PRESSURE SAFETY DESIGN PRACTICES FOR REFINERY AND CHEMICAL OPERATIONS

by

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PREFACE

Pressure Safety Design Practices for Refinery and Chemical Operations is an introductory reference guide for pressure relieving operations in petroleum refineries and chemical plants. Petroleum refineries and their related chemical manufacturing facilities often operate at high pressures and/or at conditions that can result in excessive pressure conditions. Start-up and emergency shut-down operations, transient upset conditions, or catastrophic equipment failures, as well as plant process fires and plant wide utility failures, all require fail safe systems that will alleviate high process operating conditions in a safe and expedient manner. In addition, there are many process waste streams that often require pressure let-down and proper venting before being released to the atmosphere in an environ-mentally safe manner. This volume provides general design concepts and engineering practices for pressure relieving operations, and the application of gas flaring practices, which is the controlled burning of product and waste stream releases during depressuring operations. Although the design guidelines presented in this volume are specific to petroleum refinery operations, their general application to other processes is acceptable, provided local industry and site specific codes and restrictions are applied.

The design practices compiled in this volume were derived from the literature of the American Society of Testing Materials (ASTM), the American Petroleum Institute (API), the ANSI Codes, a review of company specific literature including company design practices and general publications, and the author's personal notes compiled over the years as a consultant and practicing engineer, as well as an instructor to Venezuelan refinery and chemical plant operations.

The methodologies and in particular specific installation details, design procedures and calculation methods presented herein are not necessarily definitive, and may differ among refinery operations even though they, for the most part, are based on API recommendations and other well recognized organizations.

Also, refinery operations and particularly chemical plant processes tend to be unique, so that techniques described in this book do not necessarily apply in all cases. The presentation of material is therefore not meant to be an endorsement, either by the author or publisher, nor are any guarantees, either explicit or implicit made for designs based on the information provided in this book.

It is hoped that this volume will provide some guidance to operators, plant safety engineers and designers, and that some of the design practices are adopted particularly in older plant operations that do not apply proper pressure let down procedures and flaring practices. Many unsafe operations and even catastrophic failures have been observed by the author in overseas assignments to such countries as Ukraine and other Newly Independent States (NIS) in older plant designs. These incidents have prompted the preparation of this reference.

NOTICE

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INTRODUCTION TO PETROLEUM REFINERY OPERATIONS

INTRODUCTION

A petroleum refinery is a manufacturing operation where crude petroleum, the raw material, is converted into usable finished products. In other words, it is the manufacturing phase of the oil industry. This chapter presents a general introduction to overall refinery operations as a forerunner to the detailed information on specific processes and products which follows, and the technologies that are applied to pressure relief operations. Other chapters cover the major operations and processes used in refining, and discuss the critical properties and end uses of the products. However, it should be emphasized that a refinery is only one of the major phases of the petroleum industry; others being exploration, production, transportation, and marketing, and a variety of feedstock chemicals that supply the raw materials for various product lines. Research and engineering might also be listed, but they are, in reality, a necessary and integral part of each of the phases.

REFINERY OPERATIONS

The function of the refinery is to convert crude oil into the finished products required by the market in the most efficient, and hence most profitable manner.

The methods employed necessarily vary widely from one refinery to another, depending on the crude processed, the nature and location of the market, the type of equipment available, and many other factors. However, for simplification, it may be considered that all refining processes fall into one of four basic categories.

The first category is fractionation or distillation. This method of physically separating a mixture of compounds was the earliest process used in petroleum refining, and today is still one of the most important. However, since it is not generally possible to separate the complex petroleum mixtures into individual compounds, such mixtures are segregated into fractions or "cuts", each of which is characterized by a carefully controlled boiling range. These cuts are then further processed or utilized in the refinery operations.

The second basic type of process, essentially chemical in nature, consists of converting or chemically transforming certain of these "cuts" into products of higher commercial value. There are many ways of doing this, but all consist fundamentally of altering the molecular structure of the components. In the case of a heavy oil, the molecules may be cracked to form lighter, more valuable products, as for instance in catalytic cracking and coking. On the other hand, gaseous products may be polymerized or otherwise combined to form liquid products which may be blended into gasoline. With certain processes, e.g. catalytic reforming, both cracking and polymerization take place concurrently with the more desirable de-hydrogenation, hydrogenation, and isomerization reactions. The net result of all these transformations is the production of mixtures containing new arrays of hydrocarbons of higher value than the starting materials.

Nearly all the fractions produced by the processes mentioned above contain certain objectionable constituents or impurities. The third basic category is, therefore, treating. This group of processes includes the removal of the unwanted components, or their conversion to innocuous or less undesirable compounds. Removal of the impurities is sometimes accomplished by physical treating, as exemplified by the process for manufacturing kerosene, wherein sulfur and certain undesirable hydrocarbons are removed by extraction with liquid sulfur dioxide. Alternatively, the removal may be carried out by converting the unwanted compounds to a form more readily removed as is done in the hydrodesulfurization of diesel fuel. Here the sulfur compounds are cracked and hydrogenated. The sulfur is converted to hydrogen sulfide which can be readily separated from the heavier diesel oil by fractionation. An example of the conversion of undesirable components to innocuous compounds which remain in the product is found in the gasoline sweetening processes. There the mercaptans present give the product a foul, objectionable odor. The sweetening process merely transforms the mercaptans to organic disulfides which are less objectionable.

Although sulfur is perhaps the commonest and most troublesome of the impurities found in petroleum, it is certainly not the only one. Substances such as nickel, vanadium, and nitrogen may also be present in the crude oil. These impurities are undesirable because of the difficulty they cause during processing in the refinery or because of some detrimental effect during consumer use of the product. Furthermore, presence of certain hydrocarbons or certain types of hydrocarbons may lower the quality of a specific product. It was mentioned that aromatics are removed from kerosene by SO_2 extraction. The aromatics have undesirable burning characteristics and hence the product quality is improved if these "impurities" are removed. Lube oil treating process such as dewaxing, deasphalting, and phenol treating also fall into this category.

The fourth basic category is blending of the finished cuts into commercially saleable products such as motor gasoline, kerosene, lubricating oils, and bunker fuel oil, according to their specifications.

These four basic categories encompass the fundamental operation of a refinery. All other activities are carried out to implement them. The specifications for a given product are established to insure a satisfactory level of product performance. Specifications can be altered from time to time, but a product normally must meet the then existing product specifications. Various crudes on the other hand yield fractions with significantly different properties. At first glance, it might appear reasonable to select crudes to best match the product needs of each refinery. Many times, however, this is not economical as the money saved in eliminating various conversion and treating processes is offset by other factors. These might include crude availability, price, and transportation or specialty product requirements. A refinery is a sophisticated multi-component process operated in overall balance. The balance is set by economic considerations with the major variables being crude oil, process costs, and final products. It is thus easier to see why (1) no two refineries are exactly alike, (2) various conversion and purification processes are required, and (3) crude selection is important.

TYPES OF REFINERIES

Each refinery is designed to manufacture products as economically as possible based on the best knowledge available with regard to end product needs, future expansion plans, crude availability and other pertinent factors.

A basic modern refinery which does not produce lubricating oils or chemicals is commonly referred to as a fuel products refinery. It is designed to produce primarily motor gasoline, distillate fuels (diesel oil, jet fuel, and heating oil), and bunker (residual) fuel oil. The fuel products refineries can be considered basic and minimum as regards refinery product and processing requirements. Hydroskimming and conversion are the two major variations of this type refinery. There is a wide range of conversion levels. The term maximum conversion type has no precise definition but is often used to describe a level of conversion , where there is no net fuel oil manufacture. A fuel products refinery with specialities may manufacture lubricating oils, asphalts, greases, solvents, waxes and chemical feed stocks in addition to the primary fuel products. The number and diversity of products will naturally vary from one refinery to another.

Refineries produce chemical feed stocks for sale to the chemical affiliates and do not have responsibility for the manufacture of chemical products directly. Both operations may be carried out at the same physical location but the corporate product responsibilities are usually separate.

FUEL PRODUCTS REFINERY

Hydroskimmer

A hydroskimming refinery lends itself to locations where the market demands for the major fuel products (gasoline, gas oil, and residual fuel oil) approximate the quantities of these products obtainable by distillation from the available crudes. A typical hydroskimming refinery would include the following:

1. Atmospheric Pipestill
2. Powerforming (Catalytic Naphtha Reforming)
3. Light Ends Recovery - Fractionation
4. Treating and Blending

Figure 1 shows a simplified flow plan for a typical hydroskimming refinery. The atmospheric pipestill performs the initial distillation of crude oil into gas, naphtha, distillates, and residuum. The naphtha may be separated into gasoline blending stock, solvents, and Powerformer feed. The distillates include kerosene, jet fuel, heating oil and diesel oil. The residuum is blended for use as bunker fuel oil.

The Powerforming unit is required to upgrade virgin naphtha to produce high octane gasoline. Powerforming is a fixed bed catalytic reforming process employing a regenerable platinum catalyst. In the process, a series of reactions

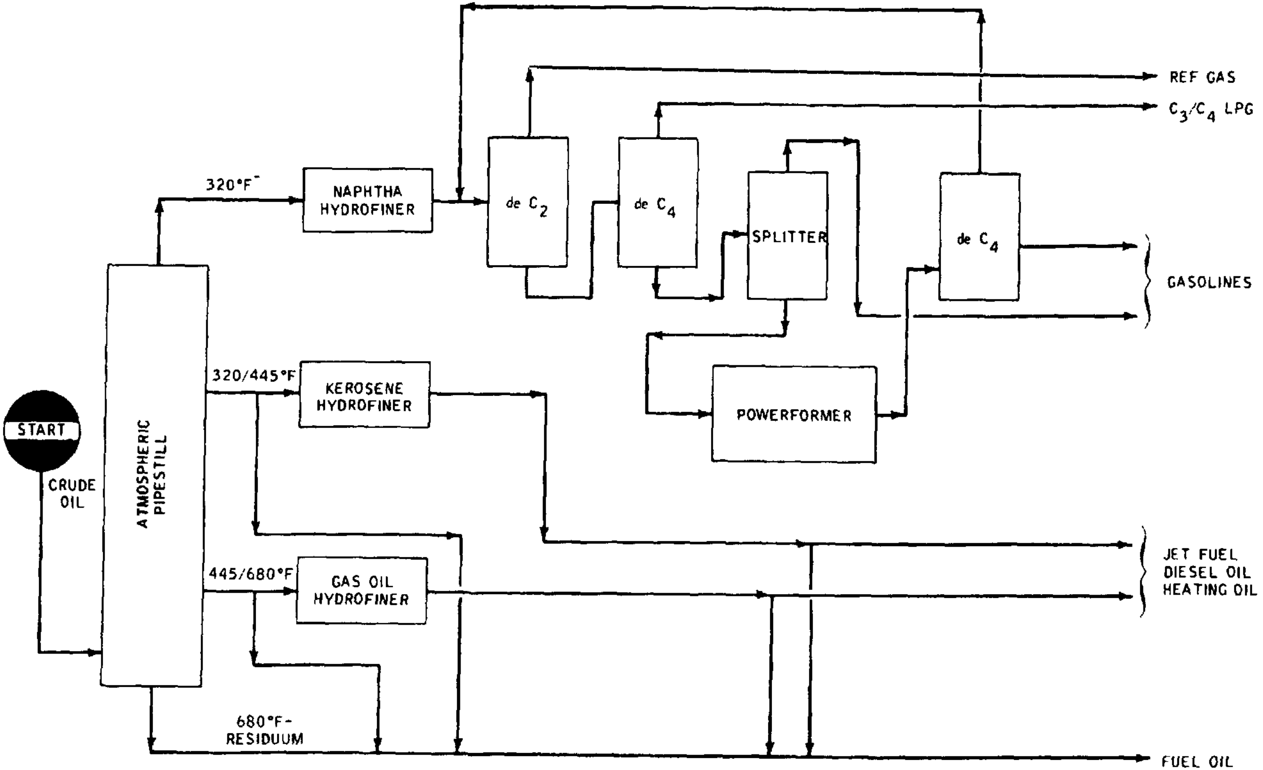

Figure 1. Flowsheet for a hydroskimming refinery operation.

takes place. The most important of these is aromatization; other reactions include isomerization, cracking, hydrogenation, and polymerization. The desired product is of approximately the same boiling range as the feed, but the molecules have been rearranged or reformed into higher octane compounds.

Light ends recovery and fractionating equipment is necessary after the Powerformer and on the pipestill overhead stream to separate the effluent mixtures into the desired boiling range cuts.

Hydrofining is used to reduce sulfur and/or other impurities and to improve odor, color, and stability of the pipestill fractions. Hydrofining is a fixed-bed catalytic process using a regenerable cobalt molybdate catalyst in a hydrogen atmosphere. The hydrogen is produced by the Powerformer with supplemental hydrogen manufactured if necessary. The difficulty of hydrofining (desulfurization) increases with increase in the hydrocarbon boiling point. Naphthas are generally desulfurized up to 99+% by hydrofining while the maximum desulfurization of distillates is usually 90 %.

The components produced by the process sequence outlined above are blended as required to meet final product rates and qualities.

Conversion

The hydroskimming type refinery is used where the gasoline demand is substantially lower and hence the final product demand is close to that yielded by single stage distillation. In areas where the demand for gasoline is relatively high, conversion processing is required. The minimum processes for a fuel products refinery designed would typically include:

1. Atmospheric and Vacuum Crude Distillation
2. Catalytic Gas Oil Cracking
3. Powerforming
4. Light Ends Recovery - Fractionation
5. Treating and Blending

Figure 2 shows a simplified flow plan for a typical conversion type refinery. The atmospheric P/S residuum can be fed to a vacuum pipestill. The vacuum tower enables the refiner to cut deeper into the crude, at the same time avoiding high temperatures (above about 750 °F) which cause thermal cracking with resultant deposition of coke and tarry residues in the equipment.

The vacuum gas oil produced by vacuum distillation is fed to a catalytic

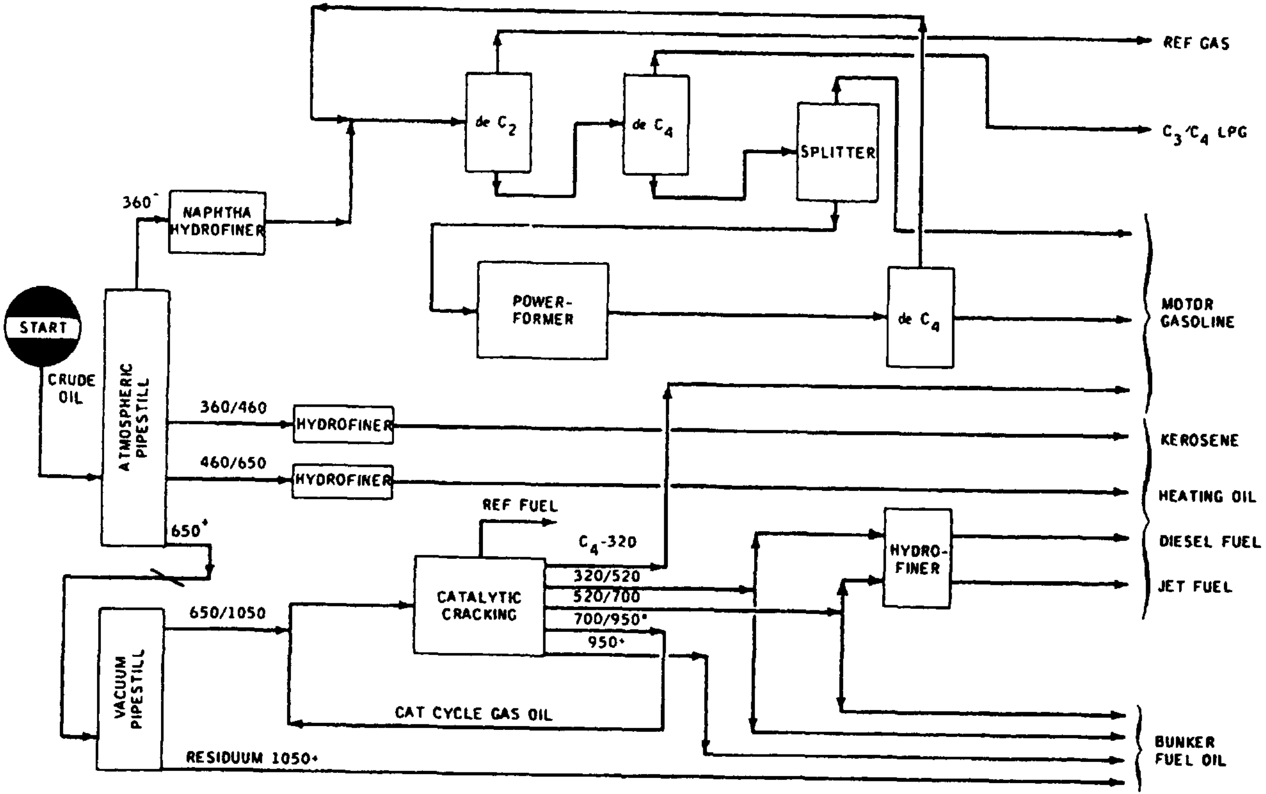

Figure 2. Flowsheet for a conversion refinery.

cracking unit for conversion into high octane gasoline blending stock. Byproducts are gas, distillate, cycle gas oil, and fractionator bottoms. The process uses a fluidized catalyst system. The catalyst is circulated continuously between the reactor where cracking takes place and the regenerator where the coke deposited on the catalyst is burned off. The major competing process is hydrocracking which offers greater conversion and flexibility but usually requires a higher investment.

Hydrocracking is a fixed bed catalytic process which cracks and hydrogenates hydrocarbon feeds. The process consumes large quantities of hydrogen and a hydrogen plant is usually necessary to support the operation. Practically any stock can be hydrocracked, including refractory feeds which resist conversion by other processes. In general, the very heavy residuum from the vacuum pipestill does not make good quality feed for catalytic cracking. In the refinery shown it is blended into residual fuel oil. Many times, however, the market for large volumes of residual fuel oil does not exist. When this is the case, additional conversion units are added to further process the vacuum pipestill bottoms. In other words, the higher the conversion of the refinery the more lighter fractions are produced. The relative levels of conversion vary from refinery to refinery.

A typical maximum conversion type refinery is shown in Figure 3. The higher conversion levels are obtained by additional processing of the bottoms and/or light ends. To increase conversion of the bottoms the amount and/or severity of processing is increased. The resulting fuel oil levels may decrease to zero. Included here in addition to the basic components of a conversion refinery may be fluid coking, delayed coking, and/or visbreaking. These processes are basically thermal cracking processes for reducing the volume and viscosity of the vacuum residuum while producing appreciable quantities of lighter products.

Each of the three processes is commercially used with selection based on particular needs at a given refinery. Some of the various characteristics include:

1. Coking-Delayed Coking and Fluid Coking are the two major variations of this process. Fluid coking produces less coke as compared with delayed coking and hence yields a better product distribution. That is, for a given product slate less crude is converted into coke. The coke produced by fluid coking, however, is of little value as it consists of fine hard particles in contrast to large pieces for delayed coke. This difference in size and texture is important to electrode manufacturers who historically have used delayed coke.

2. Visbreaking is the least expensive of the cracking processes but is limited to the lowest conversion of perhaps 20 to 25% of the feed to 680 °F material.

