \times 1 \times 131.3}{920 - 297} = 62.59 \; scfm$$

Based on the annual operating time at the low and no flow conditions, the annual quantity of natural gas auxiliary fuel needed is calculated from Equation 1.4.

$$F_{a,min} = \frac{(50.31 \times 1000 + 62.59 \times 500) \times 60}{1000} = 4,896.2 \; \frac{Mscf}{yr}$$

Natural gas is also required for purge and pilot gas. Purge gas requirements are dependent on the diameter of the flare and are calculated from Equation 1.8.

$$F_{pu} = (7.85 \times 10^{-4})(8,760)(8)^2 = 440.1 \; \frac{Mscf}{yr}$$

Since the flare tip diameter is 8 inches, pilot gas requirements are based on one pilot burner, (see Table 1.3) and are calculated by Equation 1.9. When $N = 1$,

$$F_{pi} = (70)(1)(8,760)(10^{-3}) = 613.2 \frac{Mscf}{yr}$$

The total natural gas usage is 4,896.2 + 440.1 + 613.2 = 5,949.5 Mscf/yr if complying with the net heating value combustion zone requirements of 40 CFR 63.670 (or 1,053.3 Mscf/yr if complying with the requirements of 40 CFR 63.11 or 40 CFR 60.18, since no auxiliary fuel is needed).

The annual quantity of steam used is calculated based on the steam assist rates (considering that S_2 was decreased to 866.4 lbs/hr to meet the NHV_{target} value without adding additional auxiliary fuel and S_3 was set to this same minimum) and the time in each flow condition.

$$M_{steam} = (6,179)(260) + (866.4)(7,000) + (368)(1000) + (368)(500)$$

$$= 8,223,300 \frac{lb}{yr} \ (rounded) = 8,223.3 \frac{1,000lb}{yr}$$

1.6.4 Total Annual Costs

Table 1.15 shows the calculations of the direct and indirect annual costs for the flare system as calculated from the factors in Table 1.12. Direct costs include labor, materials, and utilities. Indirect costs are the fixed costs allocated to the project, including capital recovery costs and such costs as overhead, insurance, taxes, and administrative charges. The sum of the direct and indirect annual costs yields a total annual cost of approximately $260,600.

Electrical costs of a mover system (fan, blower, compressor) would have to be included if the vent stream pressure was not sufficient to overcome the flare system pressure drop. In this example case, the pressure is assumed to be adequate.

1.7 Example Problem 2 (Flare with Flare Gas Recovery)

This example problem begins with the same process as in the previous example for the control of a vent stream associated with the distillation manufacturing of methanol, but in this example, the flare system includes a flare gas recovery system.

1.7.1 Required Information for Design

The vent stream parameters and cost data are the same as that used in this Example 1 and are listed in Table 1.13.

1.7.2 Capital Equipment

1.7.2.1 Equipment Design

The flare must still be sized based on the worst-case flow conditions, which would include times when the flare gas recovery system is not operating. Therefore, the sizing of the flare and knock-out drum are identical to Example 1.

Table 1.15: Annual Costs for Flare System - Example Problem 1

Cost Item	Calculations	Cost[a]
Direct Annual Costs, DAC		
Operating Labor		
Operator	$\dfrac{630\,h}{year} \times \dfrac{\$29.63}{h}$	$ 18,667
Supervisor	15% of operator $= 0.15 \times \$18,667$	$2,800
Operating materials		
Maintenance		
Labor	$\dfrac{0.5\,h}{shift} \times \dfrac{shift}{8h} \times \dfrac{8,760\,h}{yr} \times \dfrac{\$25.12}{h}$	$13,753
Materials	100% of maintenance labor	$13,753
Utilities		
Electricity		
Natural gas	$\dfrac{(4,896.2 + 440.1 + 613.2)Mscf}{yr} \times \dfrac{\$4.14}{Mscf}$	$24,631
Steam	$\dfrac{8,223.3 \times 10^3\,lb}{yr} \times \dfrac{\$7.70}{10^3\,lb}$	$63,319
Total DAC		$136,923
Indirect Annual Costs, IAC		
Overhead	60% of total labor and material costs $= 0.6\,(\$18,667 + \$2,800 + \$13,753 + \$13,753)$	$29,384
Administrative charges	2% of Total Capital Investment $= 0.02$ ($691,500)	$13,830
Property tax	1% of Total Capital Investment $= 0.01$ ($691,500)	$6,915
Insurance	1% of Total Capital Investment $= 0.01$ ($691,500)	$6,915
Capital recovery[b]	$0.0963 \times \$691,500$	$66,592
Total IAC		$123,637
Total Annual Cost (rounded)		$260,600

