# Chapter 7

# FLARES

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## Contents

## 7.1 Introduction

Flaring is a volatile organic compound (VOC) combustion control process in which the VOCs are piped to a remote, usually elevated, location and burned in an open flame in the open air using a specially designed burner tip, auxiliary fuel, and steam or air to promote mixing for nearly complete (> 98%) VOC destruction. Completeness of combustion in a flare is governed by flame temperature, residence time 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 VOCs are converted to carbon dioxide and water. Incomplete combustion results in some of the VOC 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, $SO_x$, $NO_x$, CO, and an additional source of ignition where not desired. However, by proper design these can be minimized.

### 7.1.1 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.

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 air supply and good mixing are required to complete combustion and minimize smoke. The various flare designs differ primarily in their accomplishment of mixing.

#### 7.1.1.1 Steam-Assisted Flares

Steam-assisted flares are single burner tips, elevated above ground level for safety reasons, that burn the vented gas in essentially a diffusion flame. They reportedly account for the majority of the flares installed and are the predominant flare type found in refineries and chemical plants.[1,2]

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 are the focus of the chapter and will be discussed in greater detail in Sections 7.2 through 7.4.

#### 7.1.1.2 Air-Assisted Flares

Some flares use forced air to provide the combustion air and the mixing required for smokeless operation. These flares are built with a spider-shaped burner (with many small gas orifices) located inside but near the top of a steel cylinder two feet or more in diameter. Combustion air is provided by a fan in the bottom of the cylinder. 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. Although air assist is not usually used on large flares (because it is generally not economical when the gas volume is large[3]) the number of large air-assisted flares being built is increasing.[4]

#### 7.1.1.3 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.[5] These streams require less air for complete combustion, have lower combustion temperatures that minimize cracking reactions, and are more resistant to cracking.

#### 7.1.1.4 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 generally (but not necessarily) have the burner arrangement at ground level, and consequently, must be located in a remote area of the plant where there is plenty of space available. They have multiple burner heads that are 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.

#### 7.1.1.5 Enclosed Ground Flares

An enclosed flare's burner heads are inside a shell that is internally insulated. This shell reduces noise, luminosity, and heat radiation and provides wind protection. A high nozzle pressure drop is usually adequate to provide the mixing necessary for smokeless operation and air or steam assist is not required. In this context, enclosed flares can be considered a special class of pressure-assisted or non-assisted flares. The height must be adequate for creating enough draft to supply sufficient air for smokeless combustion and for dispersion of the thermal plume. These flares are always at ground level.

Enclosed flares generally have less capacity than open flares and are used to combust continuous, constant flow vent streams, although reliable and efficient operation can be attained over a wide range of design capacity. Stable combustion can be obtained with lower Btu content vent gases than is possible with open flare designs (50 to 60 Btu/scf has been reported)[2], probably due to their isolation from wind effects. Enclosed flares are typically found at landfills.

### 7.1.2 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 vent 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. These large diameter flares designed to handle emergency releases, can also be used to control vent streams from various process operations. Consideration of vent stream flow rate and available pressure must be given for retrofit applications. 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. Further, if the vent stream pressure is not sufficient to overcome the flare system pressure, then the economics of a gas mover system must be evaluated, If adding the vent stream causes the maximum velocity limits or ground level heat radiation limits to be exceeded, then a retrofit application 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 a flare alone.

Streams containing high concentrations of halogenated or sulfur containing compounds are not usually flared due to corrosion of the flare tip or formation of secondary pollutants (such as $SO_2$). If these vent types are to be controlled by combustion, thermal incineration, followed by scrubbing to remove the acid gases, is the preferred method.[3]

### 7.1.3 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.

#### 7.1.3.1 Factors Affecting Efficiency

The major factors affecting flare combustion efficiency are vent gas flammability, auto-ignition temperature, heating value (Btu/scf), density, and flame zone mixing.

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 vent streams, the heating value also affects flame stability, emissions, and flame structure. A lower 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 vent 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.[5]

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 0.35) have a greater tendency to smoke and require better mixing for smokeless flaring.[3] For this reason one generic steam-to-vent gas ratio is not necessarily appropriate for all vent streams. The required steam rate is dependent on the carbon to hydrogen ratio of the gas being flared. A high ratio requires more steam to prevent a smoking flare.

#### 7.1.3.2 Flare Specifications

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.

The EPA requirements for flares used to comply with EPA air emission standards are specified in 40 CFR Section 60.18. The requirements are for steam-assisted, air-assisted, and non-assisted flares. Requirements for steam-assisted, elevated flares state that the flare shall be designed for and operated with:

- an exit velocity at the flare tip of less than 60 ft/sec for 300 Btu/scf gas streams and less than 400 ft/sec for >1,000 Btu/scf gas streams. For gas streams between 300-1,000 Btu/scf the maximum permitted velocity ($V_{max}$, in ft/sec) is determined by the following equation:

$$\log_{10}(V_{max}) = \frac{B_v + 1214}{852} \tag{7.1}$$

  where $B_v$ is the net heating value in Btu/scf.

- no visible emissions. A five-minute exception period is allowed during any two consecutive hours.

- 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 being 300 Btu/scf or greater.

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

## 7.2 Process Description

The elements of an elevated steam-assisted flare generally consist of gas vent collection piping, 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 7.1