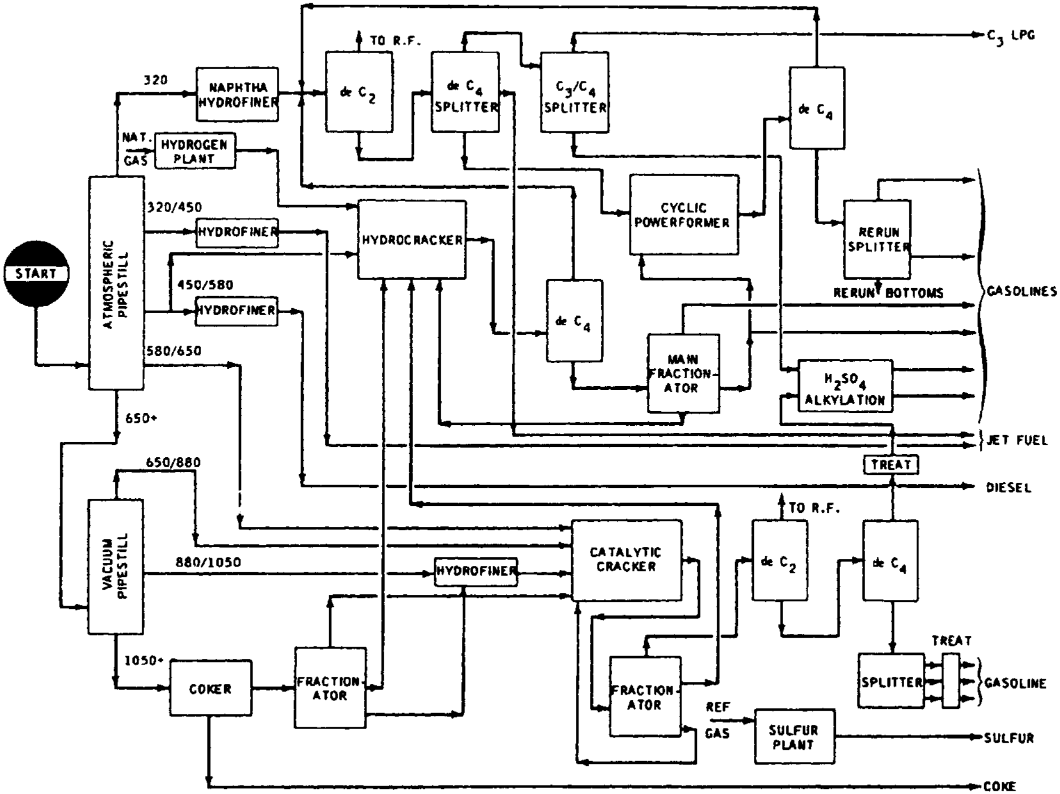

Figure 3. Flowsheet for a maximum conversion type refinery.

To obtain light ends conversion, alkylation and polymerization are used to increase the relative amounts of liquid fuel products manufactured. Alkylation converts olefins, (propylene, butylenes, amylenes, etc.), into high octane gasoline by reacting them with isobutane. Polymerization involves reaction of propylene and/or butylenes to produce an unsaturated hydrocarbon mixture in the motor gasoline boiling range.

An old variation of the conversion type is a catalytic combination unit. Development of this scheme was necessitated by the rising cost of refinery construction after World War II and by the great demand for capital for postwar expansion. The scheme reduced the investment and operating costs for refining equipment. The basic feature of the combination unit lies in the integration of the fractionation facilities of the reduced crude distillation and catalytic cracking sections.

A FUEL PRODUCTS REFINERY WITH SPECIALTIES

A fuel products refinery with specialties may manufacture products such as lubricating oils, asphalts, greases, solvents, waxes and chemical feed stocks in addition to the primary fuel products. The number and diversity of products will naturally vary from one refinery to another, but for purposes of discussion a fuel products refinery with specialties may include many of the following processes.

1. Two-Stage Crude Distillation (Atmospheric and Vacuum) - The vacuum stage can be used alternately to produce heavy gas oil for catalytic cracking feed or raw lube distillate cuts for lubricating oil manufacture.

2. Virgin Naphtha Catalytic Reforming (Powerforming) - This technique is used for the production of high octane motor gasoline, or as a source of aromatic compounds.

3. Light Ends Recovery, Fractionation, and Conversion - Propylenes and butylenes may be recovered for feed to a polymerization plant for production of high octane gasoline; or chemicals. Butylenes and isobutane may be desired for use in an alkylation plant where they are combined to make aviation gasoline and motor gasoline blendstocks. Propanes and butanes may be recovered in essentially pure form for sale as liquefied petroleum gases. It may be profitable to recover ethylene for chemical production. Certain of the light ends components, particularly ethylene, propylene, and butadiene are so in demand that processes such as steam cracking are employed specifically for their

production.

4. Fuel Products Treating- a. Sweetening - This is a process for improving odor of gasolines, kerosenes, and heating oils. The foul smelling mercaptans are converted into bisulfides whose odor is much less objectionable. Among the types in use are copper chloride, hypochlorite, Merox, Mercapfining, and air inhibitor sweetening.

b. Hydroprocessing - The nomenclature system with regard to hydrogen processing is quite confusing with an array of labels involving trade names, terms such as mild, medium, and severe, high and low pressure. Choice of terminology varies widely from company to company.

A wide variety of petroleum fractions may be treated at elevated temperature and pressure with hydrogen in the presence of a catalyst to reduce sulfur, improve stability, odor, combustion characteristics, appearance, and to convert heavy fractions to lighter more valuable products. The most severe form of hydroprocessing as discussed previously is hydrocracking. For fuel products treating, however, two less severe hydroprocessing operations are used, hydrofining and hydrotreating.

Hydrofining usually involves only minor molecular changes of the feed with hydrogen consumption in the range of about 100 to 1,000 cu.ft./bbl. Typical applications include desulfurization of a wide range of feeds (naphtha, light and heavy distillates, and certain residua) and occasional pretreatment of cat cracker feeds.

Hydrotreating essentially involves no reduction in molecular size with hydrogen consumption less than about 100 cu. ft./bbl. Primary application is to remove small amounts of impurities with typical uses including naphtha and kerosene hydrosweetening.

c. SO_2 Extraction - This is a method of solvent extraction with liquid SO_2 to remove aromatic hydrocarbons and cyclic sulfur compounds. It is used to improve the burning qualities of kerosene and diesel fuels, and to reduce sulfur. This process has practically been supplanted by other solvent extraction or by hydrotreating.

5. Fluid catalytic Cracking.

6. Hydrocracking.

7. Residuum Conversion - Included here may be fluid coking, delayed

coking, visbreaking, and residuum hydroprocessing.

8. Solvent Deasphalting - This is the solvent extraction of virgin residuum to remove asphaltenes or other tarry constituents. The deasphalted oil may be further processed into lubricating oils and greases, or used as cat cracking feed.

9. Lubricating Oil Manufacture - This will usually consist of the following processes:

a. Solvent Deasphalting.

b. Phenol Treating - An extraction process for removal of aromatic asphaltic and sulfur compounds from the lube cut.

c. Solvent Dewaxing - Waxy lube is diluted with a solvent such as propane or methyl ethyl ketone (MEK), and cooled to crystallize the wax which is then removed by filtration.

10. Grease Manufacture - Selected lube oil fractions are blended with various metallic soaps to produce high viscosity lubricating greases.

11. Wax Manufacture - A waxy distillate cut from crude or the wax byproduct from lube oil dewaxing is first deoiled. Resulting low oil content wax is hydrofined for color improvement and fractionated into appropriate melting point grades.

12. Asphalt Manufacture - Saleable asphalts are produced from the residua of selected crudes. The residuum itself may be sold as straight reduced cuts to make it easier to handle, producing the so called cut-back asphalts. Another variation is air blown or oxidized asphalts for improved tenacity, greater resistance to weathering, and decreased brittleness. Emulsified asphalts are made for application at relatively low temperatures.

13. Chemical and Other Specialty Manufacture - A wide variety of products may be derived from petroleum feed stocks, including such diverse materials as alcohols, butyl rubber, sulfur, additives, and resins. Other specialties such as solvent naphthas, white oils, Isopars, Varsol, may also be produced. As indicated previously the respective chemical affiliate usually has responsibility for products broadly classified as petrochemicals.

There are many other processes used in refineries not mentioned here. The list above is intended only to emphasize the wide diversity of processing which is common to petroleum refining and to introduce in a very general way some of the more important of these processes. Also it must be emphasized that only fundamental principles of refinery operations have been discussed and modern manufacturing techniques vary widely from company to company.

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CATALYTIC CRACKING

INTRODUCTION

The first commercial catalytic cracking unit was put into operation in 1936. Before the introduction of catalytic cracking, thermal cracking was the major petroleum conversion process. Basically, both processes are tools for the same job; that is, they destroy heavy fuel oil to produce lighter, more valuable products. Thermal cracking does this by heat alone while catalytic cracking combines the action of heat and a catalyst. This combination gives higher yields of more valuable products and, as a result, catalytic cracking replaced thermal cracking.

The *cat* products become feed to other units, such as alkylation and polymerization plants. High boiling liquid products are used to make lubes, and the gas goes into the refinery fuel systems. Cat cracking feed stocks come from atmospheric and vacuum stills, phenol extraction plants, hydrotreaters, deasphalters and cokers.

THE PROCESS

A catalytic cracking unit brings together a heavy feed and an active catalytic agent. In general, the objective of the process is to produce gasoline and heating oil from this heavy feed with a minimum of coke and gas formation. Reactions are carried out during which a carbonaceous material is deposited on the catalyst, which is then regenerated to burn off the deposit and restore the activity. While there are many reactions, such as cracking, isomerization, alkylation, dehydrogenation, etc., taking place in a cat reactor, they are complicated, interrelated and not fully understood.

The usual feed is a virgin gas oil; that is, the part of crude oil boiling between about 60 °F. and 1050°F. Sometimes material below 600°F will be included into the cat feed; but more often, it is put into diesel fuel or home heating oil. The heavy material above 1050°F is not normally used as cat feed because it often contains metallic compounds that contaminate the catalyst. Even if metals are not present, there are sometimes tarry materials that end up on the catalyst. This deposit increases the load on the regenerator, and, hence, the 1050 °F+ material is less desirable than lower boiling feeds.

Broadly speaking, *gas* oils can be considered as mixtures of aromatic, naphthene and paraffin molecules. Aromatic rings are very hard to crack and when they are cracked they tend to deposit a lot of coke on the catalyst. Isoparaffins, on the other hand, shatter easily, and the small fragments wind up in the products as gas or C_4's. Normal paraffins are harder to crack than isoparaffins or naphthenes, but they are easier to crack than aromatic molecules. Higher boiling components have larger molecules which are more easily cracked. Thus, the products and the conversion depend in part upon the type of feeds that are used. The operating conditions of the reactor are also important. This may be visualized if the feed is thought of as a chain, and the reactor as a hammer and anvil. A light tap here and there will crack only an occasional link, and the chain will be broken into a few big pieces. However, if each link of the chain is hit hard several times and the broken pieces are again recracked, most of the original chain will be destroyed, and a number of small pieces will be all that remains. The strength of the chain corresponds to the feed crackability, and the hammer represents the reactor intensity. The intensity is the combined effect of reactor temperature, pressure, amount of catalyst, and catalyst activity. The amount of conversion depends upon both the strength of the chain and how hard it is hit (reactor intensity).

Measures of Conversion

The factors that affect conversion were mentioned above, but the measure of conversion was not described. In a simple chemical reaction, conversion is easily determined by measuring the products formed or the disappearance of the starting material. A petroleum process, however, deals with a multitude of different compounds, many of which still cannot be identified, let alone measured. This makes the selection of a good measure of conversion difficult.

One common practice in petroleum processing is to define conversion as (100 % of material boilings above gasoline). While this is somewhat vague, it can be made more specific if gasoline is considered as hydrocarbons boilings between C_5 and 430°F on a true boiling or 15/5 still. Thus, % 430 conv. = 100 - % products on feed above 430°F. One important drawback with this measure is readily apparent when the feed has material below 430°F. For example, with 5% boiling below 430°F in the feed, this definition would show 5% conversion before the feed reaches the reactor. This drawback can be overcome by using the so-called "corrected 430 conversion" which considers only the conversion of feed boilings above 430°F.

Corr. 430 Conv. = 100 - (% product based on feed above 430 °F x 100)÷(100-% in feed boiling below 430 °F)

This is one of the most common measures of conversion.

While the corrected 430 conversion has proved very useful, it too does not measure all the conversion taking place. If a feed boils between 800°F and 1000°F, for example, it is possible to crack it so that a large quantity of the product would boil between 430°F. and 800°F. This is really converted material, but the 430 conversion does not define it as such.

Product Distribution

Mere conversion of heavy fuel oil to lighter products is not sufficient to guarantee profitable operation of a catalytic cracking unit. The distribution of products is also important, with local refinery demands and price structure playing an important role in dictating which product slate is desirable. Cat cracking yield patterns can be vastly different depending upon the amount of conversion required and the type of products desired. Gasoline octane is most important.

Seasonal chances in gasoline sales and heating oil sales compel some modifications to be made in conversion level. Therefore, the conversion pattern of a given catalytic cracking unit can vary from season to season. In summer operations, for instance, higher yields of motor gasoline are desired, both from direct production of C_5/430°FVT catalytic naphtha and also from conversion of butylenes and isobutane to alkylate.

Catalyst Types

Many substances exhibit catalytic properties to a greater or lesser degree, but only a very few compounds are satisfactory for commercial cracking. To be a good catalyst, a compound must have a high activity so that small quantities will do the job. High activity alone, however, is not enough. The catalyst must have the ability to produce desirable products. For example, a catalyst must not make too much coke. High coke yields are bad; first, because whatever weight of feed goes to coke is lost so far as useful products are concerned; and second, the coke deposited on the surface of the catalyst lowers its activity. This coke deposit must be burned off to regain activity and regeneration is an expensive process.

The catalyst must also be selective to valuable products. Gasoline is desirable, so a lot must be produced, but it must be high octane gasoline. C_3's and C_4's are sometimes required for polymerization, alkylation and chemical production. Certain catalysts give high yields of these compounds, especially the unsaturated components. Gases, such as methane and hydrogen, are undesirable so the yield of these products must be suppressed.

A good catalyst is also stable. It must not deactivate at the high temperature levels (1300 to 1400°F) experienced in regenerators. It must also be resistant to contamination. While all catalysts are subject to contamination by certain metals, such as nickel, vanadium, and iron in extremely minute amounts, some are affected much more than others. While metal contaminants deactivate the catalyst slightly, this is not serious. The really important effect of the metals is that they destroy a catalyst's selectivity. The hydrogen and coke yields go up very rapidly, and the gasoline yield goes down. While Zeolite catalysts are not as sensitive to metals as 3A catalysts, they are more sensitive to the carbon level on the catalyst than 3A. Since all commercial catalysts are contaminated to some extent, it has been necessary to set up a measure that will reflect just how badly they are contaminated.

Over the years, thousands of compounds have been tried as cracking catalysts. These compounds fall into two general categories: *natural* and *synthetic*. Natural catalyst, as the name denotes, is a naturally occurring clay that is given relatively mild treating and screening before use. The synthetic catalysts are of more importance because of their widespread use. Of the synthetic catalysts, two main types are: *amorphous* and *zeolitic*.

For many years the most common catalyst was an amorphous or noncrystalline type called 3A. Initially, all 3A catalyst contained 13% alumina and 87% silica. To improve activity maintenance, the alumina content was increased to 25%. Both 13 and 25% alumina grades continue to be used; the choice at a given refinery is based on the specific situation. Another amorphous type catalyst, containing silica-magnesia and called 3E, is also used.

A major step in catalyst development was the introduction of crystalline zeolitic, or molecular sieve catalysts. Their activity is very high, some of the active sites being estimated at 10,000 times the effectiveness of amorphous silica-alumina sites. Because the zeolite crystals are too small to be used directly, and because of their extremely high activity, small amounts such as 3-25% of zeolite are impregnated on amorphous clay or silica-alumina base to make commercial catalysts.

With the many types of cracking catalysts available today, catalyst selection has become a more important part of refinery planning. For a given refinery situation, both the level and type of conversion are important. For example, a 75% conversion level can be achieved at moderate reactor temperatures with highly active catalyst, producing maximum yields of good octane gasoline; or, a 75% conversion level can be achieved at high reactor temperatures with moderately active catalyst, producing maximum yields of olefinic gases and a gasoline product with very high octane. By changing catalyst types, a refiner now has wide flexibility in choosing the conversion level and product slate that best fit his particular requirements. In general, catalyst types are selected to suit feedstock quality, desired conversion level, and regional and seasonal product distribution requirements.

Catalyst Testing

Since the catalyst is so important to the cracking operation, its activity, selectivity, and other important properties should be measured. A variety of fixed or fluidized bed tests have been used, in which standard feedstocks are cracked over plant catalysts and the results compared with those for standard samples. Activity is expressed as conversion, yield of gasoline, or as relative activity. Selectivity is expressed in terms of carbon producing factor (CPF) and gas producing factor (GPF). These may be related to catalyst addition rates, surface area, and metals contamination from feedstocks.

Correlations

Catalytic crackings operations have been simulated by mathematical models, with the aid of computers. The computer programs are the end result of a very extensive research effort in pilot and bench scale units. Many sets of calculations are carried out to optimize design of new units, operation of existing plants, choice of feedstocks, and other variables subject to control. A background knowledge of the correlations used in the "black box" helps to make such studies more effective.

An important part of any cat crackings correlations package is an accurate feed characterization technique. Feed characterization is vital since it quantifies the relative ease with which a given feed will crack and the tendency of the feed to form desirable products. For many years catalytic cracking feedstocks have been characterized by routine physical inspections which are related to total paraffins, naphthenes and aromatics. This technique was sufficiently accurate for virgin gas oils, but it has proven less accurate for the blends of cracked gas oils and deasphalted and hydrotreated feeds.

Correlations have been used as a tool for catalyst selection studies. Predictions of the product yields and qualities possible with various catalysts can provide the necessary information for a refiner to study the economics of switching catalysts, for instance. With a good idea of the profitability of changing catalyst types, the refinery can justify such a change in his cat crackings operation.

Another important use of correlations is the optimization of existing unit operations. Cat cracking correlations can provide the refiner with valuable information for optimizing reactor temperature level, gasoline/distillate cut point, and feed and recycle rates. The practical application of this information can mean increased profitability for the cat cracking operation.

EQUIPMENT

The first cat cracking unit was a fixed bed reactor using natural clay pills or pebbles. Oil vapor was passed through the bed, and the cracking reaction took place on the surface of the catalyst. As the crackings reaction proceeded, coke was deposited on the catalyst which lowered its activity. After about 10 minutes, it was necessary to regenerate the catalyst by burning off the coke. The regeneration took 20 minutes, so one cycle was completed every 30 minutes. An approach toward a continuous process was made by building several reactors so that at least one would be on stream as a reactor while the others were being regenerated.

During World War II, the fluid catalytic cracker was developed to meet the high demands for aviation. Various design configurations developed in subsequent decades, including a side-by-side reactor/regenerator configuration with U-shaped catalyst transfer lines between the vessels. In 1971 FLEXICRACKING technology was introduced which features a similar side-by-side configuration with a number of improvements including increased riser cracking time and reactor operating flexibility, improved feed injection, more efficient catalyst stripping, better regenerator air distribution and various mechanical improvements.