[a] Costs are in 2017 dollars.
[b] The capital recovery cost factor, CRF, is a function of the flare equipment life and the opportunity cost of the capital (i.e., interest rate). For this example, a 15-year equipment life and 5% interest rate yields CRF = 0.0963.

The sizing of the flare gas recovery system depends on the expected flow characteristics of the vent streams to be controlled and considerations of the need to recovery high flow events. In this example, the high flow events only occur 260 hours per year and would require a flare gas recovery system capacity more than 5 times greater than the capacity needed to recover the normal gas flow. Based on these considerations, we elected to size the flare gas recovery system based on the normal flow conditions of 600 scfm. The next consideration is the number of compressors to be used in the system. Using a single compressor system for this flow rate is possible but will likely have reduced effectiveness compared to a two-compressor system due to maintenance issues related to the compressor. For this example, we chose to use a two-compressor system, with each compressor sized to recover 120 percent of the normal flow. In this manner, each compressor would be capable of recovering the normal vent flow, so one compressor would operate during normal flow and more gas during the high flow conditions could be recovered.

1.7.2.2 Equipment Costs

Costing of the flare and knock-out drum are identical to Example 1.

The flare gas recovery system for retrofit applications may require additional transport piping if there is not adequate space near the flare for the flare gas recovery system. In this example, we will assume 300 ft of transport piping is required (rather than the 200 ft assumed in Example 1). The transport piping costs are estimated to be:

$$C_P = 183 \ \left(\frac{300}{100}\right) (8)^{1.21} = \$6,797$$

The costs for the flare gas recovery systems include costs for specialized liquid seals, so there is no need for a flame arrestor. The overall capacity of the flare gas system is the number of compressors times the design flow times the capacity factor per compressor. Therefore, the capacity of the 2-compressor system is 1,440 scfm (2×600 scfm $\times 1.2$). Flare gas system recovery system equipment costs are estimated using Equation 1.27 and the system recovery capacity.

$$C_{FGR} = 731.3 Q_{cap} = 731.3 \times 1440 = \$1,053,072$$

Monitoring system costs are dependent on the applicable standard. Pilot flame monitoring is generally required. For flares that do not routinely have flare gas flow, it may be possible to use engineering calculations to comply with 40 CFR 63.670 and it will be less necessary to have fine steam controls, the only monitoring systems required for the flare with the flare gas recovery system is the pilot flame monitor and a calorimeter. The monitoring system equipment costs, based on the values provided in Table 1.9, are $4,100 + $77,300 = $81,400.

The total equipment cost is the sum of the flare, knock-out drum, transport piping, flare gas recovery system, and monitoring system costs or: $66,626 + $6,524 + $6,797 + $1,053,072 + $81,400 = $1,214,419.

PEC and *TCI* are calculated using the factors given in Table 1.11. In this example, we assume no site preparation or buildings are needed and a contingency factor of 10 percent is appropriate. The calculations are shown in Table 1.16.

$$PEC = 1.18 \times (\$1,214,419) = \$1,433,014$$

$$TCI = (1.1) \times 1.89 \times (\$1,433,014) = \$2,979,236$$

1.7.3 Operating Requirements

Operating labor is estimated at 1,260 hours annually, twice that of a flare without a flare gas recovery system, with supervisory labor at 15% of this amount. Maintenance labor for a flare with a flare gas recovery system is estimated at 1 hour per shift, which translates to 1,095 hours annually. Maintenance material costs are assumed to be equal to maintenance labor costs.