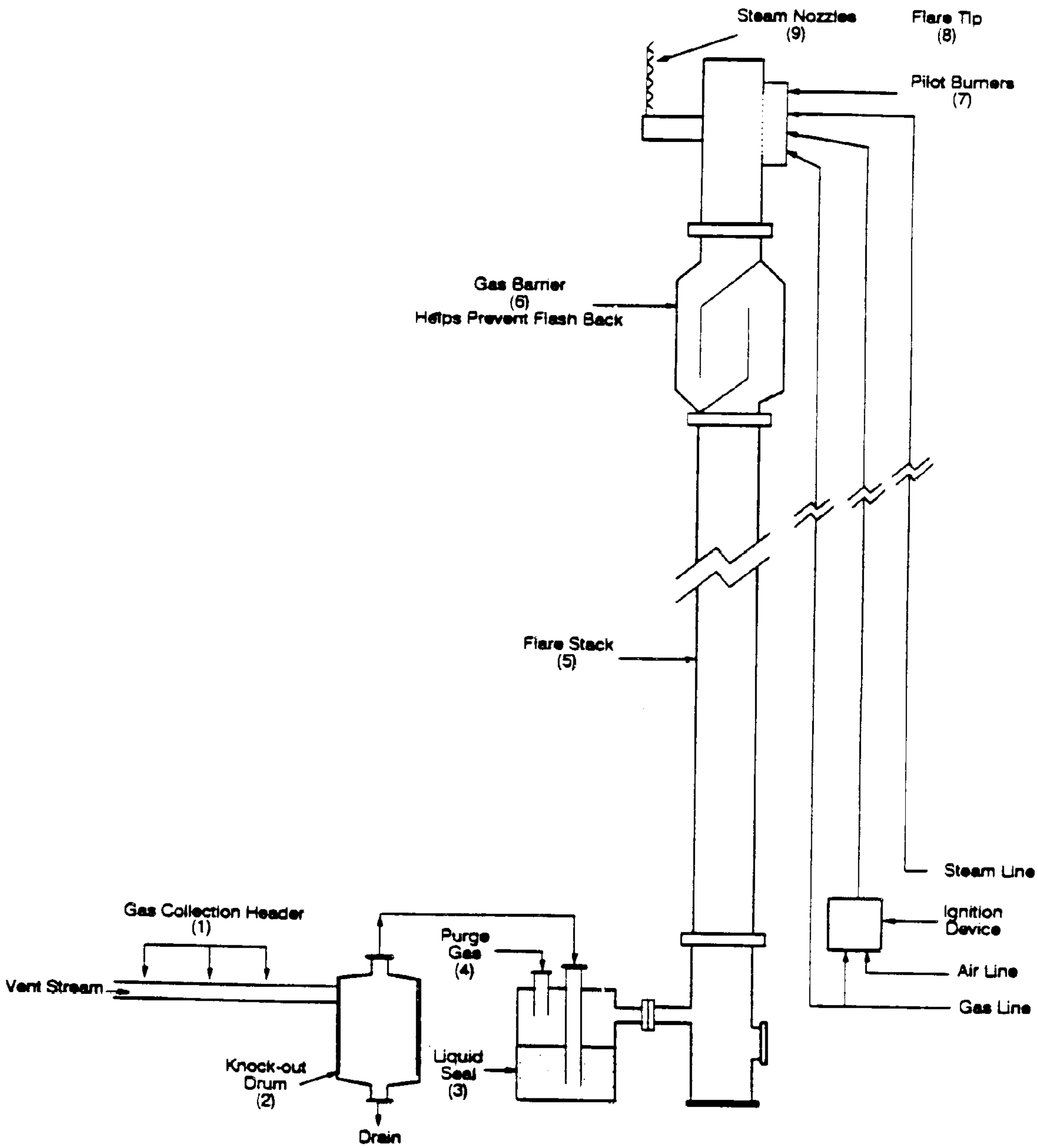


**Figure 7.1:** Steam-Assisted Elevated Flare System

is a diagram of a steam-assisted elevated smokeless flare system showing the usual components that are included.

### 7.2.1 Gas Transport Piping

Process vent streams are sent from the facility release point to the flare location through the gas collection header. 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 purging so that explosive mixtures do not occur in the flare system either on start-up or during operation.

### 7.2.2 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 7.2.) 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 vent gas flow rate with consideration given to liquid entrainment.

### 7.2.3 Liquid Seal

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 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.(51 Other devices, such as flame arresters and check valves, may sometimes replace a liquid seal or be used in conjunction with it. Purge gas (as discussed in Section 7.3.4) also helps to prevent flashback in the flare stack caused by low vent gas flow.