Flexicracking Operations

The flow plan of a Flexicracking unit is shown in Figure 1. Catalyst is circulated rapidly in U-bend transfer line between reactor and regenerator. The feed, which is usually a heavy gas oil, is preheated and injected into the stream of regenerated catalyst moving from the regenerator to the reactor. The hot catalyst vaporizes the oil (there is usually 5 to 8 times as much catalyst as oil), and the vapors help to fluidize the catalyst in the reactor. This mixture flows into the reactor where the cracking takes place at temperatures usually ranging from 900 to 1000°F. The reaction is endothermic and consumes the heat brought from the regenerator by the catalyst.

The cracked products pass out through two stages of cyclones which collect entrained catalyst and return it to the dense bed. Velocities at the outlet of the dense bed are normally 2.0-3.0 ft./sec. Upon leaving the cyclones, the vapors go to the primary fractionator which separates the heavy products from the gasoline and lighter components. The light products go on to the light ends recovery unit. The heavy material is separated and either recycled to the reactor or withdrawn from the system.

As in the fixed bed process, carbon is deposited on the surface of the catalyst and covers the active sites. This deposit lowers the activity very rapidly.

Since the activity falls so rapidly, it is desirable to regenerate as often as possible. In normal operation, the catalyst is held in the reactor only about 0.5 to 1.0 minute. This is accomplished by the rapid circulation of catalyst between reactor and regenerator, which in the larger units may be as high as 40-50 tons minute.

The spent catalyst is withdrawn from the bottom of the reactor and stripped with steam to vaporize the hydrocarbons remaining on the surface. Stripping also removes most of the hydrocarbon vapors which are entrained between the particles of catalyst. Without stripping, hydrocarbon products would be carried to the regenerator and needlessly burned consuming much of the regeneration air, and decreasing yield of useful products.

The stripped catalyst is picked up by a stream of air and carried into the regenerator where the carbon is burned at temperatures about 1100-1300°F. Entrained catalyst is again removed by cyclones and the flue gas goes out the stack. The hot, regenerated catalyst leaves the regenerator and takes with it much of the heat of combustion. This is carried over to the reactor to vaporize the feed and to balance the endothermic heat of cracking. Thus, the process is heat balanced.

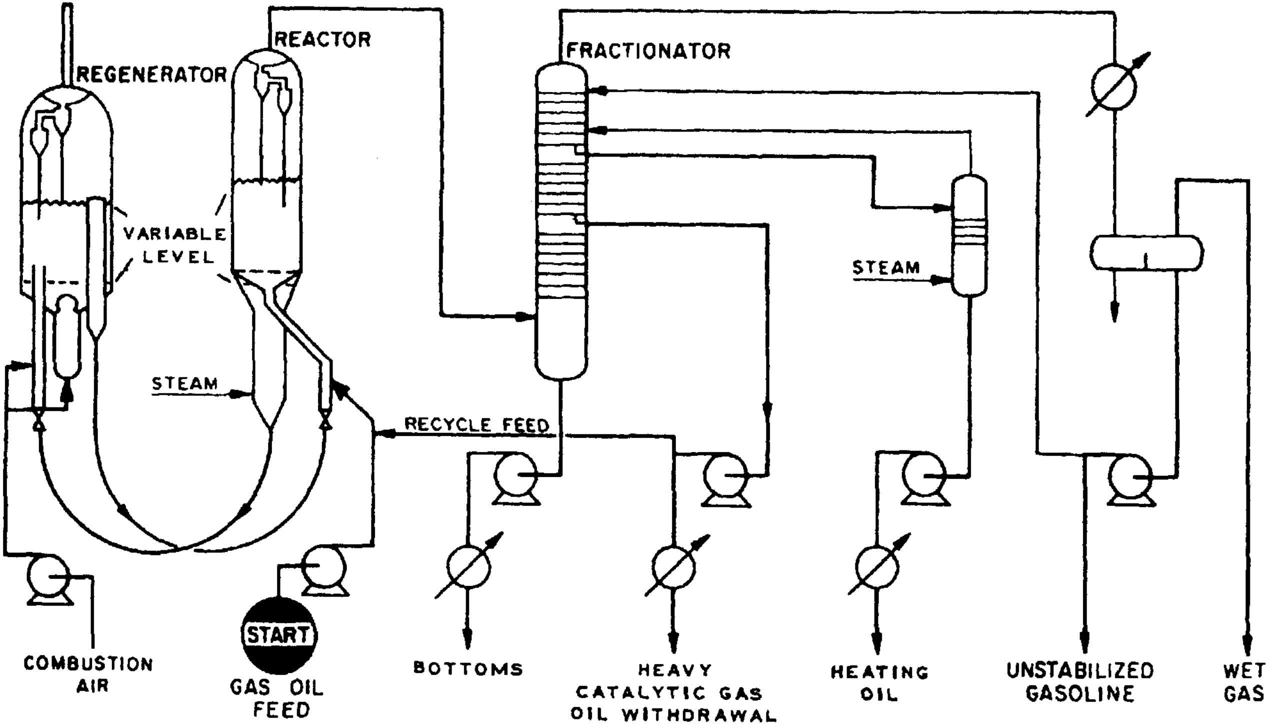


Figure 1. Flow scheme for flexicracking unit.

Combination Unit

A combination unit is a special type of unit that was developed to reduce the investment for a small refinery. In effect, one main distillation unit serves as a crude fractionator as well as the cat unit primary fractionator. This same tower also serves the naphtha reformer and visbreaker. A schematic diagram of a combination unit is shown in Figure 2. Crude oil is topped (material boiling below 650°F is removed) in the atmospheric tower, and the topped crude is sent to the combination tower along with cat products and naphtha reformer products. These latter streams provide heat to distill the topped crude and also, being more volatile than topped crude, provide a lifting effect which assists in vaporizing more of the crude.

A sidestream of material boiling between about 650°F/925°F is withdrawn as cat feed. This stream is made up of virgin material from the crude and recycle cat products. When it is available additional extraneous virgin feed also may be blended into the cat feed stream. Fractionator bottoms are withdrawn and may be sent to fuel oil.

While this unit is considerably cheaper, it also has certain disadvantages. For example, changes or upsets in any one unit may be felt throughout the refinery because of the changes in fractionator operation. However, the considerable cost saving possible with the combination type unit has permitted many small refineries to finance a catalytic cracking unit when they could not afford a conventional model.

Various Design Configurations

Various design configurations for fluid catalytic crackers are illustrated in Figure 3. Their distinguishing features can be summarized as follows:

Model II: Regenerator at higher elevation and lower pressure than reactor. Slide valves control catalyst circulation.

Model III: Regenerator and reactor at approximately equal elevation and pressure. Slide valves control catalyst circulation.

Model IV: Regenerator and reactor at approximately equal elevation and pressure. Catalyst circulates through U-bends, controlled by pressure balance and variable dense-phase riser.

UOP: Regenerator at lower elevation and higher pressure than reactor. Slide valves control catalyst circulation.

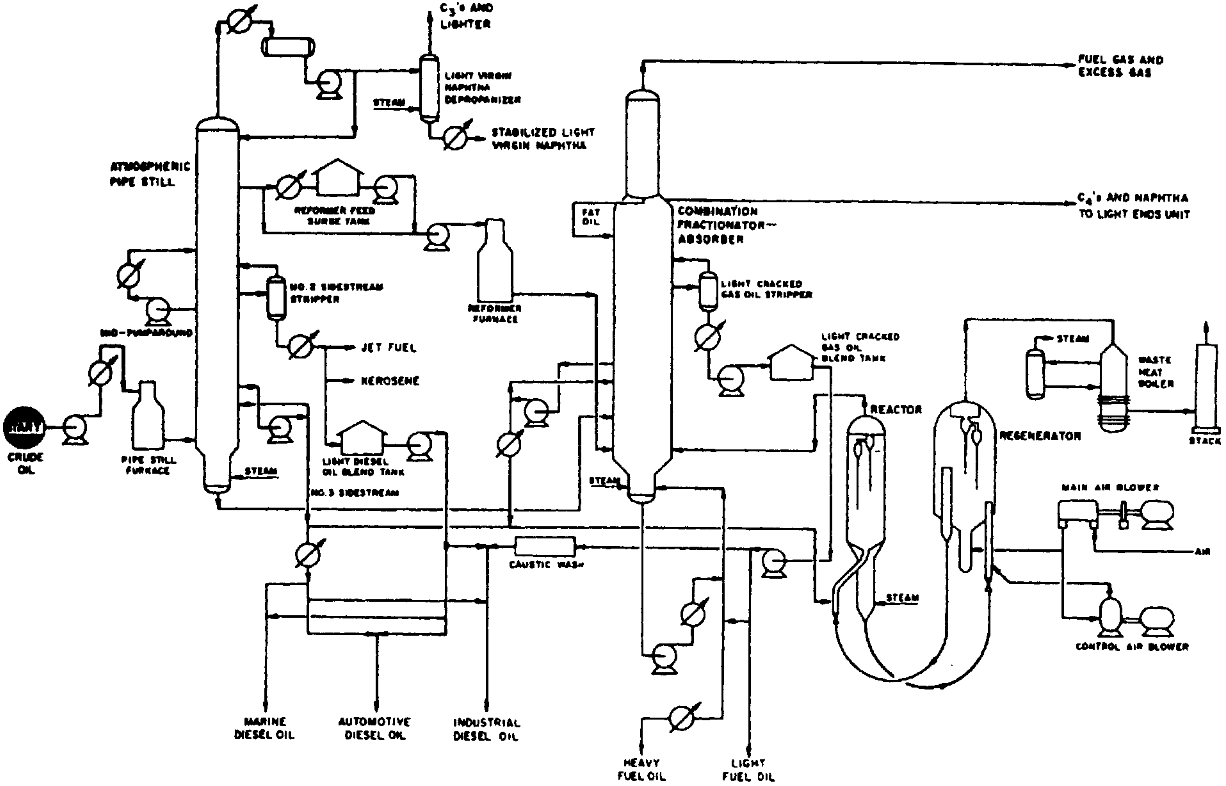
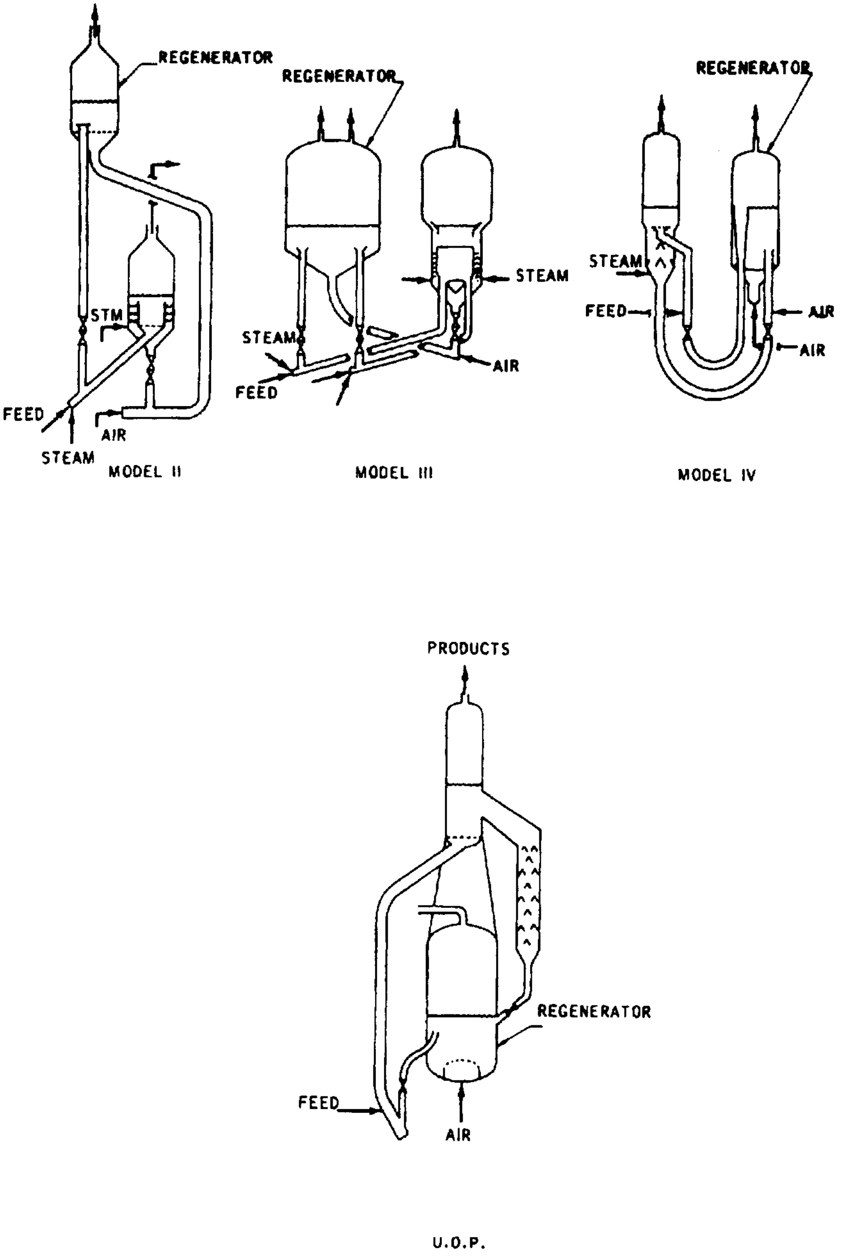


Figure 2. Flow diagram of a combination unit.



MODEL II MODEL III MODEL IV

U.O.P.

Figure 3. Various designs for fluid catalytic crackers.

Flexicracker: Side-by-side configuration with U-bends similar to Model IV design but featuring increased riser cracking, improved feed injection, and various mechanical improvements are also used. Many of the reactors for the units described above were designed originally as dense bed reactor units. Some of them have been modified to include FLEXICRACKING features such as minimized bed cracking and better feed injection.

Various companies worked on the development of Fluid catalytic cracking units. During World War II, the government requested some of the leaders in this field to pool their knowledge so as to speed the production of aviation gasoline. The fact that so many Fluid units were constructed and put into operation in such a short time shows that this joint effort was successful. However, because of this effort, many of the basic Fluid patents were held for many years in combination with other companies, some of which also developed their own Fluid designs.

Other companies have brought out their own proprietary unit designs aimed at capitalizing on the cracking characteristics of the zeolite catalysts. Some are transfer line units, in which the riser becomes the reactor and the old reactor becomes a separator vessel. Gulf, Texaco and Kellogg offer these "riser-cracker" type designs. Separate risers may be employed for fresh feed and recycle. A small dense bed at the top of the riser may be used to obtain the last bit of conversion. The specific reactor design selected is related closely to the desired product distribution and activity of the sieve catalyst used. The higher the activity, the less time is needed for reaction. Most designs differ in the "hardware" used, rather than in basic process characteristics.

Over the years, many improvements have been made to the catalytic cracking process. Basically, the total development of process and mechanical improvements in catalytic cracking since its inception can be divided into two time periods: pre-zeolitic catalyst introduction and post-zeolitic catalyst introduction. In the period from the early 40's to the early 60's, there were some improvements made in catalyst quality, but in general these catalyst changes did not dictate mechanical and operational changes in the basic units. Rather, significant improvements were made in unit service factor and flexibility. One of the most important advances came from the ability to crack heavy feeds. Original moving bed units cracked only light feeds that were completely vaporized. The new fluid units could handle both vapor and liquid feeds of high boiling points. Final cut points are now generally held to about 1050°F to avoid metals contamination of the catalyst. Many engineering improvements were made, reducing investments, improving run lengths and service factors, and lowering operating costs. The use of CO boilers to recover heat from the CO content and sensible heat in regenerator flue gases was another important step in improving the profitability of the catalytic cracking process.

When zeolitic cracking catalysts were first used, a new era in catalytic cracking had begun. The results of these research efforts were more catalyst types of higher and higher activity and selectivity. These improvements were of such significance that they dictated new ways of contacting catalyst and oil to achieve optimum catalyst performance. The increased use of riser or transfer line type cracking was required, and revamps of existing units were directed toward this goal. In addition, the desirability of low CRC (carbon on regenerated catalyst) levels for these active zeolitic catalysts dictated improved regeneration techniques. At the higher levels of regenerator pressure, flue gas expanders have been found to be attractive, recovering the energy added by the air blower.

Emission Control

Fluid catalytic cracking units present formidable emission control problems. Contaminants are present in both reactor product gas and regenerator flue gas. The reactor product contains hydrogen sulfide, ammonia, and cyanides, plus combined sulfur and nitrogen in the liquid products. Hydrogen sulfide, ammonia and cyanides are handled as part of the overall refinery waste water cleanup. The combined sulfur and nitrogen may be removed by hydrotreating.

Both particulate and gaseous contaminants are present in the regenerator flue gas. Regenerator emissions regulations are set by the Environmental Protection Agency (EPA). The most prominent particulate contaminant in the flue gas is fine catalyst dust which escapes the regenerator cyclones. Although these cyclones are 99.9% efficient, losses still amount to about 3.5 tons/day for a 45 MB/D unit. The escaping catalyst dust can be removed with electrostatic precipitators which electrically charge the particles and collect them on large metal plates. The gaseous contaminants in the flue gas streams include compounds of feed sulfur and nitrogen in addition to the carbon monoxide formed by incomplete combustion of coke in the regenerator. Harmful CO gas can be combusted to CO_2 in CO boilers which utilize the heat of CO combustion to generate steam for use within the refinery. SO_x and NO_x removal from flue gases can be accoplished by hydrotreating of the catalytic cracker feed stocks.

3

FLUID SOLIDS OPERATIONS

INTRODUCTION

Fluid bed reactors became important to the petroleum industry with the development of fluid catalytic cracking (FCC) early in the Second World War. Today FCC is still widely used. The following section surveys the various fluid bed processes and examines the benefits of fluidization. The basic theories of fluidization phenomena are also reviewed.

FLUID BED PROCESSES

Catalytic Cracking

Cracking is a process for breaking down large, high boiling petroleum molecules to produce light gases, gasoline, jet fuel, and heating oils. Cracking originally was a thermal process relying on high temperature to cause reaction. In the early fixed bed catalytic process, the catalytic process operated at lower temperatures and produced a more valuable product distribution. Because the reaction was endothermic and deposited carbon on the catalyst, cyclic operation was necessary to permit frequent regeneration and reheating of the catalyst. To meet these reaction needs, a fluid bed process was developed in which very fine catalyst particles were circulated between a reactor and a regenerator vessel. This provided much more satisfactory uniform operation and now most catalytic cracking is carried out in fluid bed processes.

Fluid cat cracking required identifying stable operating regimes for beds of fine catalyst at high gas flow rates. Highly efficient cyclone and electrostatic systems had to be developed for catalyst recovery. Finally, the principles of pressure balancing and standpipe/transfer line flow had to be developed for the catalyst circulation scheme.

Fluid Hydroforming

As the demand for high octane gasoline increased, reforming processes were developed. In reforming, low octane normal paraffins are isomerized and/or aromatized. One process developed was a fluid bed process, called fluid hydroforming. In concept, it was similar to cat cracking, employing a reactor and regenerator with solids circulation for regeneration and heat balancing. However, fluid hydroforming faced some additional problems in the area of gas-solids contacting. Due to the nature of the process, fluidization was not inherently good in hydroforming, but at the same time the reaction required better contacting than FCC to obtain high yields of high octane gasoline. Thus, a large and ultimately successful program was carried out to improve fluidization and contacting. This work led to some of the earlier theories explaining the behavior of fluidized beds as chemical reactors.

While there are hydroformers still operating, reforming today is generally carried out in fixed bed units using platinum catalysts, because of their superior product yield and distribution. Fluid platinum catalyst processes are not feasible because catalyst losses would be too great.