With the flare gas recovery system, there will be no vent gas flow to the flare during the normal, low and no flow conditions. There will be some flow to the flare during high flow condition but the flow discharged to the flare will be reduced based on the flare gas system's recovery capacity. Thus, compliance with applicable requirements in 40 CFR 60.18, 40 CFR 63.11 or 40 CFR 63.670 will only need to consider the high flow condition. Since the heat content of the gas during the high flow condition is above 300 Btu/scf, as indicated in Table 1.13, no auxiliary fuel is needed to comply with the flare requirements in 40 CFR 60.18 or 40 CFR 63.11. If the flare is required to meet the requirements in 40 CFR 63.670, then one must estimate the steam flow rates for the applicable flow scenarios using Equation 1.10 and consider the minimum design steam flow rate for the flare.

$$M_{steam} = 0.4 \times Q \times 60 \times \left(\frac{MWt}{MVC} \right)$$

At high flow conditions with flare gas recovery system running at capacity the flare gas flow rate exceeding the recovery capacity is 1,760 scfm (3,200 – 1,440). At this flow rate, $M_{steam,1} = 0.4 \times (1,760) \times 60 \times 31/385.3 = 3,398$ lb/hr.

Using the average minimum steam supply rate of 46 lbs/hr of steam per inch of flare diameter, an 8-inch diameter flare could have an expected minimum steam flow rate of $46 \times 8 = 368$ lbs/hr steam flow requirement. For this example, we assume the flare has a minimum design steam flow rate so the steam rate while there is full flare gas recovery is 368 lbs/hr.

Mass steam flow rates can be converted to a volumetric steam flow rate, S (scfm), using 18 lb/lb-mol for the molecular weight of water, MWt (lb/lb-mol) and ideal gas law molar volume correction factor, MVC, of 385.3 scf/lb-mol as follows:

$$S = \left(\frac{M_{steam}}{60} \right) \left(\frac{MVC}{MWt} \right)$$

Table 1.16: Capital Cost for Flare Systems - Example Problem 2

Cost Item	Cost[a]
Direct Costs	
Purchased Equipment Costs	
Flare system (Self-supporting)	$66,626
Knock-out drum	$6,524
Vent stream transfer line	$6,797
Flare gas recovery system	$1,053,072
Monitoring system	$81,400
Equipment Cost (Sum = A)	$1,214,419
Instrumentation (0.1A)	$121,442
Sales taxes (0.03A)	$36,433
Freight (0.05A)	$60,721
Purchased equipment cost, PEC	$1,433,014
Direct Installation Costs	
Foundations & Supports (0.12 PEC)	$171,962
Handling & Erection (0.40 PEC)	$573,206
Electrical (0.01 PEC)	$14,330
Piping (0.02 PEC)	$28,660
Insulation (0.01 PEC)	$14,330
Painting (0.01 PEC)	$14,330
Direct Installation Costs	$816,818
Site Preparation	$0
Facilities and Buildings	$0
Total Direct Costs, DC	**$2,249,832**
Indirect Installation Costs	
Engineering (0.1 PEC)	$143,301
Construction and Field Expenses (0.1 PEC)	$143,301
Contractor Fees (0.1 PEC)	$143,301
Start-Up (0.01 PEC)	$14,330
Performance Test (0.01 PEC)	$14,330
Total Indirect Costs, IC	**$458,564**
Contingencies, C (0.1 x (DC + IC)	$270,840
Total Capital Investment = 1.1 × (DC + IC) (rounded)	**$2,979,200**

[a] Costs are in 2017 dollars.

Thus, the volumetric steam flow rates are:

At high flow conditions, $S_1 = (3,398/60)(385.3/18) = 1,212$ scfm.

At typical, low and no flow conditions (with full flare gas recovery), $S_2 = S_3 = S_4 = (368/60)(385.3/18) = 131.3$ scfm.