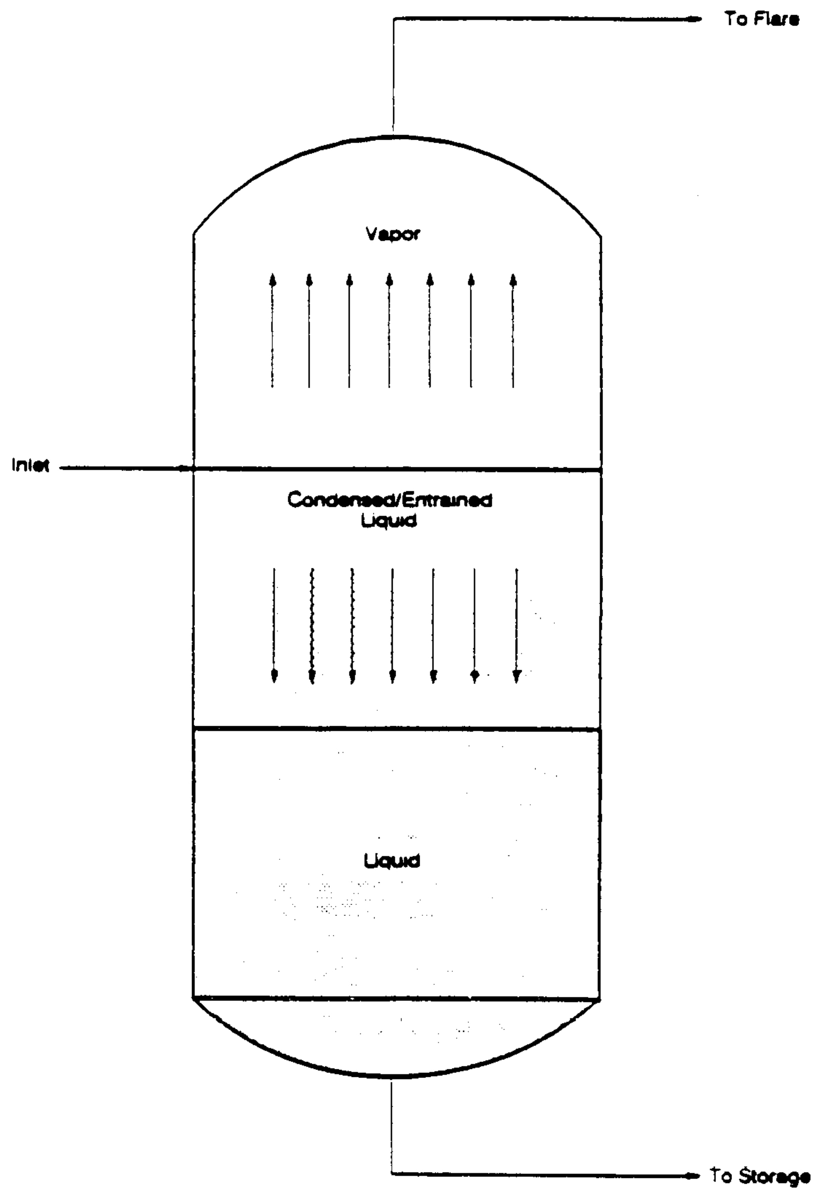


**Figure 7.2:** Typical Vertical Knock-out Drum

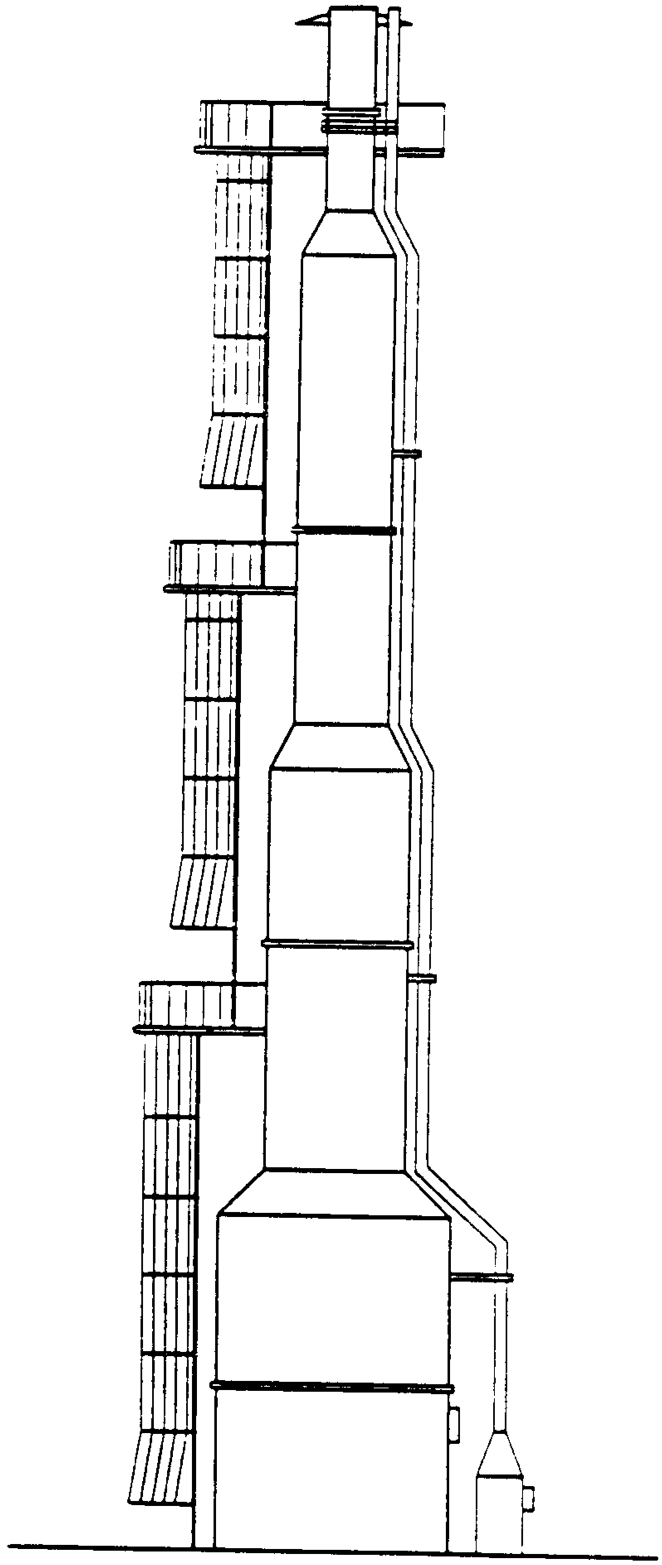

**Figure 7.3:** Self-Supported Elevated Flare

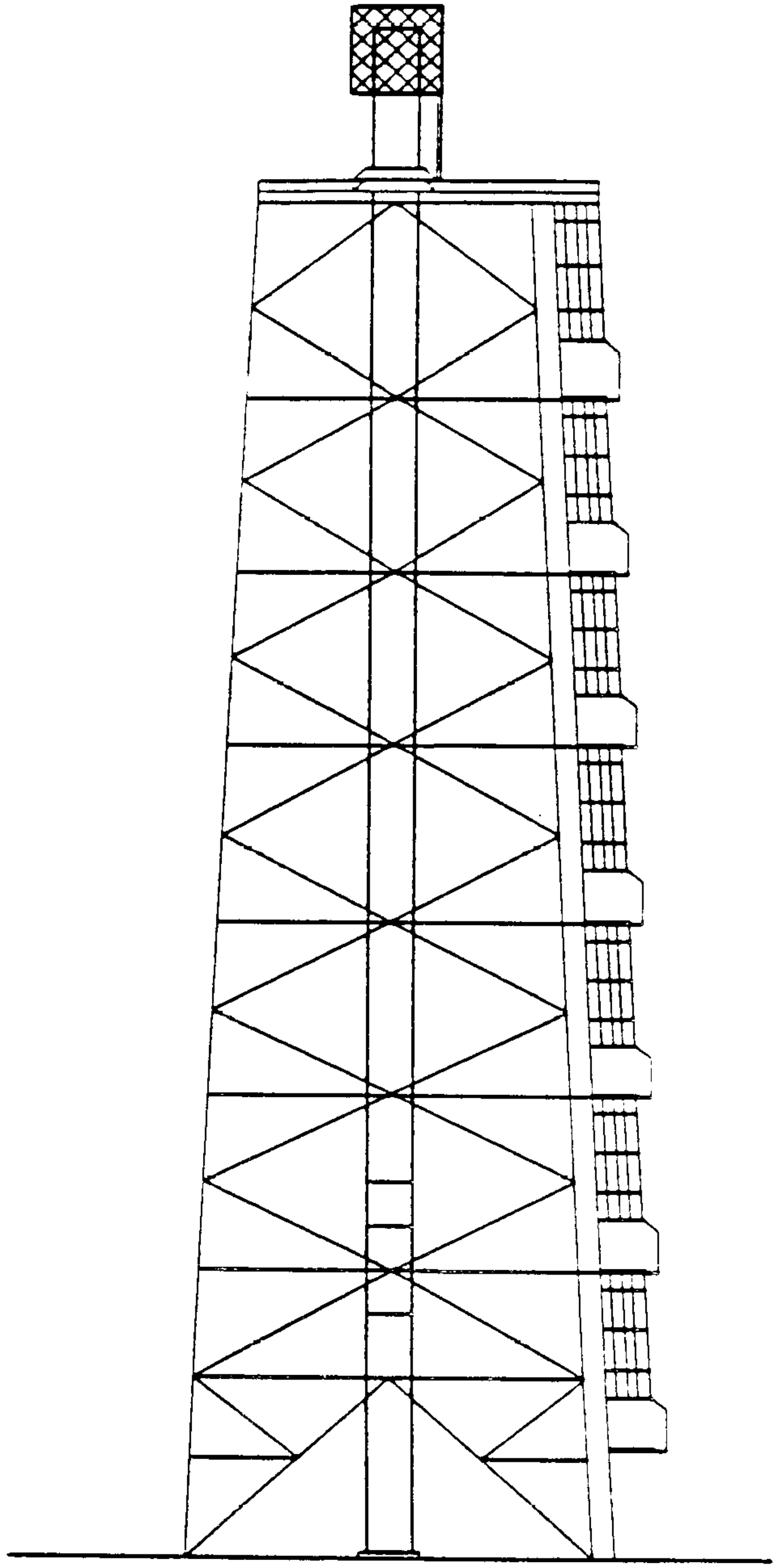

**Figure 7.4:** Derrick-Supported Elevated Flare

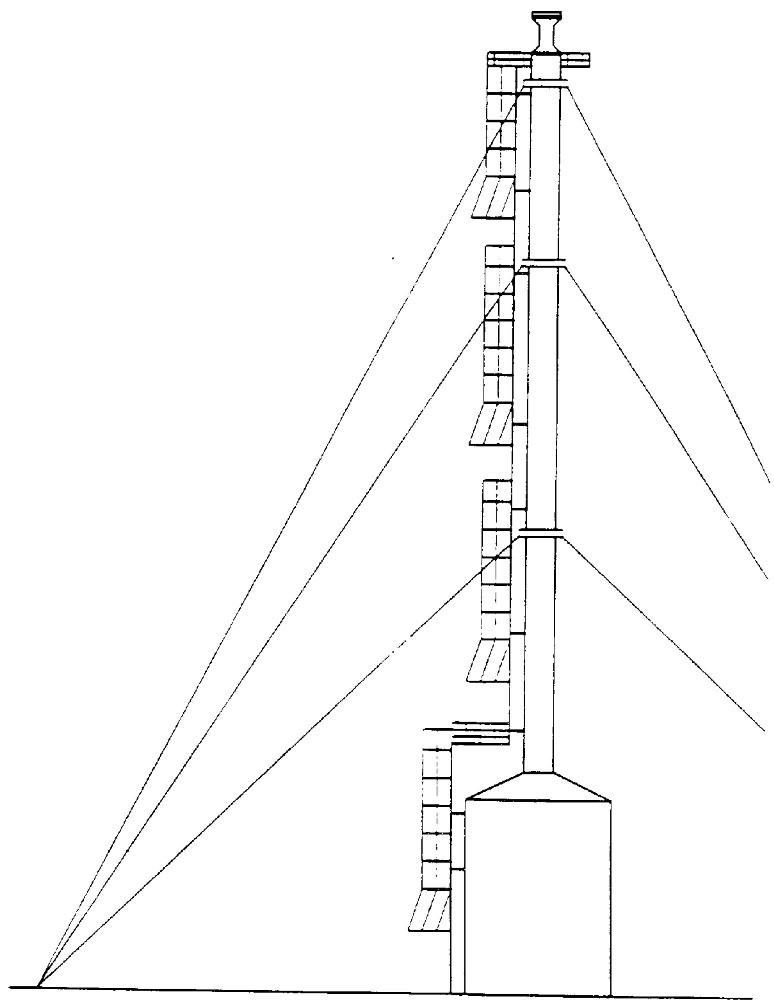

**Figure 7.5:** Guy-Supported Elevated Flare

## 7.2.4 Flare Stack

For safety reasons a stack is used to elevate the flare. The flare must he 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 7.3, 7.4, and 7.5 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 over 300 feet, while derrick towers are designed for above 200 feet.[4, 6, 7, 8, 9, 10]

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.

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.

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.[6]

## 7.2.5 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 prevent this, a gas seal is typically installed in the flare stack. One type of gas seal (also referred to as a flare seal, stack seal, labyrinth seal, or gas barrier) is located below the flare tip to impede the flow of air back into the flare gas network. There are also "seals" which act as orifices in the top of the stack to reduce the purge gas volume for a given velocity and also interfere with the passage of air down the stack from the upper rim. These are known by the names "internal gas seal, fluidic-seal, and arrestor seal".[5] These seals are usually proprietary in design, and their presence reduces the operating purge gas requirements.

## 7.2.6 Burner Tip

The burner tip, or flare 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.[2] The actual maximum capacity of a flare tip is usually limited by the vent stream pressure available to overcome the system pressure drop. Elevated flares 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 CFR 60.18.[l]