Fluid Coking

Coking is a thermal process for converting heavy, residual oils into lighter products and solid carbon. In the earliest coking process, called delayed coking, after heating and partial vaporization, the residuum is passed into a coking drum which fills up with solid coke deposits. This coke must then be drilled out. The coke can be deposited on particles of seed coke in a fluidized bed, and the coke product is in the form of freely flowing granules. Fluid coking also employs two beds with particles circulating between the coking reactor and a burner vessel, where some coke particles are burned to produce the necessary heat.

Fluid coking is very insensitive to poor gas-solids contacting, but has one problem not faced by cat cracking or hydroforming. If the heavy residual oil is fed too fast to the reactor, the coke particles will become wetted and stick together in large unfluidizable lumps. Correct control of feed rate is necessary to prevent this bogging.

FIOR

Fluid Iron Ore Reduction (FIOR) is a process for reducing ore to iron with a reducing gas in a fluid bed. For thermodynamic efficiency, iron ore reduction requires counter current flow of ore and reducing gas. This is achieved in FIOR in a multiple bed reactor. Precautions are necessary to prevent significant back mixing of solids between beds, since this would destroy counter current staging.

Other Fluidization Processes

Several other processes that are aimed at the manufacture of gasoline from coal have been applied over the years. The main reactor in these processes uses three phase fluidization in which solid coal particles, gases, and liquids are all contacted at very high temperatures and pressures. Fluid bed dryers and fluid cokers are also used in synthetic fuels manufacture.

Fluid bed boilers have also been applied as a cure to sulfur dioxide air pollution from power plants. Various schemes have been developed in which combustion of a sulfur containing fuel takes place in a fluidized bed of particles which absorb or react with sulfur dioxide. The particles are usually regenerated to recover sulfur, which often has enough by-product value to make a significant contribution to process economics.

Several important applications of fluid beds exist outside the petroleum industry. Fluid bed roasting of pyritic ores is widely used in the metallurgical industry. Calcination of lime is a commercial process. There are also fluidization processes for various nuclear processing steps.

Fluid bed processes have been subject to many problems and uncertainties in development and scale up from bench-scale reactors. The fluidization behavior of each process seems different and very often does not meet expectations based on experience with earlier plants. With hindsight fluid cat cracking seems to be an ideal system from the point of view of easy operation and straightforward scale up.

BEHAVIOR OF FLUIDIZED BEDS

Qualitative Aspects of Fluidization

When a fluid flows upward through a bed of solid particles, pressure drop across the bed increases as the flow rate increases. Eventually the pressure drop equals the weight of the bed (per unit horizontal area) at which point the particles are suspended by the flowing fluid, and the bed is said to be "fluidized". In this state, the particles in the bed are able to move freely instead of resting on each other, and the bed behaves much like a liquid. Waves can be generated and propagated along the surface of the bed. A large low density solid object will float on the surface of the bed. Particles will flow like a liquid between vessels and will stream in a jet from a hole in the side of the bed.

As the fluid flow rate is increased beyond the minimum fluidizing velocity, the bed expands. The total pressure drop across the bed remains essentially constant, but the pressure drop per unit height decreases as shown in Figure 1. At flow rates above the minimum for fluidization, bed characteristics depend greatly on whether the fluid is a liquid or a gas. For most liquid fluidized systems, the bed expands uniformly, with particles being well dispersed throughout. For most gas fluidized beds the gas flow in excess of minimum fluidization will pass up through the bed as bubbles. These bubbles have many of the properties of gas bubbles in liquids. The bubbles contain very few particles. Most particles remain in the "emulsion phase,'' packed closely together as they were at minimum fluidization conditions.

Fluidized beds have a number of properties which make them desirable in certain reaction and processing steps. They permit the use of fine particles, which have very large surface areas for catalysis and gas-solids contacting. A packed bed of fine particles would have a prohibitively high pressure drop at reasonable gas velocities. Solids are well mixed so that temperatures throughout the bed are uniform even when very exothermic or endothermic reactions are occurring. In addition, heat transfer between the fluid bed and exchanger surfaces is high. The ease of handling fluidized solids makes feasible the continuous addition and removal of solids.

A number of disadvantages are also associated with fluidized beds, although proper design can often minimize them. Perhaps of greatest concern is the problem of contacting. The gas in bubbles is not in intimate contact with solids, and the bubbles have comparatively short residence times in the reactor, thus, bypassing often occurs. The seriousness of this depends on the size and velocity of the bubbles, the design of the reactor, and the kinetics of the desired reaction. Flow patterns, particularly for solids, approach those in stirred tanks rather than those in plug flow reactors, an undesirable feature if high conversion is desired. This, of course, can be remedied by staging several fluid beds in series. Sticky or agglomerating materials often prove hard to handle in fluid beds. Entrainment of fine particles can cause serious solid losses and air pollution if efficient recovery devices are not provided. For some low pressure processes, such as burners, pressure drop may be substantially higher for a fluid bed (because the gas supports the solids) than for the more conventional process. Bubbles make

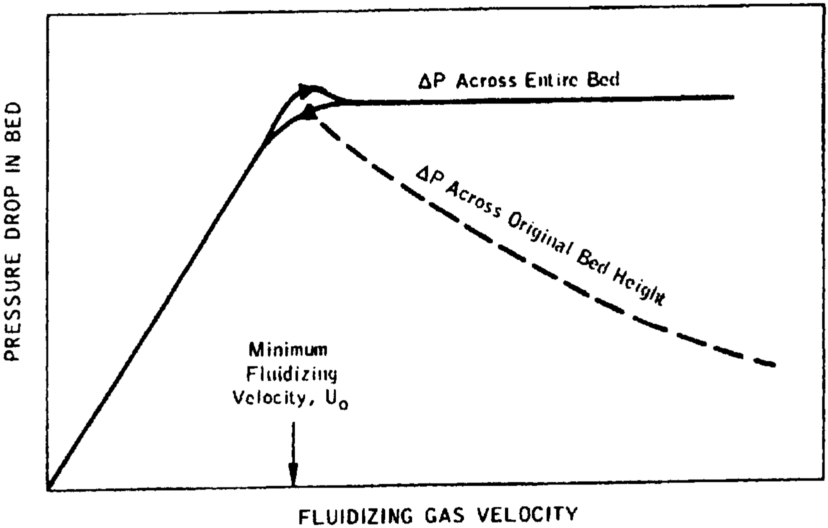


Figure 1. Pressure drop in fluid beds.

an important contribution to the behavior of most fluidized beds.

Bubble Formation and Rise Velocity

When a gas is blown steadily through an orifice into an essentially inviscid liquid, a regular stream of bubbles is formed. A theoretical expression that relates the bubble volume V to volumetric gas flow rate G and gravitational acceleration g is the following:

$$V = 1.138\,(G)^{6/5}/(g)^{3/5} \tag{1}$$

This equation has been experimentally verified in liquids, and Figure 2 shows that it applies equally well for fluidized solids, provided that G is taken as the flow rate in excess of minimum fluidization requirements. In most practical fluidized beds, bubbles coalesce or break up after formation, but this equation nevertheless gives a useful starting point estimate of bubble size.

Single gas bubbles in an inviscid liquid have hemispherical leading surfaces and somewhat flattened wakes. Their rise velocity is governed by Bernoulli's theory for potential flow of fluid around the nose of the bubble. This was first solved by G. I. Taylor to give a rise velocity U_B of :

$$U_B = 0.71\sqrt{gD_B} \tag{2}$$

where D_B, is the equivalent bubble diameter. By means of X-ray photography, this has been verified for bubbles in fluidized beds, although there is some variation in the coefficient.

When the bubble diameter approaches the diameter of the containing vessel, slug flow is said to exist. In such cases, the bubble rise velocity is given by

$$U_B = 0.35\sqrt{gD} \tag{3}$$

where D is the vessel diameter. Slug flow occurs very often in laboratory sized equipment. This is a major contributing factor to scale up effects, since bubble behavior is directly related to vessel size.

Bubble Rise During Continuous Fluidization

Bubbles in liquids rise at velocity U_B relative to the surrounding fluid. This fluid itself moves upward at velocity U when gas is introduced at superficial velocity

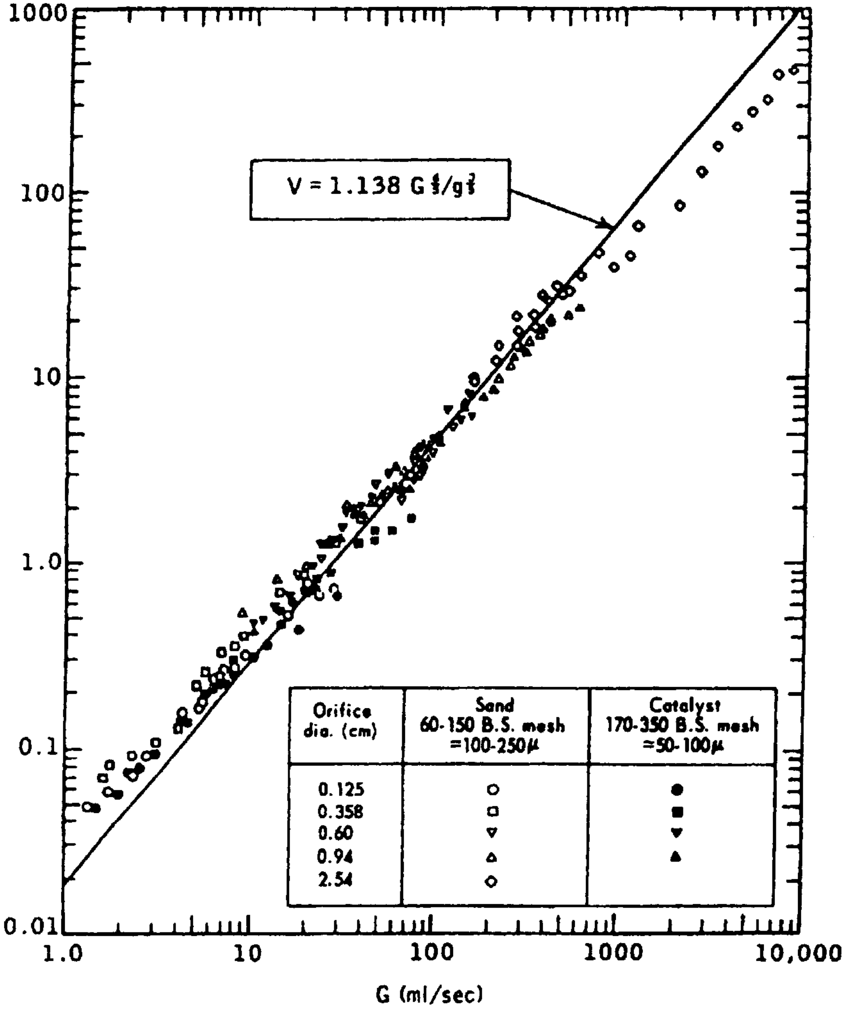


Figure 2. Bubble formation at an orifice.

U to the bottom of the bed. The absolute velocity of a bubble as seen by a stationary observer is then

$$\mathbf{U_A = U_B + U} \tag{4}$$

For fluidized beds, part of the gas flows through the emulsion at minimum fluidization velocity U_o, leaving $U - U_0$ to influence bubble behavior. Then equation (4) is modified to read :

$$\mathbf{U_A = U_B + (U - U_o)} \tag{5}$$

Bed Height and Density

Knowing the bubble rise velocity, the bed expansion can be predicted from a material balance on the bubble phase gas. Thus, total gas flow through the bubble phase equals absolute bubble velocity times the volume fraction E of bubbles in the bed.

$$\mathbf{(U\text{-}U_o) = U_A \eta} \tag{6}$$

The fraction η can be expressed in terms of the expanded bed height H and minimum fluidization height H_o, and terms may be substituted for U_A.

$$\mathbf{(U - U_o) = (U_B + U - U_o)\,(H\text{-}H_o)/H} \tag{7}$$

This may be solved to give the expansion compared to minimum fluidization conditions.

$$\mathbf{H/H_o = \rho_o / \rho = 1 + (U - U_o) / U_B} \tag{8}$$

where ρ_o and ρ are bed densities at minimum fluidizing and actual conditions.

Equation 8 has not been rigorously tested on freely bubbling beds because of the difficulty of making precise independent measurements of D_B or U_B under such conditions. It has been verified for slugging beds, however, as is shown in Figure 3. In the case of slugging, the surface level of the bed oscillates considerably, and in this case, the maximum bed height H_{max} is used for H in Equation 8.

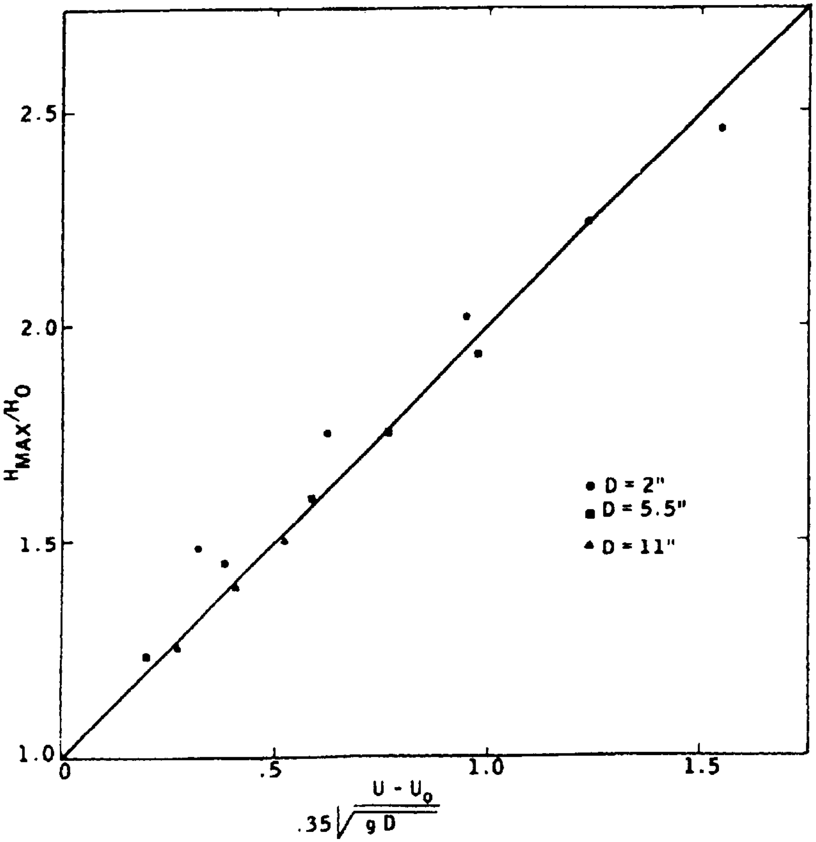


Figure 3. Bed expansion data.

Bubble Clouds and Contacting

At any instant, pressure is uniform throughout a bubble, while in the surrounding emulsion pressure increases with depth below the surface. Thus, there is a pressure gradient external to the bubble which causes gas to flow from the emulsion into the bottom of the bubble, and from the top of the bubble back into the emulsion. This flow is about three times the minimum fluidization velocity across the maximum horizontal cross section of the bubble. It provides a major mass transport mechanism between bubble and emulsion and hence contributes greatly to any reactions which take place in a fluid bed. The flow out through the top of the bubble is also sufficient to maintain a stable arch and prevent solids from dumping into the bubble from above. It is thus responsible for the fact that bubbles can exist in fluid beds, even though there is no surface tension as there is in gas-liquid systems.

The flow pattern of gas within the emulsion phase surrounding a bubble depends on whether the bubble velocity U_B is less than or greater than minimum fluidization velocity U_0. For $U_B < U_0$, Figure 4(A) is representative. The presence of the void of the bubble causes a distortion of the gas flow lines. For $U_B > U_0$, the much different case of Figure 4(B) results. Here a gas element which leaves the bubble cap rises much more slowly than the bubble, and as the bubble passes, it returns to the base of the bubble. Thus, a cloud of captive gas surrounds a bubble as it rises. The ratio of cloud diameter to bubble diameter may be written

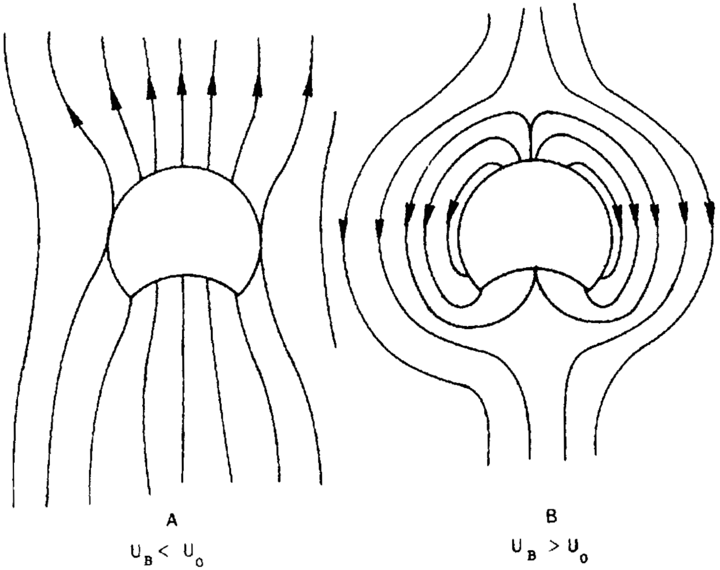
$$D_C/D_B = \{U / (U_B - U_0)\}^{1/3} \tag{9}$$

In addition to this convective cross flow of gas from the bubble into the emulsion phase of the cloud, mass transfer also occurs by diffusion into the emulsion.

The effectiveness of a fluidized bed as a chemical reactor depends to a large extent on the amount of convective and diffusive transfer between bubble gas and emulsion phase, since reaction usually occurs only when gas and solids are in contact. Often gas in the bubble cloud complex passes through the reactor in plug flow with little back mixing, while the solids are assumed to be well mixed. Actual reactor models depend greatly on kinetics and fluidization characteristics and become too complex to treat here.

Coalescence

In a freely bubbling fluidized bed, small bubbles often tend to coalesce into larger



Figures 4. Gas flow through bubbles.

ones. This can occur when a bubble enters the wake region of the bubble above it. The wake of a bubble moves upwards with the bubble, and when a trailing bubble enters a wake region, it moves at U_B relative to the solids in the wake, which are already moving at U_A. The trailing bubble is thus sucked up into the leading bubble. With the possibility of coalescence, the picture suddenly becomes more complex, because bubble velocities and sizes are no longer constant but change with height in the bed. Rigorous models for coalescence have not been verified under realistic fluidization conditions. It is often possible, however, to make educated guesses as to bubble growth rates so that reasonable calculations can be made.

Maximum Bubble Size

Measurements in large fluidized beds of fine particles indicate that bubble coalescence often ceases within a short distance above the gas distributor plate. Indications from density measurements or single bubble velocities are that bubble velocity U_B and diameter D_B often reach maximum stable values, which are invariant with height or fluidizing gas velocity.