For flares required to meet the requirements in 40 CFR 63.670, the projected combustion zone NHV must be assessed to determine if the steam rates projected above allow compliance with the 270 Btu/scf combustion zone NHV operating limit. One can simply input the gas flow rates, heat content and steam rates into Equation 1.3 and if the calculated auxiliary fuel rate is negative, then no auxiliary fuel is required. Alternatively, one can calculate the combustion zone NHV assuming no auxiliary fuel is used to see if it meets the required limit, and then calculate the amount of auxiliary fuel needed using Equation 1.3 only for those scenarios where the target is not met. The combustion zone net heating value, NHV_{cz}, when there is no auxiliary fuel is calculated as follows:

$$NHV_{cz} = NHV_{vg}\left(\frac{Q}{Q+S}\right)$$

The combustion zone NHV with no auxiliary fuel for the high flow condition with flare gas recovery is, $NHV_{cz,1} = (750)(1760)/(1760+1212) = 444$ Btu/scf. Therefore, no auxiliary fuel is needed.

Under 40 CFR 63.670, the NHV_{cz} limit does not apply when no "regulated material" is discharged to the flare. Thus, with the flare gas recovery system operating, "no flow" conditions effectively occur for 8,500 hours per year. Consequently, with the flare gas recovery system in-place, it would not be necessary or desirable to meet the NHV_{target} value during times of no waste gas flow. Therefore, no auxiliary fuel is needed to meet the NHV_{target} value during Condition 2, 3, or 4 (with full flare gas recovery).

Natural gas is still required for purge and pilot gas. Purge gas requirements are dependent on the diameter of the flare and are calculated from Equation 1.8.

$$F_{pu} = (7.85 \times 10^{-4})(8,760)(8)^2 = 440.1\frac{Mscf}{yr}$$

Since the flare tip diameter is 8 inches, pilot gas requirements are based on one pilot burner, (see Table 1.3) and are calculated by Equation 1.9. When $N = 1$,

$$F_{pi} = (70)(1)(8,760)(10^{-3}) = 613.2\frac{Mscf}{yr}$$

The total natural gas usage is $440.1 + 613.2 = 1,053$ Mscf/yr.

Flare gas recovery can offset natural gas purchases. The quantity of natural gas purchases that can be offset is determined using Equation 1.19 for each flow condition. The volume of gas recovered during a flow condition is calculated using Equation 1.4. For the high flow condition

(Condition 1), the amount of flare gas recovered is limited by the capacity of the flare gas recovery system installed, which is 1,440 scfm.

$V_{rec,1} = (1440 \times 60 \times 260)/1000 = 22,464$ Mscf

$V_{rec,2} = (600 \times 60 \times 7000)/1000 = 252,000$ Mscf

$V_{rec,3} = (50 \times 60 \times 1000)/1000 = 3,000$ Mscf

$F_{offset,1} = (22,464 \times 750)/920 = 18,313$ Mscf

$F_{offset,2} = (252,000 \times 450)/920 = 123,261$ Mscf

$F_{offset,3} = (3,000 \times 450)/920 = 1,467$ Mscf

$F_{offset,total} = 18,313 + 123,261 + 1,467 = 143,041$ Mscf

This quantity of flare gas recovered may be reduced by the fraction of time flare gas recovery system is operational to account for downtime due to maintenance or other reasons. Because we are using a 2-compressor system designed to operate one compressor at a time for most of the year, we assume the system is operating at all times. Note that this does not mean the system recovers 100 percent of the flared gas. In this example, 27,456 Mscf [(3200-1440)×60×260/1000] of flared gas was not recovered under Condition 1, suggesting the system only recovered approximately 90 percent of all of the gas sent to the flare.

The annual quantity of steam used is calculated based on the steam assist rates and the time in each flow condition.

$$M_{steam} = (3,398)(260) + (368)(7,000) + (368)(1000) + (368)(500)$$

$$= 4,011,500 \frac{lb}{yr} \ (rounded) = 4,011.5 \frac{1,000lb}{yr}$$

The power need associated with running the compressors (mover system) associated with the flare gas recovery system is estimated using Equation 1.18. We apply Equation 1.18 on a per compressor basis because, in this example, only 1 compressor is used at a time for most flow conditions. We sized each compressor for a capacity of 120 percent of the typical flow (Condition 2) or 720 scfm (600 × 1.2). The power need per compressor is:

$$E_{FGR} = 0.0806 \times 720 = 58.032 \ kW/compressor$$

For Condition 1, 2 compressors are used for 260 hours, so the total electrical usage is 30,177 kW-hr (58.032×2×260). For Conditions 2 and 3, only one compressor is needed. The total operating time while running one compressor is 8,000 hrs (7,000+1,000), so the electrical consumption for Conditions 2 and 3 combined is 464,256 kW-hr (58.032×1×8000). We assume the compressor were not running during the no flow condition. Therefore, the total annual electricity consumption is: 30,177 + 464,256 = 494,433 kW-hr.