### 7.2.7 Pilot Burners

EPA regulations require the presence of a continuous flame. 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 infra-red sensor or, more rarely, (for ground flare applications) an ultra-violet sensor.[4]

### 7.2.8 Steam Jets

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 combustion efficiency. 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.[1] 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.[5] 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. Steam injection is usually 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, 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.

### 7.2.9 Controls

Flare system control can be completely automated or completely 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 heat content (Btu/scf) and automatically adjusting the amount of auxiliary fuel to maintain the required minimum of 300 Btu/scf for steam-assisted flares. Steam consumption can likewise be minimized by controlling flow based on vent gas flow rate. Steam flow can also be controlled using visual smoke monitors. Automatic ignition panels sense the presence of a flame with either visual or thermal sensors and reignite the pilots when flameouts occur.

## 7.3 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. 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.

### 7.3.1 Auxiliary Fuel Requirement

The flare tip diameter is a function of the vent gas flow rate *plus* the auxiliary fuel and purge gas flow rate. The purge gas flow rate is very small relative to the vent gas and fuel flow rates, so it may be ignored when determining the tip diameter. The flow rate of the auxiliary fuel, if required, *is* 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, F, is calculated based on maintaining the vent gas stream net heating value at the minimum of 300 Btu/scf required by rules defined in the *Federal Register* (see next section):

$$Q\,B_v + F\,B_f = (Q + F)(300\ \text{Btu/scf}) \tag{7.2}$$

where

$Q$ = the vent stream flow rate, scfm
$B_v$ and $B_f$ are the Btu/scf of the vent stream and fuel, respectively.