Small bubbles give smoother more uniform fluidization and greater bed expansion than large ones. In addition, they have a greater relative interfacial area for mass transfer and have a longer residence time in a fluid bed, so that chemical or physical reaction between gas and solids is enhanced by small bubbles. The correlation between smooth physical fluidization and good reactor performance had long been recognized. It was known that increasing gas density and viscosity and decreasing particle density and size (high *fines* content) improved reactor behavior. The theory of maximum stable bubble size can explain why these changes tend to produce smaller bubbles. Referring to Figure 5, as a bubble rises in a fluid bed, shear at the bubble-emulsion interface sets up circulation patterns of gas within the bubble. The circulation velocity is roughly equal to U_B. When this velocity approaches the free fall velocity of particles in the wake, they are entrained into the bubble, and too much entrainment destroys the bubble. Factors which reduce the free fall velocity of bed particles therefore reduce the stable bubble size. Figure 6 shows a bubble in the process of breaking up in this manner. Figure 7 shows the effect of fines content on bed density and bubble size.

Entrainment

Economic considerations dictate that a commercial fluid bed operates at as high a gas velocity as practicable. A frequent limiting factor is entrainment from the bed which is a very strong function of gas velocity. A well defined fluid bed can be maintained even at gas velocities well in excess of the free fall velocity of the biggest particles

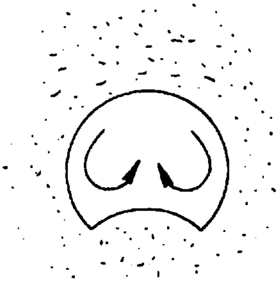


Figure 5. Internal circulation and maximum bubble size.

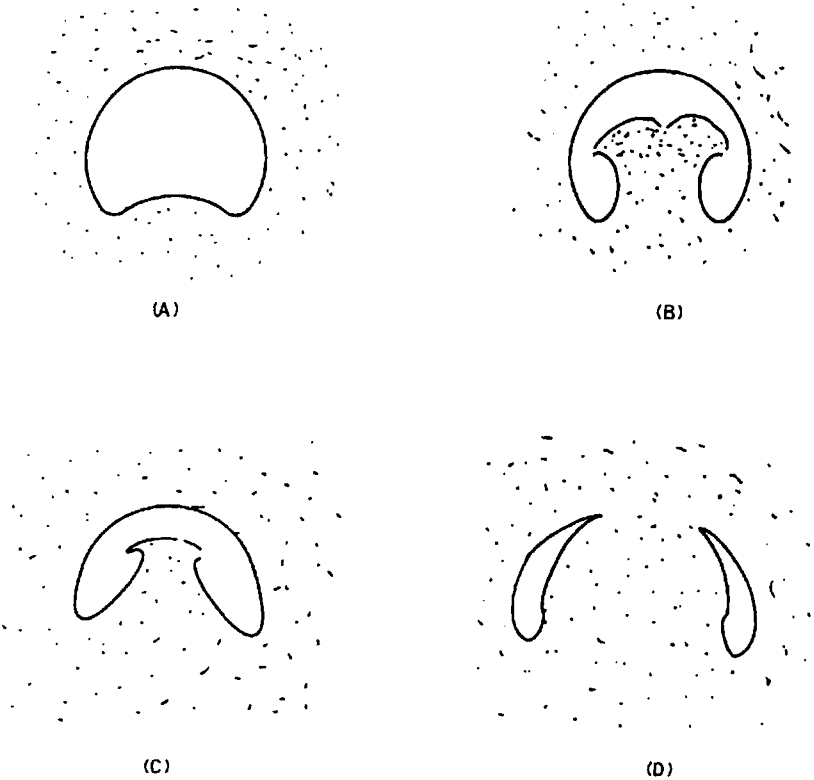


Figure 6. Experimental observation of bubble break-up.

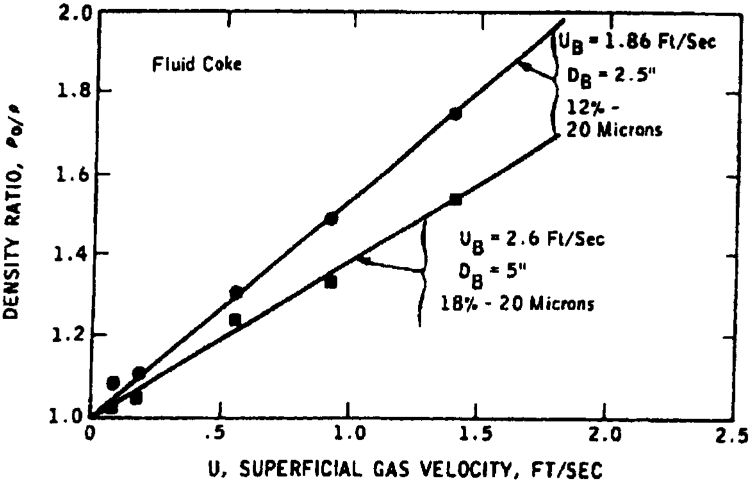


Figure 7. Bed expansion with small bubbles.

in the bed. Though entrainment can occur only at a finite rate, the rate can be high, and entrained particles must be returned to the bed or the bed will rapidly disappear. Entrainment in the regenerator of a large cat cracker can be as high as 40 tons per minute, and the importance of efficient entrainment recovery is obvious.

Entrainment may be separated into two zones. Near the bed, it decreases rapidly with height, but above the so called transport disengaging height (TDH) it is constant. The primary contribution to entrainment at low heights above a bed is the solids thrown up from the wakes of large fast moving bubbles when they break the surface. Gas velocities at the surface vary greatly both in time and with position, and high velocity streams are capable of carrying large solids concentrations. These streams dissipate as jets, however, and as they become slower and broader with height above the bed, their carrying capacity decreases and some of the solids fall back to the bed. Finally, far enough above the bed, the gas velocities have become uniform, and the solids have lost any excess kinetic energy due to violent bubble eruptions, so entrainment reaches a constant level. This explains the seemingly anomalous observation that entrainment for coarse particles may be greater than for fine ones. If entrainment is measured below the TDH, the fluidization behavior of the bed can be the dominant factor, with coarse particles giving much worse, less uniform fluidization characteristics and higher entrainment. Entrainment also depends very strongly on gas velocity. Figure 8 shows typical entrainment rates for cat cracking catalyst and fluid coke.

Heat Transfer

Fluid beds have very favorable heat transfer characteristics in three important modes: thermal conduction within the bed; heat transfer between gas and solids; and heat transfer between bed and heat transfer surfaces. Thermal conductivity within the bed depends to a large extent on specific solids flow patterns and gas bubbling characteristics. It usually is greater in large units than in small ones. Temperature differences between points in a uniformly fluidized bed are never more than a few degrees, and any greater differences are usually strong evidence of malfunctioning of the bed.

Heat transfer between gas and solids is exceedingly hard to measure because it is so rapid. Although the coefficient is low, the available surface area and the relative specific heat of solid to gas are so large that temperature equilibration occurs almost instantaneously. Experiments on injection of argon plasmas into fluidized beds have shown quenching rates of up to fifty million degrees Kelvin per second. Thus, in a properly designed bed, gas to solids heat transfer is not normally a matter of concern.

Heat transfer from surfaces to a fluidized bed is a strong function of the rate at

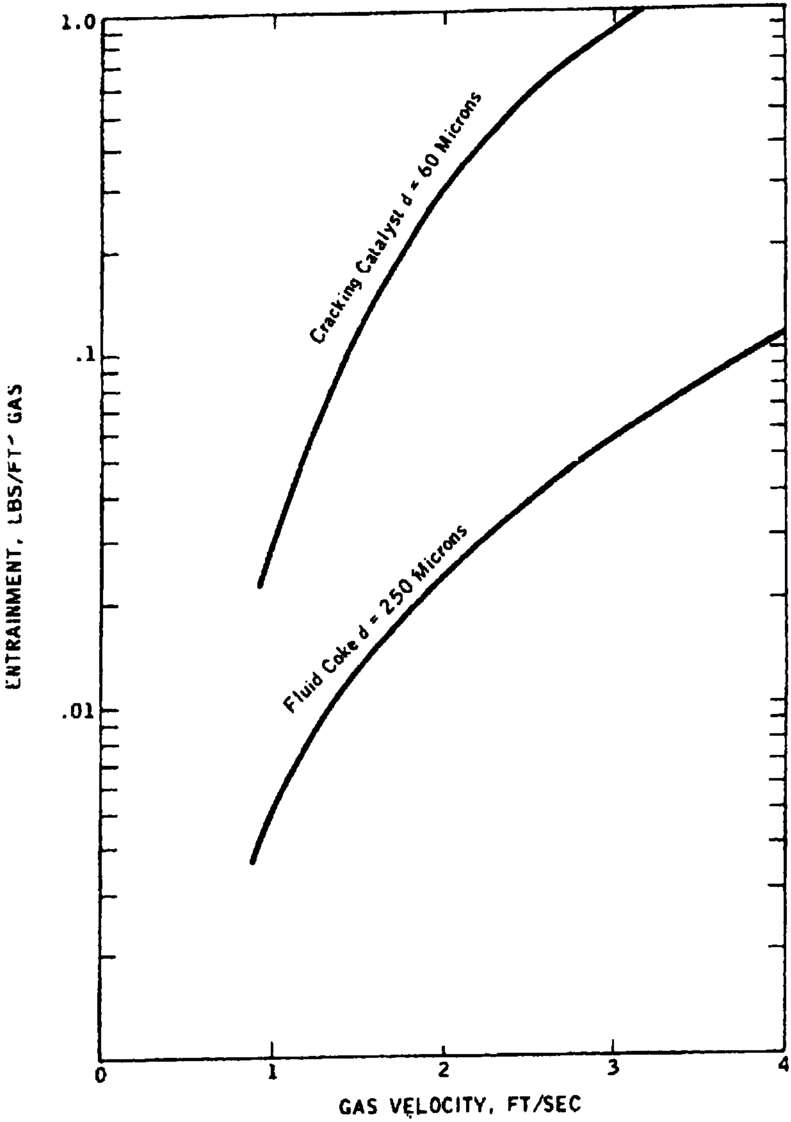


Figure 8. Typical entrainment.

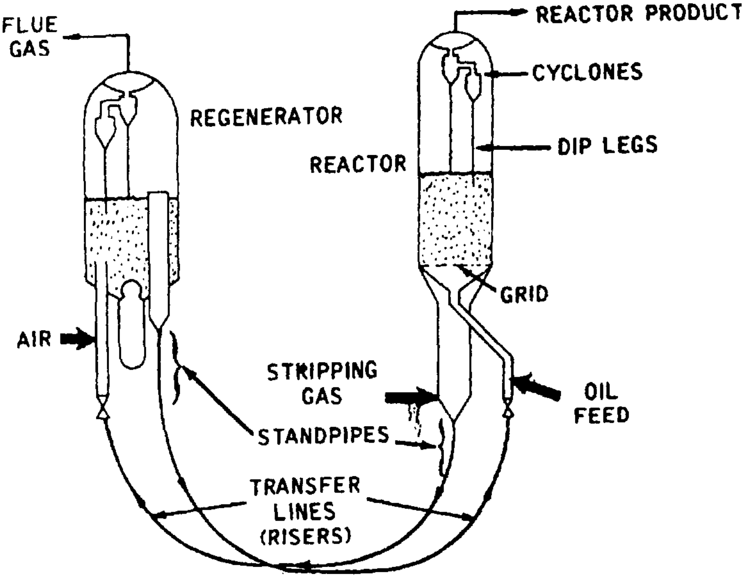


Figure 9. Fluid solid system model IV catalytic cracking unit.

which particles reach the surface. This in turn depends on the size of the vessel, the geometry of the heat transfer surface, and the quality of fluidization.

DESIGN CONSIDERATIONS FOR FLUID BED SYSTEMS

Many factors enter into the design of a fluid bed reactor which are unknown in more familiar reactor types. These can be illustrated with reference to fluid catalytic cracking.

A typical cat cracker is shown in Figure 9. Gas oil is fed to the reactor (right hand vessel). It contacts the hot finely divided catalyst in the reactor, and is vaporized and broken down catalytically into smaller molecules mostly in the gasoline boiling range. The product gases leave the fluid bed, and after passing through cyclones which remove entrained catalyst, leave the reactor and are sent to fractionation. The reaction is endothermic, and it deposits a carbonaceous residue on the catalyst particles which reduces the catalytic activity. The catalyst is therefore continuously removed for regeneration and reheating. It passes down through the reactor stripper in which unconverted hydrocarbons are removed and then through a standpipe and transfer line to the regenerator vessel. Here air is used as the fluidizing gas, and the coke deposits are burned off the catalyst particles. The burning is highly exothermic and provides the heat necessary for the cracking reaction. The hot regenerated catalyst is then returned to the reactor.

Particle Size Selection

When particle size can be set by the designer, the optimum balance between good fluidization and low levels of entrainment usually dictates a size in the 40-80 micron range, as is used in cat cracking. In a case such as Fluid Iron Ore Reduction (FIOR), the particles are reactants, and costs of crushing iron ore dictate the use of much larger particle sizes. In fluid coking, good fluidization and contacting are not very critical, and particles of 200-1000 microns are used. For fluid bed boilers, very high gas velocities may be necessary for economically sized vessels, and particles several millimeters in diameter might be considered.

Bed Design

Permissible gas velocities are usually set by entrainment, and for a given throughput the vessel diameter is thus determined. The amount of catalyst or other bed particles is set by reaction kinetics and the bubble-solids contacting expected. Very often there is a scale-up debit involved in fluid bed reactors. As mentioned earlier, small reactors are often in slug flow, where reactor diameter determines fluidization behavior, and in such units bed diameter limits bubble size. Large reactors thus give bigger bubbles and less gas-solids contacting than small ones. Some large reactors are subject to *gulf stream* flow in which most of the gas passes up through the center of the bed and solids circulate rapidly up the center and down at the walls. Gulf streaming can also cause scale-up debits because it permits bubbles to channel very rapidly up through the center of the bed, again giving poor contacting. Fortunately in cat cracking, bubbles are usually small, and gulf streaming is normally absent so that scale-up problems are not as critical as in some less well behaved processes. With the amount of catalyst known, bed height can be determined from the bed density relationships outlined earlier. Bed height is often controlled by overflow wells for particle removal.

Grid Design

The first requirement of the grid is that it distribute gas evenly across the fluid bed. This is accomplished by having a large number of holes and a high pressure drop across the grid. The high pressure drop assures that flow will be uniform through all holes and that flow stability will not be much influenced by behavior of the fluid bed.

Often grid design is primarily dictated by the peculiarities of the specific process under consideration. Some typical problems which have dictated grid design include erosion or fouling due to particles carried up through the grid by fluidizing gas, dumping from one bed to a lower one, formation of reaction deposits on upper surfaces, and physical support of a large bed at very high temperatures.

Outage and Cyclone Selection

The vertical distance between the bed surface and the inlet to cyclones is known as outage. Two factors usually determine outage. First of all, entrainment is a strong function of outage up to the TDH, and too low an outage can put very serious loads on the cyclone systems. Secondly, there is an appreciable pressure drop in the cyclones, which hinders the tendency of solids to flow from the cyclones back into the bed. Solids return to the bed is achieved because the solids build up hydrostatic pressure in the cyclone diplegs, and they will flow back to the reactor if the diplegs are long enough. Since most cyclones are mounted internally, the dipleg length can dictate the vessel height and outage.

Perhaps the most important concern in cyclone selection is that they have the required collection efficiency. In a large cat cracker, two stages of cyclones in series must have a combined efficiency of 99.99% to keep catalyst losses to an acceptable level of 5 tons per day. Efficiency can be increased by higher gas velocity, by smaller cyclone diameter, and by selection of high efficiency geometry rather than a high throughput type.

Minimum cyclone diameter is about 3 ft. based on maintenance considerations. Most cyclones are installed inside reactor vessels, and problems of geometric layout of multiple cyclones also can affect the size and hence the number of cyclones selected.

The penalty of higher efficiency is usually higher pressure drop. Pressure drop through cyclones is of some concern because of gas compression costs. A more critical reason is usually the effect on dipleg operability and length mentioned previously.

Standpipe and Riser Operation

Catalyst is transferred between reactor and regenerator vessels through standpipes and risers. Particles flow out of a vessel downward through a standpipe, which is aerated with a small amount of gas. Bulk density in the standpipe is fairly high, and the hydrostatic pressure produced at the bottom is substantial. The particles then go through a U bend and up through a transfer line. The transfer line is aerated with a high gas flow, and this causes the density there to be much lower than in the standpipe. This resulting pressure inbalance causes the desired catalyst flow. Accurate pressure balances between vessels across the standpipe/riser loops are essential to proper design of a fluid solids unit.

4

CATALYTIC REFORMING

INTRODUCTION

With the advent of catalytic reforming, the thermal process became known as thermal reforming to distinguish it from the catalytic process. The use of a catalyst results in much higher octane levels and yields than can be obtained in thermal reforming. In addition, significant amounts of valuable hydrogen are produced as a by-product. Hydrogen is in great demand in refineries, and the catalytic reformer is often the only source available for the many important refining hydroprocesses such as hydrocracking, hydrodesulfurization, and hydrotreating. Catalytic reforming has thus displaced thermal reforming, and is in many refineries producing high octane motor fuels. The intent of this chapter is to describe the purpose, nature, and characteristics of this refining process.

THE CATALYTIC REFORMING PROCESSES

The different commercial processes fall into two major classes: platinum, and non-precious metal catalysts. The platinum catalysts contain from 0.3 to 0.8 wt% platinum on an alumina or alumina-silica base. In addition, most contain a halogen, such as chlorine, to increase activity. The nonprecious metal catalysts are also supported on alumina, but for the reactive component Fluid Hydrofining uses molybdena, TCR uses chromia, and Hyperforming uses cobalt-molybdena. The early catalyst could not be regenerated and so these units were designed for very high pressures and high recycle rates to provide long on-oil cycles. Other platinum fixed-bed processes were introduced. These processes fall into two classes: semi-regenerative, where the whole unit is shut down at intervals for regeneration, and cyclic, where an extra fixed bed reactor is provided for continuous catalyst regeneration. Powerforming is the only process which provides both types of units; Ultraforming is a cyclic process; while all the rest can be classified as semi-regenerative.

Several processes based on non-precious metal also exist. Because of high catalyst deactivation rates with these catalyst systems, they all require some form of continuous regeneration. The Fluid Hydroforming process uses fluid solids techniques to move catalyst between reactor and regenerator; TCR and Hyperforming use some form of a moving bed system.

Many refinery streams are blended to produce automobile gasoline. The main components of the refinery gasoline pool, however, are catalytically cracked and are virgin naphthas. The gas oil is catalytically cracked and the residuum is visbroken. Propylene and butylene produced in these operations are catalytically polymerized to make gasoline. Thus some means is needed to raise the octane number of this portion without undue loss in yield. The reforming operation involves thermally cracking the virgin naphthas to produce an olefinic reformate with increased octane. The nature of thermal reforming in the early days was such, however, that degredation was severe and the octane increase obtained was limited. The introduction of a catalytic process, therefore, provided a significant improvement. The advantages of catalytic reforming over the thermal process were:

- significantly higher octane levels
- much higher gasoline yields
- substantial amounts of hydrogen produced
- superior reformate quality - improved lead susceptibility, lower sensitivity, low sulfur levels, absence of diolefins which adversely effect engine cleanliness.

POWERFORMING PRODUCTS

A catalytic reforming process produces similar products. The relative amounts may differ, however, depending on the catalyst selectivity and process conditions, the main product, of course, is a high octane C_5 + gasoline fraction.

In addition to gasoline, the process produces significant quantities of hydrogen. The value of this hydrogen is high due to increased refinery application of hydrogen consuming processes. In addition to numerous past hydrotreating applications for product quality improvement, these include processes such as hydrocracking, cat feed hydrofining and fuel oil desulfurization. These processes need large quantities of hydrogen and would require expensive grass roots hydrogen production facilities if sufficient Powerformer hydrogen were not available.