1.7.4   Total Annual Costs

Table 1.17 shows the calculations of the direct and indirect annual costs for the flare system as calculated from the factors in Table 1.12. Direct costs include labor, materials, and utilities. Indirect costs are the fixed costs allocated to the project, including capital recovery costs and such costs as overhead, insurance, taxes, and administrative charges. The direct annual costs show a credit of $423,000 due to the value of the recovered gas. The sum of the direct and indirect annual costs yields a total annual cost of just under $40,000, indicating that the flare gas recovery system covers most of the cost for the entire flare and flare gas recovery system over the life of the system.

Table 1.17: Annual Costs for Flare System - Example Problem 2

Cost Item	Calculations	Cost[a]
Direct Annual Costs, DAC		
Operating Labor		
Operator	$\dfrac{1,260h}{year} \times \dfrac{\$29.63}{h}$	$ 37,334
Supervisor	15% of operator $= 0.15 \times \$18,667$	$5,600
Operating materials		
Maintenance		
Labor	$\dfrac{1h}{shift} \times \dfrac{shift}{8h} \times \dfrac{8,760h}{yr} \times \dfrac{\$25.12}{h}$	$27,506
Materials	100% of maintenance labor	$27,506
Utilities		
Electricity	$\dfrac{494,433kW\text{-}hr}{yr} \times \dfrac{\$0.0688}{kW\text{-}hr}$	$34,017
Natural gas	$\dfrac{(440 + 613 - 143,041)Mscf}{yr} \times \dfrac{\$4.14}{Mscf}$	-$587,830
Steam	$\dfrac{8,223.3 \times 10^3 lb}{yr} \times \dfrac{\$7.70}{10^3 lb}$	$30,889
Total DAC		-$424,977
Indirect Annual Costs, IAC		
Overhead	60% of total labor and material costs $= 0.6\,(\$37,334 + \$5,600 + \$27,506 + \$27,506)$	$58,768
Administrative charges	2% of Total Capital Investment $= 0.02\,(\$2,979,236)$	$59,585
Property tax	1% of Total Capital Investment $= 0.01\,(\$2,979,236)$	$29,792
Insurance	1% of Total Capital Investment $= 0.01\,(\$2,979,236)$	$29,792
Capital recovery[b]	$0.0963 \times \$2,979,236$	$286,900
Total IAC		$464,838
Total Annual Cost (rounded)		$39,900

[a] Costs are in 2017 dollars.
[b] The capital recovery cost factor, CRF, is a function of the flare equipment life and the opportunity cost of the capital (i.e., interest rate). For this example, a 15-year equipment life and 5% interest rate yields CRF = 0.0963.

1.8   Acknowledgments

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

- Aereon (Austin, TX)

- Airgas USA (Morrisville, NC)

- American National Standards Institute (ANSI) (Washington, DC)

- Applied Instrument Technology (Upland, CA)

- COSA Xentaur Corporation (Yaphank, NY)

- Emerson (St Louis, MO)

- Flaregas Corporation (Spring Valley, NY)

- John Zink Hamworthy Company (Tulsa, OK)

- Kaldair Incorporated (Houston, TX)

- Marathon Petroleum (Findlay, OH)

- NAO Incorporated (Philadelphia, PA)

- Peabody Engineering Corporation (Stamford, CT)

- Piedmont HUB, Incorporated (Raleigh, NC)

- RE Mason (Charlotte, NC)

- Siemens USA (Washington, DC)

- Wasson-ECE Instrumentation (Fort Collins, CO)

- Zeeco (Broken Arrow, OK)

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