Rearranging gives:

$$F\ (\text{scfm}) = Q\,\frac{300 - B_v}{B_f - 300} \tag{7.3}$$

The annual auxiliary fuel requirement, $F_a$, is calculated by:

$$F_a(\text{Mscf/yr}) = (F\ \text{scfm})(60\text{min/hr})(8760\text{hr/yr}) = 526F \tag{7.4}$$

Typical natural gas has a net heating value of about 1,000 Btn/scf. 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.[2]

### 7.3.2 Flare Tip Diameter

Flare tip diameter is generally sized on a velocity basis, although pressure drop must also be checked. Flare tip sizing for flares used to comply with EPA air emission standards is governed by rules defined in the Federal Register (see 40 CFR 60.18). To comply with these requirements, the maximum velocity of a steam-assisted elevated flare is determined as follows:

| Net Heating Value of Vent Stream $B_v$ (Btu/scf) | Maximum Velocity $V_{max}$ (ft/sec) |
|---|---|
| 300 | 60 |
| 300 – 1,000 | $\log_{10}(V_{max}) = (B_v + 1{,}214)/852$ |
| >1,000 | 400 |

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

$$D_{min}(\text{in}) = 12\sqrt{\frac{\frac{4}{\pi}\ \frac{Q_{tot}}{60\ (\text{sec/min})}}{0.8\ V_{max}}} = 1.95\sqrt{\frac{Q_{tot}}{V_{max}}} \tag{7.5}$$

where

$Q_{tot}$ = $Q + F$ (measured at stream temperature and pressure)

The flare tip diameter, $D$, is the calculated diameter, $D = D_{min}$, 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 90 inches.[5]

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, 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 must be weighed. (Refer to Section 7.3.8 for typical pressure drop relationships.)

## 7.3.3 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.[1,2] 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$^2$.[1] The intensity of solar radiation is in the range of 250-330 Btu/hr-ft$^2$.[1] 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 flare height normally used is 30 feet.[5] Equation (7.6) by Hajek and Ludwig may be used to determine the minimum distance, *L*, required from the center of the flare flame and a point of exposure where thermal radiation must be limited.[1]

$$L^2\ (ft^2) = \frac{fR}{4\ \ K} \tag{7.6}$$

where

|   |   |   |
|---|---|---|
|  | = | fraction of heat intensity transmitted |
| $f$ | = | fraction of heat radiated |
| $R$ | = | net heat release (Btu/hr) |
| $K$ | = | 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, *L*, 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). For estimating purposes, however, assume all of the heat radiated is transmitted (*i.e.*, r = 1). The following is a summary of heat radiated from various gaseous diffusion flames:[1]

| Gas | Flare Tip Diameter (in) | Fraction of Heat Radiated (*f*) |
|---|---|---|
| Hydrogen | <1 | .10 |
| | 1.6 | .11 |
| | 3.3 | 1.6 |
| | 8.0 | 1.5 |
| | 16.0 | 1.7 |
| Butane | <1 | .29 |
| | 1.6 | .29 |
| | 3.3 | .29 |
| | 8.0 | .28 |
| | 16.0 | .30 |
| Methane | <1 | .16 |
| | 1.6 | .16 |
| | 3.3 | .15 |
| Natural Gas | 8.0 | .19 |
| | 16.0 | .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 $f$ = 0.2 will give conservative results.[4] The heat release, $R$, is calculated from the flare gas flow rate, $W$, and the net heating value, $B_v$, as follows:

$$R \text{ (Btu/hr)} = (W \text{ lb/hr})(B_v \text{ Btu/lb}) \tag{7.8}$$

## 7.3.4 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. Purge gas, typically natural gas, $N_2$, or $C0_2$, 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, purge gas injection should be at the farthest upstream point in the flare transport piping.

The minimum continuous purge gas required 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).[6, 7, 8, 9, 10] Using the conservative value of 0.04 ft/sec and knowing the flare diameter (in), the annual purge gas volume, $F_{pu}$, can be calculated:

$$F_{pu}\ (\text{Mscf/yr}) = (0.04\ \text{ft/sec})\left(\frac{\frac{\pi D^2}{4}}{144}\ \text{ft}^2\right)(3{,}600\ \text{sec/hr})(8{,}760\ \text{hr/yr}) \qquad (7.9)$$
$$= 6.88 D^2\ (\text{Mscf/yr})$$

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.[5]

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 purge gas consumption from these uses is assumed to be minor.

### 7.3.5 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 1000 Btu per scf gas) per pilot burner.[6, 7, 8, 9, 10] The number of pilot burners, N, based on flare size is:[6, 7, 8, 9, 10]

| **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:

$$F_{pi}\ (\text{Mscf/yr}) = (70\ \text{scf/hr})(N)(8{,}760\ \text{hr/yr}) = 613\ N \quad (7.10)$$

### 7.3.6 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.[6, 7, 8, 9, 10] 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.[2]

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.[2] For a typical refinery, the average steam requirement is typically 0.25 lb/lb, with this number increasing to 0.5 lb/lb in chemical plants where large quantities of unsaturated hydrocarbons are flared.[10]

For general consideration, the quantity of steam required, *S*, can be assumed to be 0.4 pounds of steam per pound of flare gas, *W*. Using a 0.4 ratio, the amount of steam required is:

$$S\ (\text{lbs/yr}) = 0.4(W\ \text{lb/yr})(8{,}760\ \text{hr/yr}) = 3{,}500(W\ \text{lbs/hr}) \quad (7.11)$$

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.

### 7.3.7 Knock-out Drum

As explained previously, the knock-out drum is 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. It is assumed here that the drum is not sized for emergency releases and that liquid flow is minimal. Flares designed to control continuous vent streams generally have vertical knockout drums, whereas emergency flares typically have horizontal vessels. The procedure described below applies to *vertical* drums exclusively. A typical vertical knock-out drum is presented in Figure 7.2.

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.[1] The dropout velocity, *U*, of a particle in a stream, or the maximum design vapor velocity, is calculated as follows:[11]

$$U\ (\mathrm{ft/sec}) = G\sqrt{\frac{{}_l - {}_v}{{}_v}} \tag{7.12}$$

where

| | | |
|---|---|---|
| G | = | design vapor velocity factor |
| $p_l$ and $p_v$ | = | liquid and vapor densities, lb/ft$^3$ |

Note that in most cases,

$$\frac{{}_l - {}_v}{{}_v} \approx \frac{1}{{}_v}$$

The design vapor velocity factor, *G*, ranges from 0.15 to 0.25 for vertical gravity separators at 85% of flooding.[11]

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

$$A\ (\mathrm{ft}^2) = \frac{Q_a\ \mathrm{ft}^3/\mathrm{min}}{(60\ \mathrm{sec/min})(U\ \mathrm{ft/sec})} \tag{7.12}$$

where $Q_n$ is the vent stream flow in actual ft$^3$/min, or *Q* adjusted to the vent stream temperature and pressure.