Because some hydrocracking occurs, Powerforming also produces saturated C_1 to C_4 light hydrocarbons. The methane and ethane formed normally are consumed as refinery fuel. Propane and butane products are frequently marketed as LPG. The relative quantities of each of these products vary considerably with feed quality, operating conditions and octane severity.

As severity is increased, C_5+ gasoline yields decrease, with a corresponding increase in C_1 to C_4 products. Hydrogen yields increase with severity until the level at which no further aromatics are produced; as severity is increased even further, hydrogen yields then decrease.

Also of importance to the industry is the demand for aromatic products. The high octane reformate, which is rich in aromatics, is a major source of aromatics for petrochemical operations. Powerformers can produce aromatics such as benzene and toluene.

CHEMICAL CHANGES IN POWERFORMING

Powerforming is basically a conversion process in which catalytically promoted chemical reactions convert low octane feed components into high octane products. The key to a good reforming process is a highly selective dual-function catalyst. The dual nature of this catalyst relates to the two separate catalyst functions : atomically dispersed platinum to provide dehydrogenation/hydrogenation activity; chlorine on the alumina base to promote acid-type reactions. The two functions complement one another to provide highly selective conversion of paraffins and naphthenes to high octane aromatics, with a minimum amount of degradation by hydrocracking.

Major Reactions

The overall reaction scheme by which high concentrations of aromatics are produced in Powerforming is rather complex. However, it can be broken down into four important and distinct types of reactions. In Figure 1, these four reactions are illustrated by the reactions of C_6 hydrocarbon feed components.

The dehydrogenation reaction is an extremely rapid endothermic reaction which converts alkylcyclohexanes to aromatics almost quantitatively. It is promoted by the catalyst platinum function and is so rapid that it is normally limited by diffusion into the catalyst particle.

The isomerization reaction, which is acid-site controlled, includes the conversion of alkylcyclopentanes into alkylcyclohexanes, which, in turn, are quickly converted to aromatics by dehydrogenation. In addition, isomerization also includes the conversion of feed n-paraffins into higher octane I-paraffins.

Dehydrocyclization refers to the conversion of feed paraffins into alkylcyclohexane and alkylcyclopentane naphthenes. These, in turn, are subsequently converted by isomerization and dehydrogenation into aromatics. Dehydrocyclization is controlled by both acid and platinum functions and is the most sensitive indicator of catalyst selectivity.

Hydrocracking involves the rupture of paraffin C-C bonds to yield lower molecular weight paraffins. Although the cracking reaction is largely acid-site controlled, the catalyst platinum function is involved in the rapid hydrogenation of the olefins produced by the cracking reaction. Since hydrocracking converts gasoline components into gas, it is the least desirable reforming reaction, and reforming conditions are normally selected so as to minimize the extent to which hydrocracking occurs.

All of the above reactions are reversible, with the exception of hydrocracking, so that thermodynamic equilibrium limitations are important considerations. To the extent possible, therefore, operating conditions are selected which will minimize equilibrium restrictions on conversion to aromatics. This conversion is favored at higher temperatures and lower operating pressures.

Figure 1 shows that the valuable by-product hydrogen produced in reforming comes largely from alkylcyclohexane conversion to aromatics. The alkylcyclohexanes are either introduced with the feed or are obtained by conversion of feed alkylcyclopentanes and paraffins. Some additional hydrogen is produced from the paraffin conversion to naphthenes as well. Hydrocracking, on the other hand, results in the consumption of hydrogen. Thus feeds and operating conditions which provide increased conversion to aromatics, likewise promote higher hydrogen yields.

Aside from the above reforming reactions, a small amount of feed components are converted to polymeric hydrogen deficient products which deposit on the catalyst as "coke." A coke buildup results in activity and selectivity loss which ultimately requires catalyst regeneration. In semi-regenerative operation, the coking rate is maintained at a low level to provide cycles of at least three to six months. In cyclic units, coking conditions are inherently much more severe so that frequent regenerations are required.

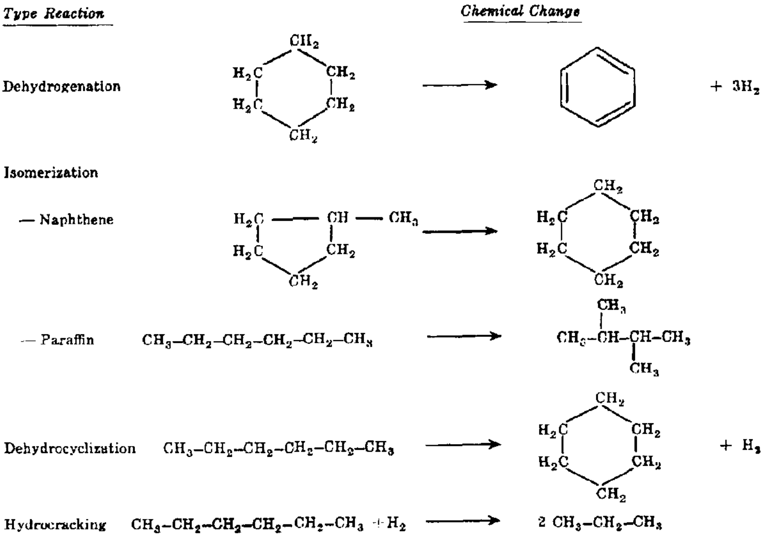


Figure 1. Major Powerforming reactions.

MAJOR POWERFORMING VARIABLES

The extent to which each of the above reactions occur is strongly influenced by feed quality and the levels selected for the major process variables: pressure, temperature, recycle rate, and frequency of regeneration. From a process viewpoint, these variables affect catalyst requirement, gasoline yield, and coke make.

Pressure

An increase in operating pressure increases hydrocracking, decreases dehydrocyclization and provides a less favorable equilibrium for conversion to aromatics. This results in a significant reduction in gasoline yields and somewhat lower conversions. Catalyst requirements, therefore, increase. However, coke make, and hence catalyst deactivation, is greatly reduced, so that, where long cycles are desired, high pressure units are favored. Most units operate in the pressure range of 250 to 500 psig or higher.

Temperature

A typical Arrhenius dependency of conversion on temperature is found in Powerforming; reaction rates are found to double about every 28°F. Thus at higher temperatures, catalyst requirements can be significantly reduced. As temperature is increased, furthermore, the desired conversion of paraffins to aromatics is enhanced, resulting in higher gasoline yields at a given octane severity. This improvement arises from a more favorable thermodynamic equilibrium and also because the dehydrocyclization reaction is favored at higher temperatures. Coke make, however, is also strongly dependent on operating temperature; coking rates may double every 20-25°F. Operating temperatures are thus largely selected on the basis of economic considerations. Typically, average operating temperatures are in the range of 880 to 960°F.

Recycle Gas Rate

A large quantity of hydrogen-rich separator gas is normally recycled with the feed stream. Recycle rates may vary from 2,000 to 10,000 MSCF/B. The recycle gas serves to suppress catalyst coke make but normally has relatively little direct effect on gasoline yields or catalyst requirement. However, at lower recycle levels, where an increase in recycle rate may significantly increase reactor hydrogen partial pressure, the effect is similar to a small increase in total pressure, i.e. yields are somewhat reduced and catalyst requirement slightly increased. The control provided by recycle rate over deactivation provides considerable flexibility in unit designs for operations at temperatures and pressures which would otherwise not provide feasible on-oil cycles.

Regeneration Frequency

The design frequency of regeneration is normally from three to six months for semi-regenerative units, and one reactor every 24 hours in cyclic units. For either case, an increase in regeneration frequency would result in a reduction in average catalyst coke level. Thus, gasoline yields would increase and catalyst requirements decrease.

Feed Quality

Powerformer feed quality can vary over a wide range. Feed paraffinicity, for example, can vary from 25%, for high quality hydrocrackate feeds, to 75% for Middle East virgin feeds. Such wide variations in feed quality can have a great effect on process performance. An increase in feed paraffinicity normally results in a lower yield of aromatics, so that catalyst requirements increase and gasoline yields decrease. Coke make, however, tends to be somewhat lower. Feed end point also can have an important effect. As end point is increased, feed quality improves significantly in that catalyst requirements decrease and yields increase. Higher end point feeds, however, give appreciably higher coking rates.

ENGINEERING ASPECTS

To appreciate the engineering aspects of Powerforming, it would be helpful to go over some of the engineering design considerations.

1. As noted above, the reaction is carried out at about 930°F. This means the reactants must be heated to this temperature and the products subsequently cooled. Thus a large section of any catalytic reformer is heat exchange equipment in the form of exchangers, furnaces, and condensers. In a 20,000 B/SD unit over 100 million BTU's are transferred per hour. The furnace design and the process heat integration are therefore important engineering considerations in optimizing the design of units.

2. The Powerformer reaction absorbs heat; it is largely an endothermic reaction. This heat of reaction is in the order of 200-350 BTU/pound, depending on the type of feed. Since the individual Powerforming fixed bed reactors operate

adiabatically, the high endothermicity of the Powerforming reactions, particularly in the lead reactors, results in a large temperature drop. As temperature falls, the reforming reaction rate decreases. Thus to maintain a relatively high rate of reaction and keep catalyst requirements at a minimum, intermediate reheat is required between each of the reactors. The more such stages, the higher the average catalyst temperature for given inlet temperatures. An increase in recycle gas rate also reduces the temperature drop because of the increased sensible heat. Ultimately, selection of the number of reactors must be based on economic considerations for the specific unit under design.

3. Compression costs in Powerforming are high. Recycle gas rates ranging from 2,000-10,000 SCF/B of feed must be compressed to overcome pressure drop in the heat exchangers, furnaces, and reactors. When regeneration is employed, considerable amounts of air also must be compressed from atmospheric to regeneration pressure. In a typical 20,000 B/D Powerformer with 7 MSCF/B of recycle gas, a 5,000 brake horsepower compressor is required. Both investment and operating costs for the compressor thus contribute significantly to the unit costs. The selection of recycle rate is, therefore, also an important design consideration. Normally, the higher the catalyst requirement, the higher the recycle rate that can be justified on economic grounds.

THE POWERFORMING PROCESS

Powerforming is a version of the platinum reforming process. Basically there are two types of unit: semi-regenerative and cyclic. The choice of unit and the exact process conditions used will depend to a large extent on the particular application. However, certain generalizations regarding each of the two types can be made.

Semi-Regenerative Powerforming

As mentioned above, semi-regenerative implies infrequent catalyst regeneration. The unit is on stream for two or more months between regenerations. When it is necessary to regenerate the catalyst, the entire unit is shut down. The long on-oil cycles in this type of operation normally require either higher reactor pressures, higher recycle rates, or lower temperatures in order to limit catalyst deactivation. However, semi-regenerative Powerforming process flexibility has been extended so that pressures as low as 300 psig are feasible for conventional mogas feeds, and 250 psig designs are used for light narrow boiling feeds for benzene-toluene production.

A schematic flow plan of a typical semi-regenerative Powerformer is shown in Figure 2. Starting with the feed, flow is as follows. The feed is mixed with recycle gas and partially preheated by indirect heat exchange with the product stream. Final preheat is obtained in the preheat furnace. The temperature to which the feed and recycle are heated in this furnace depends on the condition of the catalyst. With a fresh or freshly regenerated catalyst, reactor inlet temperatures may be as low as 875°F. As coke is laid down on the catalyst, its activity decreases requiring a higher inlet temperature to maintain the same octane product. Thus temperature is slowly raised over the onstream period to maintain constant octane. The maximum reactor inlet temperature is set by rapid catalyst deactivation. This is in the neighborhood of 1000°F.

After the reactants pass through the several reactors and reheat furnaces, they are cooled by heat exchange with the feed stream. Three reactors with intermediate reheats are shown in the figure although the actual number may be 2, 3, or 4 depending on the feed and octane level desired. Final cooling is obtained in condensers. The gas-liquid stream then passes into a high pressure separator from which unstabilized reformate is withdrawn and the major portion of the gas is dried and then recycled. The excess gas, called make or tail gas, then flows to an absorber for recovery of C_4 and heavier hydrocarbons. The "wild" reformate passes to a stabilizer or similar light ends equipment.

When regeneration is required, the entire unit is shut down. The unit is purged with inert gas and the recycle gas compressor is used to circulate flue gas during regeneration to control the temperature rise. The only additional equipment required for this type of regeneration is an air compressor and facilities for chlorine injection. The flame front temperature during regeneration is controlled to less than 1000°F by limiting oxygen content to less than 1%. When the flame front has passed through the reactor bed indicating that the carbon has been completely burned from the catalyst, oxygen content of the flue gas is increased to full air and circulation continued. Following this period, water and chlorine are added to the oxygen containing circulating gas stream in order to replace chloride which has been stripped from the catalyst during reaction and regeneration. This chlorine treat is necessary in order to fully restore catalyst activity. The chlorine serves to redistribute agglomerated platinum formed during regeneration and to impart a certain amount of catalytic activity to the catalyst itself. Following the chlorine treat, circulation of the oxygen containing gas is continued for several hours to assure complete platinum dispersion. The unit is then purged with an inert gas. Subsequently a hydrogen containing gas is circulated to reduce the water content of the catalyst to a minimum. The system is then repressured and oil feed returned to the unit.

The common practice in semi-regenerative Powerformers is to desulfurize the reactor feed. This reduces the deactivation rate of the Powerforming catalyst,

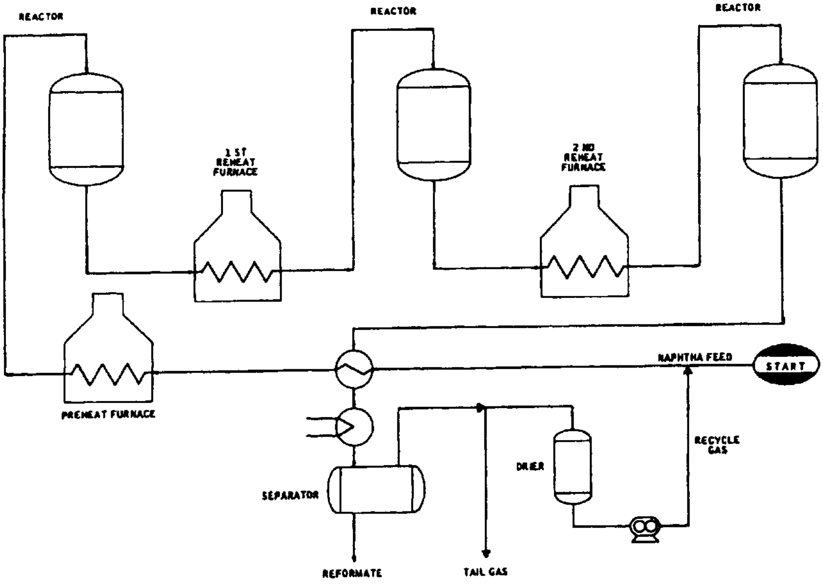


Figure 2. Semi regenerative power.

and also serves to reduce unit corrosion which would otherwise result from the formation of H_2S during the Powerforming reaction. Elimination of corrosion is extremely important because of scale formation in furnaces and transfer lines which flakes off and tends to plug up the catalyst beds.

When the feed is desulfurized, the Hydrofining unit is commonly integrated with the Powerformer to conserve heat. The stripper on the hydrofiner product which removes H_2S is combined with the Powerformer absorber used to recover C_4+ from the tail gas. This tower is termed an absorber stripper. Powerformer tail gas is used to strip H_2S from the hydrofiner product. The hydrofiner product in turn serves to absorb the C_4 + from this tail gas and return it to the Powerformer.

Cyclic Powerforming

The equipment used for cyclic Powerformers is quite similar to that just described. However, the cyclic design provides an extra, or swing, reactor so that catalyst can be regenerated while the unit remains on stream. The swing reactor is so manifolded to the system that it can replace any of the other reactors so that they may be regenerated. A flow plan is shown in Figure 3. Again three reactors on oil are indicated although 3, 4 or even 5 may be used. The ability to regenerate during operation means that relatively short cycle lengths are possible. Usual cycle lengths are about 60-80 hours. The short cycles mean that we can tolerate a more rapid catalyst deactivation rate. Thus cyclic Powerformers normally run at lower pressures, lower recycle rates and higher temperatures than do semi-regenerative units.

The regeneration procedure employed is similar to that used in the semi-regenerative process except that separate regeneration facilities are required. This includes the swing reactor, the necessary manifolding, a recycle flue gas compressor, an air compressor and flue gas scrubber. When a reactor is to be regenerated, it is taken off stream and replaced by the swing reactor. It is first purged with recycle gas and then with flue gas from an inert gas generator. After purging, flue gas is circulated through the regeneration equipment and air is added. Flame front temperatures are limited to 1000°F, as in the case of semi-regenerative operation. After the carbon burns, oxygen content of the flue gas is increased to 6% and the catalyst is dried, chlorine treated and purged as in the case of semi-regenerative operation before it is replaced on stream.

Requirements for desulfurizing the feed are much the same as in the semi-regenerative process. While the continuous regeneration facilities would permit operation with higher sulfur feeds, extensive H_2S corrosion results, making feed

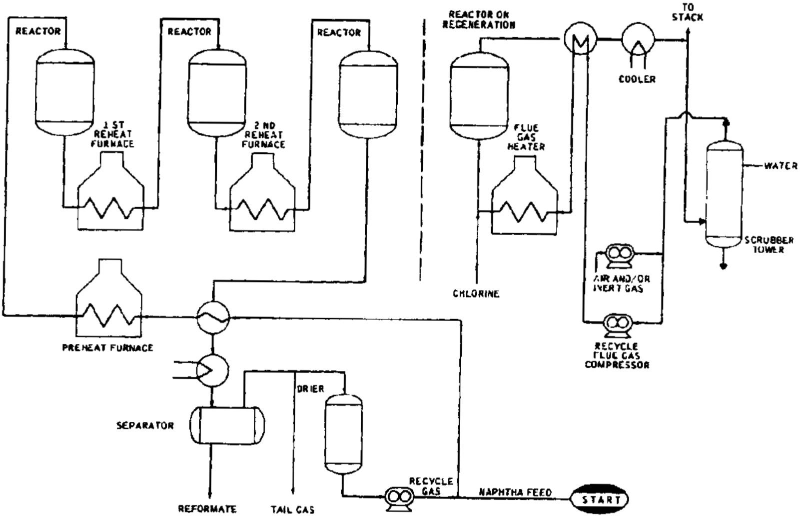


Figure 3. Cyclic Powerformer.

desulfurization an economical way to reduce unit maintenance costs.

POWERFORMING PROCESS SELECTION

There are many factors to be considered in selecting the type of Powerformer to use. It largely depends on the local situation and includes such things as feed stock, octane level, the value of gasoline yield, the use for by-product hydrogen, plant size, etc. While we can not completely generalize, we can mention some of the more important considerations.

For mogas operations, the specific application may favor either a cyclic or semi-regenerative unit. Because of their greater inherent tolerance to high coking rates, cyclic units are more widely used for high end point feeds and high octane severities. Operating conditions are also more severe, including lower pressures, lower recycle rates, and reduced catalyst loadings. In semi-regenerative units, while lower pressures are feasible, they are achieved at the expense of higher recycle rates and/or higher catalyst loadings. Highest C_5+ and hydrogen yields are thus achieved in cyclic units, but at the expense of somewhat higher plant investments for the extra reactor and additional manifolding and valving.

Unit size is also an important consideration. For large units, the higher catalyst requirement for semi-regenerative units becomes a larger percentage of total investment so that cyclic units are more attractive. This is especially true for paraffinic feeds and high octane severities. Ultimately, the selection of cyclic versus conventional semi-regenerative and low pressure semi-regenerative Powerforming requires an economic evaluation. The higher hydrogen yields and the availability of an uninterrupted supply of high purity hydrogen tail gas from a cyclic unit is an important credit in applications. The incremental costs for a cyclic Powerformer normally cannot be justified, and a low cost conventional pressure semi-regenerative unit is the usual choice.

A low pressure semi-regenerative unit provides yields that approach those obtained in cyclic Powerforming but with savings in investment. For smaller units and moderate severities, the low pressure semi-regenerative unit would likely be the most attractive alternative if selectivity improvement credits are sufficiently high.

For aromatics production, similar considerations apply. Maximum yields of xylenes and other heavy aromatics can be obtained in cyclic units, but, again, at somewhat higher investments. The process selection, thus, again requires the balancing of process credits versus debits for the specific application. For light aromatics (benzene-toluene) production, however, the situation tends to favor a low pressure semi-regenerative unit. Coking rates for these light feeds are so low that semi-regenerative pressures can be reduced markedly to provide yields that compete favorably with those obtained with the more expensive cyclic units.

There are several versions of the two-step process. One involves the use of 5A molecular sieves in conjunction with Powerforming. In this combination process, the 5A sieves may be used to remove n-paraffins from the feed to improve feed quality. Alternatively, the Powerformate may be sieved to remove the low octane n-paraffin components. Extraction of aromatics from Powerformate is another interesting two-step process.

5

FUELS HYDROTREATING

INTRODUCTION

Hydrotreating is a process which improves the quality of a variety of petroleum stocks by treating with hydrogen in the presence of a catalyst. Hydrotreating may be applied to a variety of solvents, distillate fuels and residual fuels. When referring to residual fuels, the process is termed Hydro-desulfurization (HDS) since the sole object is sulfur removal. When treating other stocks, the process is referred to as Hydrofining. When treating stocks other than residual fuels, depending on the precise feed and purpose of the operation, Hydrofining will improve the odor, color, stability, combustion characteristics and other important quality characteristics. It also removes sulfur, nitrogen and other nonhydrocarbon components. When applied to catalytic cracking feed stocks, Hydrofining significantly improves cracking quality. Carbon yield is reduced, gasoline yield is increased, and the quality of the catalytic cracking products is significantly better. The need for low sulfur residual fuel oils to alleviate the air pollution problem has led to the development of the required hydrotreating technology. The subsequent chapter describes the Hydrotreating process.

THE HYDROTREATING PROCESS

Figure 1 shows a schematic diagram of a Hydrotreater. Feed stock is mixed with a hydrogen containing gas and heated to reaction temperature in a furnace and passed to the top of a reactor (s). The reactor contains the catalyst in the form of extrudates or pills. The oil and hydrogen containing gas pass downward through the reactor. Depending upon the feed stock and operating conditions, all of the oil may be vaporized or as much as 80-90% may remain in the liquid phase. In

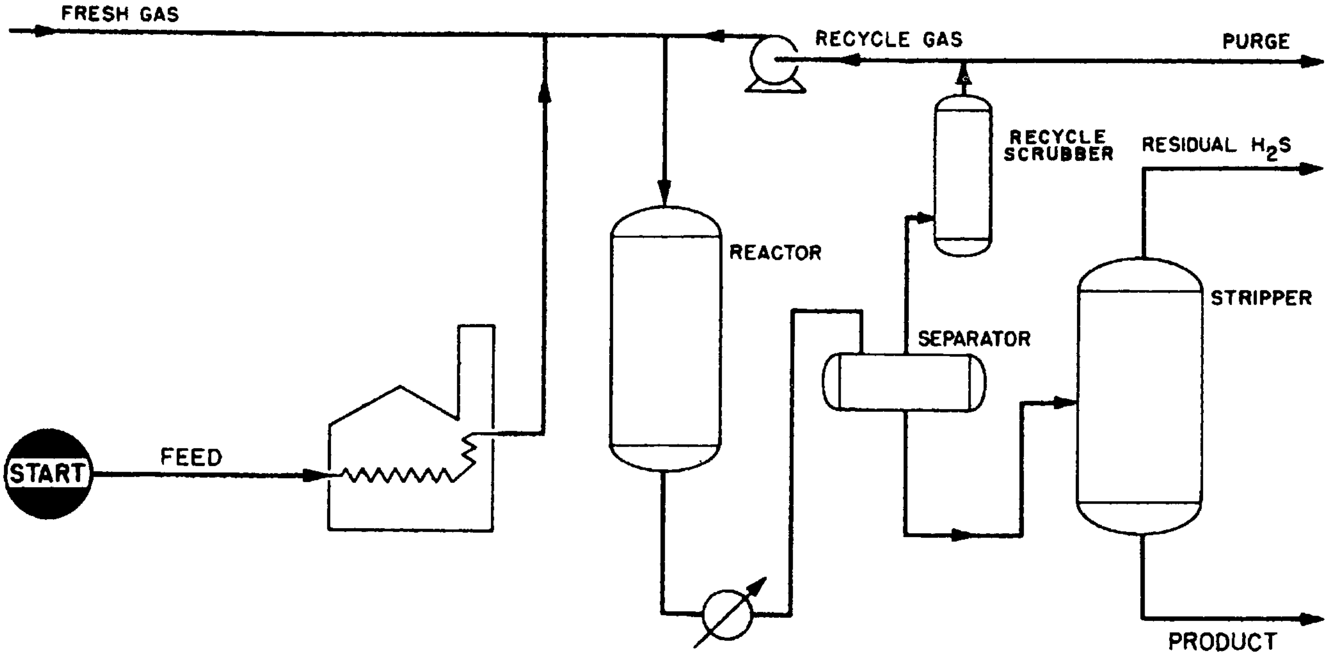


Figure 1. Schematic flow plan-hydrotreater.

the case of residual fuel oils, the feed remains essentially all liquid. The reactor effluents are cooled and passed to a gas-liquid separator wherein the hydrogen containing gas is separated and recycled to the feed for reuse. The recycled gas is usually scrubbed to remove the H_2S. This is done because of the inhibiting effect of H_2S on the kinetics of hydrotreating and also to reduce corrosion in the recycle circuit. Sometimes, when treating a light stock with a very low sulfur content, the recycle gas is not scrubbed because the H_2S is at an acceptably low level. The liquid is passed to a stripper to remove residual H_2S and other light gases; then it may be fractionated into several cuts. In many cases, the liquid products are given a light caustic wash to assure complete removal of H_2S. Small quantities of H_2S, if left in the product, will oxidize to free sulfur upon exposure to air, and will cause the product to exceed corrosion specifications.

The Hydrofining process is actually one of many processes that exist, but all are very similar in nature.The main difference in the various processes is in "know how". Each process differs by catalysts, equipment and/or methods, but these are rather narrow since the general field of hydrogenation is an old and well established art.

Figure 2 shows three different types of reactions that occur during Hydrofining. Group I shows hydrodesulfurization of four sulfur types: mercaptans, disulfides, thiophenes and benzothiophenes. The mercaptans and disulfide types are represenative of a high percentage of the total sulfur in lighter virgin oils, such as virgin naphtha and heating oil. Thiophenes and benzothiophenes appear as the predominant sulfur form in heavy virgin oils and even more in cracked stocks of all boiling ranges. By a fair margin, thiophenic sulfur is the hardest to remove.

Group II shows the reactions of oxygen compounds. Phenols (and thiophenols) occur in catalytic cracking products. Peroxides are often found in cracked stocks after exposure to air. Oxygen compounds and poor storage stability go hand in hand. Hydrofining provides stable and clean burning fuels, and the Hydrofinates are almost always free of oxygen compounds.

While not shown here, Hydrofining also removes nitrogen from various nitrogen compounds. Nitrogen is one of the causes of instability. Removal of nitrogen is much more difficult than sulfur removal.

Group III shows the Hydrofining reactions of a straight chain monoolefin and a diolefin. These reactions are typical for all kinds of olefins. Diolefins and certain or all of the cyclic olefins are found to be most reactive with respect to gum formation. Thus, Hydrofining improves stability by removing (saturating) reactive hydrocarbons as well as oxygen containing compounds.

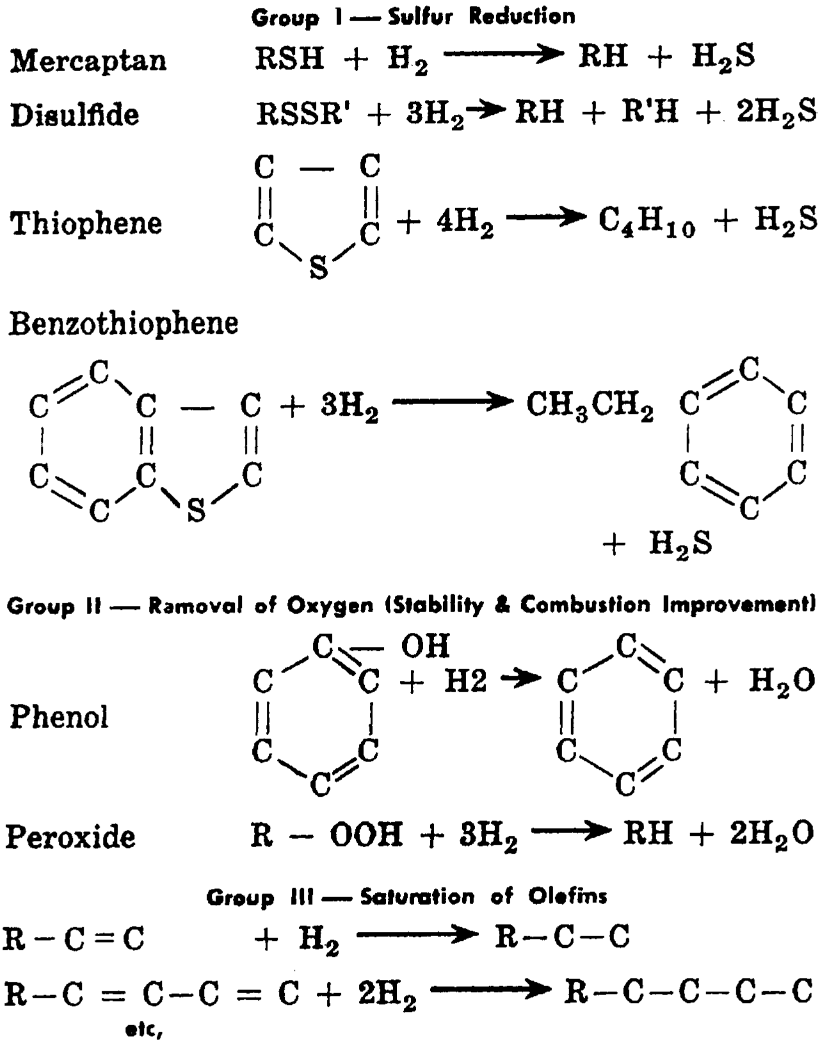


Figure 2. Typical hydrofining reactions.

Saturation of olefins other than reactive olefins usually is not desired. The added hydrogen is often expensive or useful elsewhere, and it does not provide any real improvement in product quality. Actually, product quality may be reduced in the case of gasolines. Research octane number losses may be correlated with increasing olefin saturation. So in many cases, hydrodesulfurization conditions are selected with an eye toward minimizing olefin saturation over and above that needed for product quality improvement. There is one exception: saturation of certain olefins shows substantial improvements in Motor octane number. This is true for iso- and n-pentenes and to a lesser extent for higher boiling isoolefins. The higher n-olefins show octane losses upon saturation.

The ranges of operating conditions that are used in Hydrotreating vary significantly mainly because of the very broad application of Hydrotreating. The ranges are wide due in part to the fact that light products such as naphthas require much lower treating severity than that required to desulfurize gas oils or residual fuel oils. Within a given boiling range, say heating oil, treating conditions can vary depending on the nature of the stock, virgin or cracked, and the specific purpose of the Hydrofining operation, sweetening, deep desulfurization, or improvement in burning characteristics (lowering carbon residue on 10% bottoms). When treating virgin naphtha, for example, 99% desulfurization can be obtained at conditions of 550°F, 4 V/hr/V feed rate, 400 psig and 1000 SCF/B treat gas. These same conditions applied to diesel oil would give only about 25% desulfurization although the carbon residue would show adequate improvement. Additionally, if the temperature for the diesel oil were increased to 700°F, which is approaching the maximum allowable from a catalyst deactivation standpoint, the desulfurization would be about 85% (the other conditions being held constant).

The discussion that follows will show the effects of several operating variables on product inspections. The effects of the variables are illustrated best by deep desulfurization of heavier gas oils.

Effect of Feed Rate: The effect of feed rate on hydrodesulfurization of vacuum gas oil is shown in Figure 3. Halving the feed rate in this range approximately halves the product sulfur.

Effect of Pressure: Figure 3 shows the effect of pressure on product sulfur. In the 400-800 psig range, doubling the pressure reduces the product sulfur by about one third. Pressure also has an effect on catalyst life. In general, as the pressure is increased the catalyst deactivates at a lower rate. However, beyond a certain point, further increases in pressure have only a small effect on deactivation rate. An example of this is for atmospheric resids; typical data

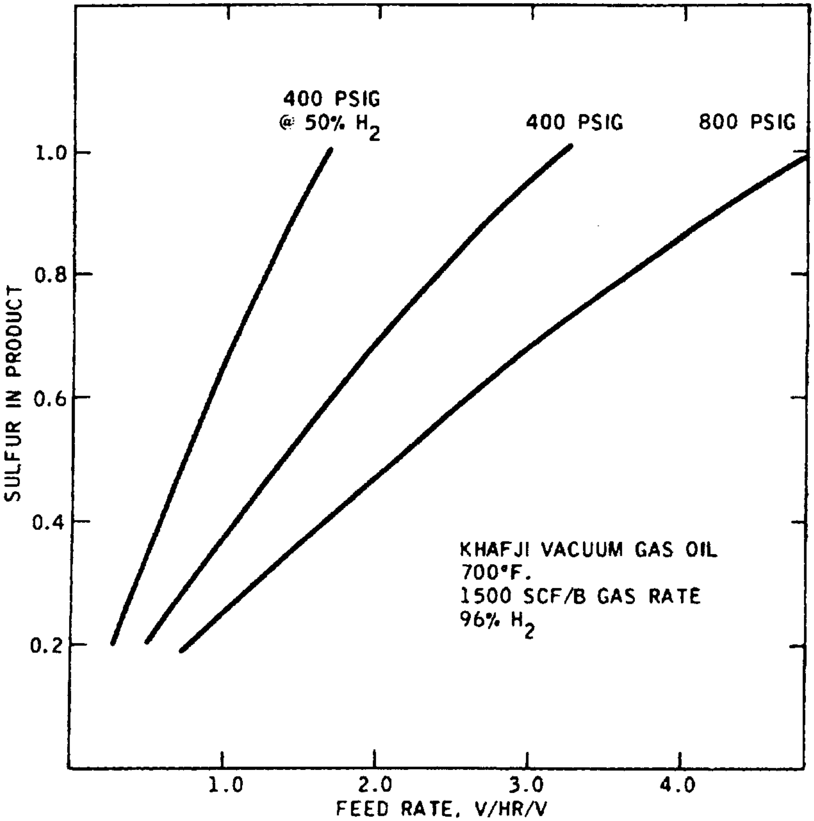


Figure 3. Effect of feed rate on product sulfur.

indicate about the same deactivation rate at pressures of 800 psig and higher (such stocks are not processed below 800 psig).

Effect of Hydrogen Concentration: The effect of hydrogen concentration is very similar to the effect of total pressure, i.e., increasing hydrogen concentration increases the sulfur reduction. Hydrogen partial pressure (total pressure multiplied by hydrogen concentration) correlates the data quite well. However, there appears to be an effect of concentration over and above its factor in partial pressure. High hydrogen concentrations, like high total pressure, improve catalyst life.

Effect of Treat Gas Rate: Treat gas rate is usually expressed as SCF/B at the reactor inlet. Very low rates provide inferior resuits and probably shorten catalyst life. Above about 1.5-3 MSCF/B, changes in rate do not usually change results. The effects of gas rate, if any, are probably related to a reduction in hydrogen concentration as the gas passes through the reactor. The reaction consumes hydrogen and manufactures light hydrocarbon diluents. At high gas rates, changes in concentration are quite small and indiscernible. At low gas rates, serious drops in hydrogen concentration occur and product sulfur rises because of this loss in concentration as shown earlier. The effect of gas rate is also probably related to the reduction in H_2S concentration as gas rate is increased. Hydrogen sulfide is a product of the hydrodesulfurization reactions and has an inhibiting effect. Since the H_2S formed is fixed, the concentration falls off as the gas rate increases.

Effect of H_2S, Carbon Oxides, Etc.: Hydrogen sulfide in the treat gas has an inhibiting effect on the kinetics of hydrotreating. Being a product of the desulfurization reactions, H_2S must diffuse from the catalyst surface into the bulk gas stream. Any H_2S present beyond that formed, further slows down the rate of diffusion with a consequent decrease in the amount of desulfurization for a given amount of catalyst. Therefore, additional catalyst would be required.

The H_2S can be removed by a process such as MEA scrubbing of the treat gas. However, the economics must be justified for each case.

Carbon monoxide has been found to poison cobalt molybdate catalysts. It causes not only instantaneous deactivation but a cumulative deactivation as well. It should be removed from treat gas entirely or at least reduced to a very low value. Carbon dioxide also must be removed since it is converted to CO in the reducing atmosphere employed in Hydrofining. Liquid water can damage the structural integrity of the catalyst. Water, in the form of steam does not necessarily hurt the catalyst. In fact 30 psig steam/air mixtures are used to regenerate the catalyst. Also, steam appears to enhance the catalyst activity in residual fuel oil desulfurization.

The presence of oxygen in treat gas would be expected to be innocuous, since it would be expected to combine with hydrogen and form water upon contact with the catalyst. However, the presence of oxygen does degrade product color and should be removed where color is important, for example, kerosenes, gasoline, etc. The oxygen also may catalyze polymerization reactions in the preheat circuit and some of this polymer may pass through the reactor and degrade product color.

Effect of Catalyst: The catalysts used in hydrotreating are: molybdena on alumina, cobalt molybdate on alumina, nickel molybdate on alumina or nickel tungstate. Which catalyst is used depends on the particular application. Cobalt molybdate catalyst is generally used when sulfur removal is the primary interest. The nickel catalysts find application in the treating of cracked stocks for olefin or aromatic saturation. One preferred application for molybdena catalyst is sweetening, (removal of mercaptans). The molybdena on alumina catalyst is also preferred for reducing the carbon residue of heating oils.

APPLICATIONS OF HYDROFINING

As mentioned earlier, Hydrofining may be applied to a host of products to improve their quality. Subsequent paragraphs will show the results that can be obtained.

Virgin Naphtha

Hydrofining is applied to virgin naphthas mainly in the form of a pretreatment step for the feed to catalytic reformers (Powerforming). Sulfur levels of 5 parts per million (ppm) or less are required to avoid deactivation of the platinum reforming catalyst.

Virgin naphtha hydrofining processing conditions have been standardized at 550°F, 4 V/hr/V, 300-400 psig and 400-500 SCF/B of 70% H_2 treat gas. Such conditions will make a 4 ppm sulfur product of most stocks of interest.