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

$$(d_{min}\ (in) = (12\ in/ft)\sqrt{\frac{4}{\ }(A\ ft^2)} \tag{7.13}$$

$$= 13.5\sqrt{A}$$

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

$$d = d_{min} \text{ rounded to the next largest size} \tag{7.14}$$

Some vertical knockout 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.[5]

The vessel thickness, *t*, is determined based on the following:[13]

| **Diameter, *d* (inches)** | **Thickness, *t* (inches)** |
|---|---|
| $d < 36$ | 0.25 |
| $36 \leq d < 72$ | 0.37 |
| $72 \leq d < 108$ | 50.5 |
| $108 \leq d < 144$ | 0.75 |
| $d \geq 144$ | 1.0 |

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.[11] For small volumes of liquid, as in the case of continuous VOC vent control, it is 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:

$$h\ (in) = 3d \tag{7.15}$$

### 7.3.8 Gas Mover System

The total system pressure drop is a function of the available pressure of the vent stream, the design of the various system components, and the flare gas flow rate. 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 following approximate pressure drop relationships:[5]

| | |
|---|---|
| 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. |

Typical total system pressure drop ranges from about 1 to 25 psi.[5]

## 7.4 Estimating Total Capital Investment

The capital costs of a flare system are presented in this section and are based on the design/sizing procedures discussed in Section 7.3. The costs presented are in **March 1990** dollars.*

Total capital investment, TCI, includes the equipment costs, EC, for the flare itself, the cost of auxiliary equipment, 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 vs automatic control) and the number of appurtenances selected, such as knock-out drums, 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.

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

## 7.4.1 Equipment Costs

Flare vendors were asked to provide budget estimates for the spectrum of commercial flare sizes. These quotes [6, 7, 8, 9, 10] were used to develop the equipment cost correlations for flare units, while the cost equations for the auxiliary equipment were based on references [12] and [13] (knock-out drums) and [14] and [15] (piping). The expected accuracy of these costs is ± 30% (*i.e.*, "study" estimates). Keeping in mind the height restrictions discussed in Section 7.2.4, 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, $C_F$ presented in Equations 7.16 through 7.18 are calculated as a function of stack height, $L$ (ft) (30 ft minimum), and tip diameter, $D$ (in), and are based on support type as follows:

<u>Self Support Group:</u>

$$C_F\ (\$) = (78.0 + 9.14D + 0.749L)^2 \tag{7.16}$$

<u>Guy Support Group:</u>

$$C_F\ (\$) = (103 + 8.68D + 0.470L)^2 \tag{7.17}$$

<u>Derrick Support Group:</u>

$$C_F\ (\$) = (76.4 + 2.72D + 1.64L)^2 \tag{7.18}$$

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 310 stainless steel.

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 the same diameter as the flare tip[6] 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, $C_p$, are presented separately in Equation 7.19 or 7.20 and are a function of pipe, or flare, diameter $D$.[15]

$$C_p\ (\$) = 127D^{1.21}\ (\text{where } 1'' < D < 24'') \tag{7.19}$$

$$C_p\ (\$) = 139D^{1.07}\ (\text{where } 30'' < D < 60'') \tag{7.20}$$

The costs, $C_p$, include straight, Schedule 40, carbon steel pipe only, are based on 100 feet of piping, and are directly proportional to the distance required.

The costs for a knock-out drum, $C_K$, are presented separately in Equation 7.21 and are a function of drum diameter, $d$ (in), and height, $h$ (in).[12, 13]

$$C_K\ (\$) = 14.2\left[dt\ (h + 0.812d)\right]^{0.737} \tag{7.21}$$

where $t$ is the vessel thickness, in inches, determined based on the diameter.

Flare system equipment cost, EC, is the total of the calculated flare, knock-out drum, and piping costs.

$$EC\ (\$) = C_F + C_K + C_p \tag{7.22}$$

Purchased equipment costs, PEC, is equal to equipment cost, EC, plus factors for ancillary instrumentation (*i.e.*, control room instruments) (.10), sales taxes (0.03), and freight (0.05) or,

$$\text{PEC}\ (\$) = \text{EC}\ (1 + 0.10 + 0.03 + 0.05) = 1.18\ \text{EC} \tag{7.23}$$

## 7.4.2 Installation Costs

The total capital investment, TCI, is obtained by multiplying the purchased equipment cost, PEC, by an installation factor of 1.92.

$$\text{TCI}\ (\$) = 1.92\ \text{PEC} \tag{7.24}$$

These costs were determined based on the factors in Table 7.1. The bases used in calculating annual cost factors are given in Table 7.2. These factors encompass direct and indirect 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, performance testing, and contingencies. 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 .[16]

## 7.5 Estimating Total Annual Costs

The total annual cost, TAC, is the sum of the direct and indirect annual costs. The bases used in calculating annual cost factors are given in Table 7.2

### 7.5.1 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 is estimated at 630 hours annually.[3] A completely manual system could easily require 1,000 hours. A standard supervision ratio of 0.15 should be assumed.

Maintenance labor is estimated at 0.5 hours per 8-hour shift. Maintenance materials costs are assumed to equal maintenance labor 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, $C_{pu}$:

$$C_f\ (\$/yr) = C_{pi} + C_a + C_{pu} \tag{7.25}$$

where, $C_{pi}$ is equal to the annual volume of pilot gas, $F_{pi}$, multiplied by the cost per scf, *i.e.*:

$$C_{pi}\ (\$/yr) = (F_{pi}\ scf/yr)(\$/scf) \tag{7.26}$$

$C_a$ and $C_{pu}$ are similarly calculated.

Steam cost ($C_s$) to eliminate smoking is equal to the annual steam consumption 8,760 *S* multiplied by the cost per lb, *i.e.*:

$$C_s\ (\$/yr) = (8{,}760\ hr/yr)(S\ lb/hr)(\$/lb) \tag{7.27}$$

The use of steam as a smoke suppressant can represent as much as 90% or more of the total direct annual costs.

### 7.5.2 Indirect Annual Costs

The indirect (fixed) annual costs include overhead, capital recovery, administrative (G & A) charges, property taxes, and insurance. Suggested indirect annual cost factors are presented in Table 7.2.

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

**Table 7.1:** 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 | 0.12 B |
| 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 Annual Costs, DC** | |
| 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 |
| Contingencies | 0.03 B |
| Total Indirect Costs, IC | 0.35 B |
| | |
| Total Capital Investment = DC + IC | 1.92 B + SP + Bldg. |

The system capital recovery cost, CRC, is based on an estimated 15-year equipment life. (See 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 7%, the capital recovery factor is 0.1098. The system capital recovery cost is the product of the system capital recovery factor, CRF, and the total capital investment, TCI, or:

$$CRC\ (\$/yr) = CRF \times TCI = 0.1098 \times TCI \tag{7.28}$$

As shown in Table 7.2, G & A, taxes, and insurance can be estimated at 2%, 1%, and 1% of the total capital investment, TCI, respectively.

# 7.6 Example Problem

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

## 7.6.1 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

Vent stream parameters and cost data to be used in this example problem are listed in Table 7.3.

**Table 7.2:** Suggested Annual Cost Factors for Flare Systems

| Cost Item | Factor |
|---|---|
| Direct Annual Costs, DC | |
| Operating labor{3} | |
| Operator | 630 man-hours/year |
| Supervisor | 15% of operator |
| Operating materials | — |
| Maintenance | |
| Labor | ½ hour per shift |
| Material | 100% of maintenance labor |
| Utilities | |
| Electricity | All utilities equal to: |
| Purge gas | (Consumption rate) x |
| Pilot gas | (Hours/yr) x (unit cost) |
| Auxiliary fuel | |
| 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[a] | 0.1315 x Total Capital Investment |
| Total Annual Cost | Sum of Direct and Indirect Annual Costs |

[a]See Chapter 2.

**Table 7.3:** Example Problem Data

| | |
|---|---|
| Vent Stream Parameters | |
| Flow rate | 63.4 acfm[a] |
| | 399.3 lb/hr |
| Heat content | 449 Btu/scf[b] |
| System pressure | 10 psig[c] |
| Temperature | 90 $^oF$ |
| Liquid density[17] | 49.60 $lb/ft^3$[d] |
| Vapor density[17] | 0.08446 $lb/ft^3$[d] |
| Cost Data (March 1990)[18,19] | |
| Operating hours | 8,760 hrs/yr |
| Natural gas | 3.03 $/1000 scf |
| Steam | 4.65 $/1000 lbs |
| Operating labor | 15.64 $/hr |
| Maintenance labor | 17.21 $/hr |

[a]Measured at *flare tip*. Flow rate has been adjusted to account for drop in pressure from 10 psig at source to 1 psig at flare tip.
[b]Standard conditions: 77$^oF$, 1 atmosphere.
[c]Pressure at source (gas collection point). Pressure at flare tip is lower: 1 psig.
[d]Measured at standard conditions.

## 7.6.2 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 7.3, 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 7.4.1.

### 7.6.2.1 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 Federal Register requirements. Since the heating value is in the range of 300 to 1,000 Btu/scf, the maximum velocity, $V_{max}$, is determined by Equation 7.1.

$$\log_{10} V_{max} = \frac{449 \text{ Btu/scf} + 1{,}214}{852} = 1.95$$

so,

$$V_{max} = 89.5 \text{ ft/sec}$$

Because the stream heating value is above 300 Btu/scf, no auxiliary fuel is required. Hence, $Q_{tot}$ equals the vent stream flow rate. Based on $Q_{tot}$ and $V_{max}$, the flare tip diameter can be calculated using Equation 7.5.

$$D_{min} = 1.95 \sqrt{\frac{Q_{tot}}{V_{max}}} = 1.95 \sqrt{\frac{63.4 \text{ acfm}}{89.5 \text{ ft/sec}}} = 1.64 \text{ in}$$

The next largest commercially available standard size of 2 inches should he selected for *D*.

The next parameter to determine is the required height of the flare stack. The heat release from the flare is calculated using Equation 7.7.

$$R \text{ (Btu/hr)} = (W \text{ lb/hr})(B_v \text{ Btu/lb})$$

First the heat of combustion, or heating value, must be converted from Btu/scf to Btu/lb. The vapor density of the vent stream at standard temperature and pressure is 0.08446 lb/scf.

So,

$$B_v = \frac{449 \text{ Btu/scf}}{0.08446 \text{ lb/scf}} = 5316 \text{ Btu/lb}$$

and,

$$R = (399.3 \text{ lb/hr})(5{,}316 \text{ Btu/lb}) = 2{,}123{,}000 \text{ Btu/hr}$$

Substituting *R* and appropriate values for other variables into Equation 7.6:

$$\begin{aligned} L^2 \, (\text{ft}^2) &= \frac{fR}{4 \ K} \\ &= \frac{(1)(0.2)(2{,}123{,}000 \text{ Btu/hr})}{4 \ (500 \text{ Btu/hr-ft}^2)} \\ &= 68 \text{ ft}^2 \end{aligned}$$

gives a height of $L$ = 8.2 ft. The smallest commercially available flare is 30 feet, so $L$ = 30 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 7.11 yields a *maximum* velocity of:

$$\begin{aligned} U &= G\sqrt{\frac{{}_l - {}_v}{{}_v}}, \text{ ft/sec} \\ &= 0.20\sqrt{\frac{49.60 - 0.08446}{0.08446}} \\ &= 4.84 \text{ ft/sec} \end{aligned}$$

Given a vent gas flow rate of 63.4 scfm, the *minimum* vessel cross-sectional, diameter is calculated by Equation 7.12:

$$
\begin{aligned}
A &= \frac{Q_a \text{ acfm}}{(60 \text{ sec/min})(U \text{ ft/sec})} \\
&= \frac{63.4}{(60)(4.84)} \\
&= 0.218 \text{ ft}^2
\end{aligned}
$$

This results in a minimum vessel diameter of:

$$
\begin{aligned}
d_{\min} &= 13.5\sqrt{A} \\
&= 13.5\sqrt{0.218} \\
&= 6.3 \text{ inches}
\end{aligned}
$$

The selected diameter, *d*, rounded to the next largest 6 inches is 12 inches. Using the rule of the height to diameter ratio of three gives a vessel height of 36 inches, or 3 feet.

#### 7.6.2.2 Equipment Costs

Once the required flare tip diameter and stack height have been determined the equipment costs can be calculated. Since the height is 30 feet, the flare will be self-supporting. The costs are determined from Equation 7.16.

$$
\begin{aligned}
C_F &= (78.0 + 9.14D + 0.749L)^2 \\
&= [78.0 + 9.14(2 \text{ inches}) + 0.749(30 \text{ ft})]^2 \\
&= \$14,100
\end{aligned}
$$

Knock-out drum costs are determined using Equation 7.21, where *t* is determined from the ranges Presented in Section 7.3.7. Substituting 0.25 for *t*:

$$
\begin{aligned}
C_K &= 14.2[dt\ (h + 0.812d)]^{0.737} \\
&= 14.2[(12)(0.25)(36 + 0.812(12))]^{0.737} \\
&= \$530
\end{aligned}
$$

Transport piping costs are determined using Equation 7.19.

$$
\begin{aligned}
C_p &= 127D^{1.21} \\
&= 127(2)^{1.21} \\
&= \$290
\end{aligned}
$$

The total auxiliary equipment cost is the sum of the knock-out drum and transport piping costs, or \$530 + \$290 = \$820.

The total capital investment is calculated using the factors given in Table 7.1. The calculations are shown in Table 7.4. Therefore:

$$\text{Purchased Equipment Cost} = \text{"B"} = 1.18 \times A$$
$$= 1.18 \times (14{,}920) = \$17{,}610$$

And:

$$\text{Total Capital Investment (rounded)} = 1.92 \times B$$
$$= 1.92 \times (17{,}610) = \$33{,}800.$$

**Table 7.4:** Capital Costs for Flare Systems
Example Problem

| Cost Item | Cost |
|---|---|
| Direct Costs | |
| Purchased equipment costs | |
| Flare (self supporting) | $14,100 |
| Auxiliary equipment[a] | 820 |
| Sum = A | $14,920 |
| Instrumentation, 0.1A | 1,490 |
| Sales taxes, 0.03A | 450 |
| Freight, 0.05A | 750 |
| Purchased equipment cost, B | $17,610 |
| Direct installation costs | |
| Foundation and supports, 0.12B | 2,110 |
| Handling & erection, 0.40B | 7,040 |
| Electrical, 0.01B | 180 |
| Piping, 0.02B | 350 |
| Insulation, 0.01B | 180 |
| Painting, 0.01B | 180 |
| Direct installation cost | $10,040 |
| Site preparation | — |
| Facilities and buildings | — |
| Total Direct Cost | $27,650 |
| Indirect Costs (installation) | |
| Engineering, 0.10B | 1,760 |
| Construction and field expenses, 0.10B | 1,760 |
| Contractor fees, 0.10B | 1,760 |
| Start-up, 0.01B | 180 |
| Performance test, 0.01B | 180 |
| Contingencies, 0.03B | 530 |
| Total Indirect Cost | $6,170 |
| Total Capital Investment (rounded) | $33,800 |

[a]Includes costs for knock-out drum and transport piping.

### 7.6.3 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. Maintenance material costs are assumed to be equal to maintenance labor costs.

As stated above, since the heat content of the example stream is above 300 Btu/scf (*i.e.*, 449 Btu/scf) no auxiliary fuel is needed. Natural gas is required, however, for purge and pilot gas. Purge gas requirements are calculated from Equation 7.8.

$$F_{pu} = 6.88D^2 = 6.88(2)^2 = 27.5 \text{ Mscf/yr}$$

Since the flare tip diameter is less than 10 inches, pilot gas requirements are based on one pilot burner, (see Section 7.3.5) and are calculated by Equation 7.9.

$$F_{pi} = 613N$$

When N = 1,

$$F_{pi} = 613 \text{ Mscf/yr}$$

Steam requirements are calculated from Equation 7.10:

$$S \text{ (lb/yr)} = 3{,}500\, W$$

Inserting the methanol mass flow rate of 399.3 lb/hr yields:

$$\begin{aligned} S &= (3{,}500)(399.3 \text{ lb/hr}) \\ &= 1{,}400 \text{ Mlb/yr} \end{aligned}$$

### 7.6.4 Total Annual Costs

The sum of the direct and indirect annual costs yields a total annual cost of $61,800. Table 7.5 shows the calculations of the direct and indirect annual costs for the flare system as calculated from the factors in Table 7.2. 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.

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.

**Table 7.5:** Annual Costs for Flare System
Example problem

| Cost Item | Calculations | Cost |
|---|---|---|
| Direct Annual Costs, DC | | |
| Operating Labor | | |
| Operator | $\frac{630\ h}{year} \times \frac{\$15.64}{h}$ | $ 9,850 |
| Supervisor | 15% of operator = 0.15 x 9,850 | 1,480 |
| Operating materials | | ---- |
| Maintenance | | |
| Labor | $\frac{0.5\ h}{shift} \times \frac{shift}{8\ h} \times \frac{8{,}760h}{yr} \times \frac{\$17.21}{h}$ | 9,420 |
| Material | 100% of maintenance labor | 9,420 |
| Utilities | | |
| Electricity | | ---- |
| Purge gas | $\frac{27.5\ Mscf}{yr} \times \frac{\$3.03}{Mscf}$ | 80 |
| Pilot gas | $\frac{613\ Mscf}{yr} \times \frac{\$3.03}{Mscf}$ | 1,860 |
| Steam | $\frac{1{,}400 \times 10^3 lb}{yr} \times \frac{\$4.65}{10^3 lb}$ | 6,510 |
| Total DC (rounded) | | $38,600 |
| Indirect Annual Costs, IC | | |
| Overhead | 60% of total labor and material costs = 0.6(9,850 + 1,480 + 9,420 + 9,420) | 18,100 |
| Administrative charges | 2% of Total Capital Investment = 0.02 ($33,800) | 680 |
| Property tax | 1% of Total Capital Investment = 0.01 ($33,800) | 340 |
| Insurance | 1% of Total Capital Investment = 0.01 ($33,800) | 340 |
| Capital recovery[a] | 0.1098 x $33,800 | 3,710 |
| Total IC (rounded) | | 23,200 |
| Total Annual Cost (rounded) | | $61,800 |

[a]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 example, for a 15 year equipment life and 7% interest rate, CRF = 0.1098.

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- NAO Incorporated (Philadelphia, PA)
- Peabody Engineering Corporation (Stamford, CT)
- Piedmont HUB, Incorporated (Raleigh, NC)

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