Cracked Naphthas

On cracked naphthas, Hydrofining provides not only desulfurization, but also improvements in gum, stability, and engine cleanliness characteristics.

Hydrofining has all the advantages of acid treating without the disadvantages. For example, acid treating does not readily remove refractory sulfur compounds such as thiophene; the treated products must be rerun to remove polymers with a consequent yield loss; and disposal of the acid sludges is a serious problem.

In treating cracked stocks such as steam cracked naphtha or visbreaker naphtha, which are highly olefinic in nature, nickel molybdate or nickel tungstate catalysts are generally employed. These catalysts have much higher activity for olefin saturation reactions than does cobalt molybdate.

Solvents

Hydrofining has been applied to Varsols and various other solvents for the control of odor, sulfur, and corrosion characteristics. For example, Hydrofining of Iranian and Kuwait distillates demonstrated its effectiveness as a means of producing "White Spirit", a high-quality solvent naphtha distributed in the United Kingdom.

Kerosene

With higher boiling stocks, mild Hydrofining of kerosene effects desulfurization, color improvement, and a reduction in wick char. Hydrofining improves odor and by reducing sulfur content makes the kerosene less corrosive.

It should also be noted that this process does not alter the smoke point. Smoke point is a function of aromatics content and mild Hydrofining does not hydrogenate aromatics. To accomplish this, treating over a more active catalyst such as nickel tungstate at pressures of at least 800 psig is required.

Heating Oils

Both virgin and cracked heating oils respond to very mild Hydrofining. The process provides tremendous improvements in odor largely because of mercaptan removal. Color is also improved. As mentioned earlier, Carbon Residue on 10% bottoms is indicative of the burning characteristics of heating oils. Hydrofining improves not only the CR-10% level, but also the stability. In addition, it has been found that certain heating oils when blended show incompatibility in CR-10%. For example, the CR-10% of a blend can be higher than that on either component of a two component blend. Hydrofining corrects this incompatibility problem.

Diesel Fuel

Hydrofining is employed to desulfurize high sulfur diesel stocks, both virgin and cracked. The stability of cracked diesel stocks is also improved. In the diesel range, operating conditions become more severe. Compared to naphthas, temperatures are increased from the 550-600°F level to 700°F.

Conventional Hydrofining of diesel oils does not improve octane number because octane number improvement, like smoke point improvement in kerosenes, requires saturation of aromatics. Higher pressures are needed to gain appreciable aromatics saturation and cetane number improvement.

Heavy Gas Oils and Residual Oils

Hydrofining is also applicable to heavy atmospheric gas oils, atmospheric residua and vacuum gas oils. For the latter two stocks, the process is usually referred to as hydrodesulfurization rather than hydrofining. Hydrotreating of the gas oils (atmospheric and vacuum) improves their catalytic cracking characteristics and also produces low sulfur fuel oil blending stocks. The hydrotreating of atmospheric residua is done strictly for the purpose of producing a low sulfur fuel oil. In treating these stocks, substantially more severe conditions are required than for the lighter stocks previously discussed. Temperatures up to 760°F are employed at pressures of 800 psig and higher. Feed rates can be as low as 0.2-0.3 V/hr/V. Further, catalyst regeneration is required due to the fouling which occurs at these severe conditions. With some atmospheric residua, the fouling of catalyst is severe enough to preclude regeneration for future reuse. In such cases the entire catalyst charge is replaced at the end of each cycle.

It should be noted that the atmospheric residuum has a very high metals content, 320 ppm. This is a major factor in the difficulty of desulfurizing such stocks. Middle East residua can be fairly easily desulfurized to the 75-80% level at similar conditions. Even though the feed sulfur levels for Middle East stocks are higher (ca 4-4.5 wt% sulfur), the metals level is lower (ca 100 ppm). Hydrotreating reduces the sulfur, nitrogen, aromatic rings and Conradson carbon. The effect of this is to increase the gasoline yield in cat cracking and reduce the coke deposited on the cat cracking catalyst.

6

DISTILLATION AND LIGHT ENDS PROCESSING

INTRODUCTION

All refining operations may be classed as either conversion processes or separation processes. In the former, the feed undergoes a chemical reaction such as cracking, polymerization, or desulfurization. Separation processes take advantage of differences in physical properties to split the feed into two or more different products. Distillation, the most common of all refinery separation processes, uses differences in boiling points to separate hydrocarbon mixtures.

Distillation may be defined as the separation of the constituents of a liquid mixture by partial vaporization of the mixture, followed by separate recovery of the vapor and liquid residue. Since crude petroleum is the most complex mixture of liquids found in nature, it is not surprising that distillation is one of the most important processes in modern petroleum refining.

REFINERY EQUIPMENT

Distillation has become an indispensable part of a wide variety of refining processes other than crude distillation. Although the equipment used in these various services differs widely in appearance and complexity, it should be remembered that all distillation operations are fundamentally the same. They all depend on the fact that the liquids to be separated have different boiling points. Whenever such a mixture is heated enough to vaporize part of it, the vapor will have a higher concentration of low boiling point materials than the liquid, while the liquid will be richer in high boiling point materials than the vapor.

Another way of stating this is to say that the lower boiling material is more *volatile* than the higher boiling material. When there is only a small difference in the boiling points of two hydrocarbons (or hydrocarbon mixtures), we say that their *relative volatility* is low. When their boiling points are widely different, they have a high relative volatility. The higher the relative volatility, the easier it is to separate the components by means of distillation. Thus, the relative volatility of the desired products is an important factor in determining the complexity of the equipment required.

Another major factor affecting the choice of equipment is the boiling range of the feed stock. Crude oil is made up of hydrocarbons varying in boiling point from methane to asphalt, a range of about 1500°F. The hydrocarbons referred to as light ends (methane through butane) are so low boiling that they are gases at normal atmospheric temperature and pressure. When it is necessary to condense these materials, their boiling point is usually raised by operating the distillation unit under pressure so that available cooling water may be used. In most cases, it would be far more costly to run at lower pressures and condense the light product by means of refrigeration.

At the high end of the temperature scale, there are desirable distillate fractions whose atmospheric boiling points are well above the temperature at which cracking starts (roughly 700°F). To distill these materials without cracking, their boiling points must be lowered by reducing the oil partial pressure. This is accomplished by inducing a vacuum on the system, by diluting the oil vapors with inert gas (usually steam), or by using both methods simultaneously.

The stability of the stock with respect to heat determines the maximum temperature which can be employed in the distillation process. The complexity of the process is also affected by the product purity demanded by product specifications or by the limitations of subsequent processing equipment. The relative values of the several products determine how much money can be spent in designing the unit for increased yields of the more valuable products.

Atmospheric Pipe Stills

A single-stage pipe still used for distilling whole crude is shown in Figure 1. The unit shown is designed to separate six products: gas, overhead distillate, three sidestream distillates, and undistilled residue, or bottoms. Modern pipe stills usually have several sidestreams.

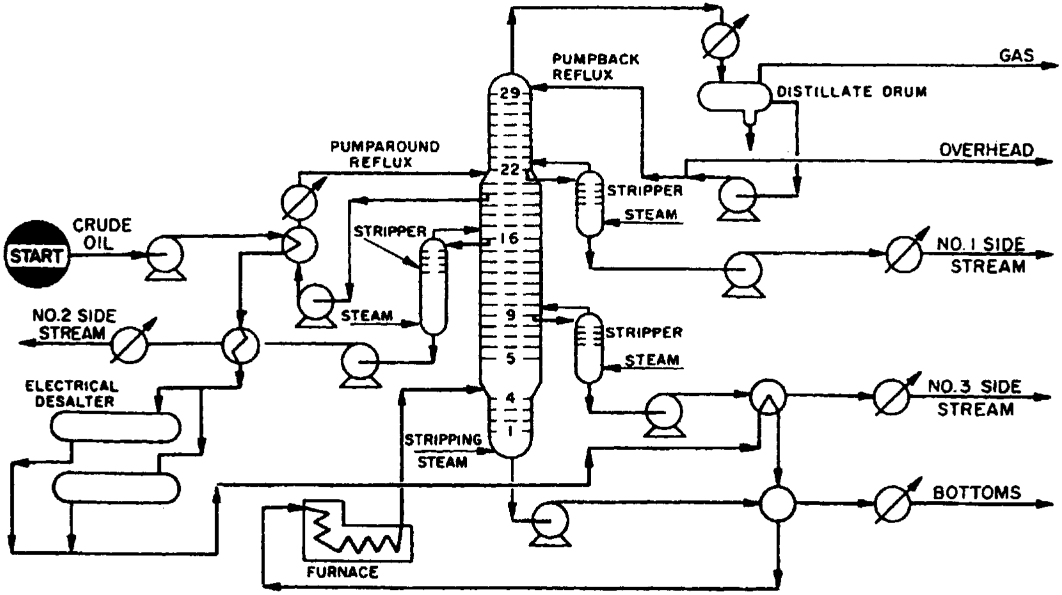


Figure 1. Atmospheric pipe still.

Generally, an atmospheric pipe still is operated at slightly above atmospheric pressure, although pressures as high as 30 psig have been used. The distillate drum temperature is determined by the temperature of the cooling water used in the overhead condenser; normally the drum runs at about 100°F. The volume of gas produced varies with the operating pressure, the character of the crude, and the boiling range of the overhead distillate. In some cases, all of the vapor leaving the tower is condensed and no gas is produced.

The overhead distillate is most frequently a full boiling range naphtha (up to about 400°F final boiling point) with the first sidestream product being a kerosene or jet engine fuel cut. The second sidestream on the pipe still shown in Figure 1 might be used for diesel fuel or heating oil. The third sidestream might be a heavier diesel fuel or gas oil for feed to a catalytic cracking unit. An atmospheric pipe still is normally designed with enough flexibility to process more than one crude or for more than one distribution of products from a single crude. Additional sidestreams contribute to this flexibility and facilitate the production of narrow boiling range stocks for special uses. The bottoms product in the pipe still shown would probably be blended into heavy fuel oil.

The temperature at which the crude enters the tower is usually determined by one of the following considerations:

1. When kerosene is one of the distillate products, the color stability of this stock has limited the flash zone temperature to as low as 650°F in older designs to avoid incipient cracking. Cracking is a function of both time and temperature, and in new designs this limit has been increased by carefully designing the unit to minimize liquid residence time at the high temperature points of the system. If the kerosene product is upgraded in quality by subsequent hydrotreating, the flash zone temperature can be as high as 725°F or more.
2. In single-stage units which do not produce kerosene or other critical stocks, flash zone temperatures may be as high as 750°-775°F. The principal limitation is the point at which cracking of distillates to less valuable gas or the rate of coke formation in the furnace tubes becomes excessive.
3. In two stage units, it is often economical to distill more gas oil in the vacuum stage and less in the atmospheric stage than the maximum attainable.

Gas formed in the atmospheric tower bottoms piping at high temperatures tends to overload the vacuum system and thereby to reduce the capacity of the vacuum tower.

The volume of crude vaporized at the flash zone is approximately equal to the total volume of distillate products. Of course, the vapor at this point contains some undesirable heavy material and the liquid still contains some valuable distillate products. The concentration of heavy ends in the vapor is reduced by contact with liquid on the trays as the vapor passes up the tower. This liquid reflux is induced by removing heat farther up in the tower.

Reflux from the first tray above the flash zone, along with the flashed crude, enters a smaller diameter stripping section at the bottom of the tower. Here, open steam injected below the bottom plate rises through the descending liquid, stripping out the lower boiling hydrocarbons. These vapors pass up through the flash zone to the fractionating section of the column.

Liquid products are withdrawn from plates 9, 16, and 22 as sidestreams. These are contacted with steam in external stripping columns having 4 to 6 plates in order to remove light dissolved hydrocarbons. Since the liquid withdrawn from any plate in a pipe still is in equilibrium with all the lighter vapor passing that plate, it always will contain material which is lower in boiling point than the desired product.

In the example shown, part of the heat is removed by returning cold overhead distillate to the top of the tower where it is revaporized by the heat supplied by condensation of part of the rising vapors. The remaining heat is removed by circulating a liquid *pump- around* stream over plates 20 and 21. This stream is heated by condensation and cooling of the rising vapors, leaves the tower to give up its heat to the feed in an external heat exchanger, then returns to the tower at a higher point.

It would be possible to remove all of the heat by pumping cold reflux from the distillate drum to the top of the tower and thus eliminate the cost of the pumparound circuit. Where more than one sidestream is withdrawn, however, it is usually economical to withdraw part of the heat in a pumparound reflux system farther down the tower. The following economic factors affect the choice:

1. Reduction in top reflux permits a reduction in the tower diameter in the section above the pump- around, thereby reducing investment. However, the reduction in reflux requires the addition of more fractionating plates to maintain the same separation efficiency.
2. Since temperatures farther down the tower are higher than at the top, heat is available at a higher level. This permits transfer of heat to the incoming feed, thus reducing fuel requirements and furnace investment.
3. Cooling water cost is reduced and total heat exchanger plus condenser surface may be reduced.

Steam used in the sidestream strippers and in the stripping section of the main column is condensed in the overhead condenser. This water settles to the bottom of the distillate drum and is drawn off through a small water pot in the bottom. In most installations, ammonia gas is injected into the overhead line to raise the pH of this water and reduce corrosion.

Critical temperatures throughout the tower are controlled by automatic instruments and products are withdrawn under various combinations of flow and level control. A pipe still is capable of running for days on end with only minor adjustment by the operators except, of course, when a change in crude type or in product distribution is required.

Removing Salt from Crude Oil

Crude oil usually contains water which carries inorganic salts in solution. As the crude is heated, the water evaporates and part of the salt is deposited on heat exchanger surfaces, thus impairing their efficiency. To prevent this type of heat exchanger fouling, an electrical *desalting unit* is usually installed as an integral part of the atmospheric pipe still when the crude to be run contains more than 20 to 40 pounds of salt per thousand barrels of oil, or about 0.01 weight per cent. In the desalter, fresh water is mixed with the crude and passed to a coalescing zone. Here a high voltage is applied which causes most of the small water droplets to combine into larger ones. These are separated from the crude in a subsequent settling zone. In some cases, chemicals are added to prevent the formation of emulsions.

Before desalters came into common use, crude pipe stills were frequently equipped with *flash drums* to minimize salt deposition on hot surfaces. In the flash drum system, the crude is heated to about 300°F. under enough pressure to suppress vaporization. The pressure is released as the crude enters the flash drum and all of the water (along with a small amount of crude) is flashed off, leaving the salt as a suspension in the oil. The flashed vapor is recombined with the crude near the furnace outlet or in the flash zone of the fractionating tower.

While a desalter costs more to install than the flash drum system, it has the advantage of removing up to 95% of the salt from the oil permanently. Because less salt reaches the fractionating tower in a unit equipped with a desalter, a smaller quantity of corrosion products is formed due to high temperature breakdown and hydrolysis, and the salt content of the residual fuel oil is much lower.

Vacuum Pipe Stills

Two rather distinct types of vacuum distillation equipment have evolved in the oil industry. The differences lie in their functions and in details of construction, not in the principles of operation. For convenience, these two types are commonly referred to as *lube* vacuum pipe stills and *fuel* vacuum pipe stills.

The lube vacuum pipe still, shown in Figure 2, was developed in the period between the two World Wars. As the name implies, this type of unit was designed to meet the growing demand for heavy lubricating oil distillates. With this unit, it is possible to produce distillates which boil as high as 1050°-1100°F at atmospheric pressure.

The lube vacuum pipe still is similar in many ways to the atmospheric pipe still previously described. The reduced crude feed (or whole crude, in an integrated two-stage pipe still) is preheated by indirect exchange with the hot product streams, then flows through the furnace and transfer line to the flash zone. The flash zone operates at an absolute pressure of 125-150 millimeters of mercury. Liquid from the flash .zone is stripped with steam to increase distillate yield and steam is sometimes injected into the heating coil also. By lowering the boiling point of the feed, coil steam may reduce the peak coil temperature required for a given yield, thus minimizing gas formation and product degradation.

Cracking imposes an additional penalty in a vacuum unit in that it forms gas which cannot be condensed at the low pressures employed. This gas must be vented by compressing it to atmospheric pressure. This is accomplished by means of steam jet ejectors.

Ideally, it would be possible to operate a vacuum pipe still without ejectors, with the overhead vapors composed only of steam. In practice, however, leakage of air into the system and the minor cracking which occurs make it necessary to provide a means of removing non-condensibles from the system. In addition to the distillation of atmospheric residuum, the lube vacuum pipe still is also used for rerunning of off specification lube distillates.

The *fuel* vacuum pipe still, shown in Figure 3, is actually a simplified model of the lube unit, rather than a distinct breed of its own. This type of still did not come into general use until shortly after World War II. As demand for high quality motor gasoline soared in the immediate postwar period, refineries began digging deeper into the crude to augment the supply of cat cracking feed stocks. The lube vacuum pipe still was not ideally suited to this goal because it was designed to prepare several well-fractionated lube distillates. The large number of plates needed for fractionation imposes a relatively large pressure drop between the condenser and the flash zone; the resulting high flash zone pressure reduces the yield of heavy gas oil which can be obtained at a given temperature.

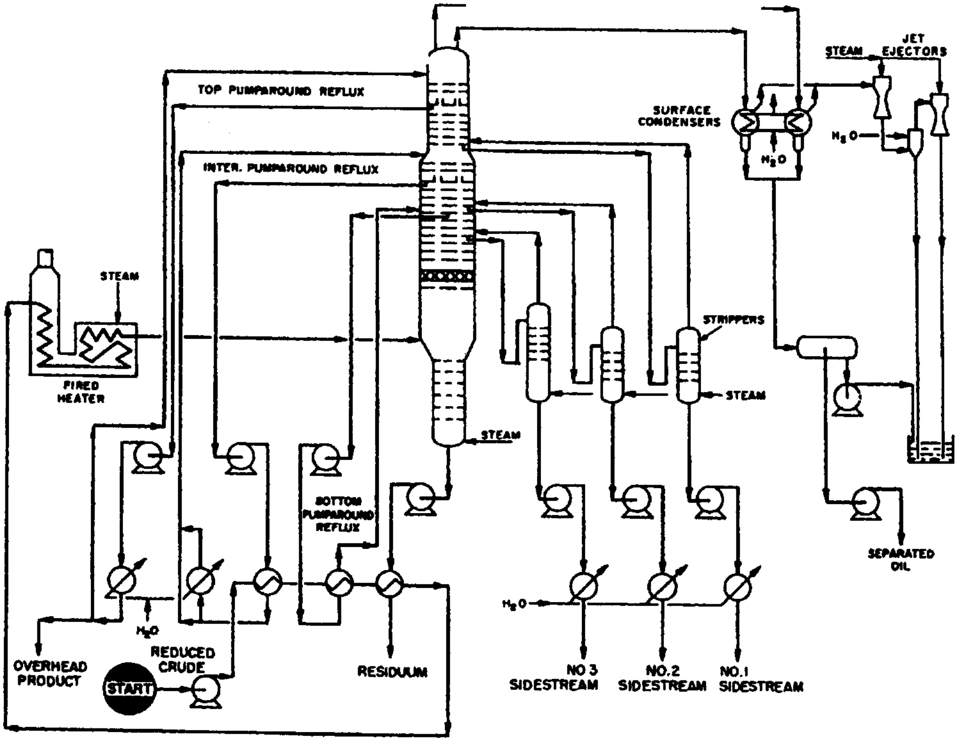


Figure 2. Lube vacuum pipe stil.

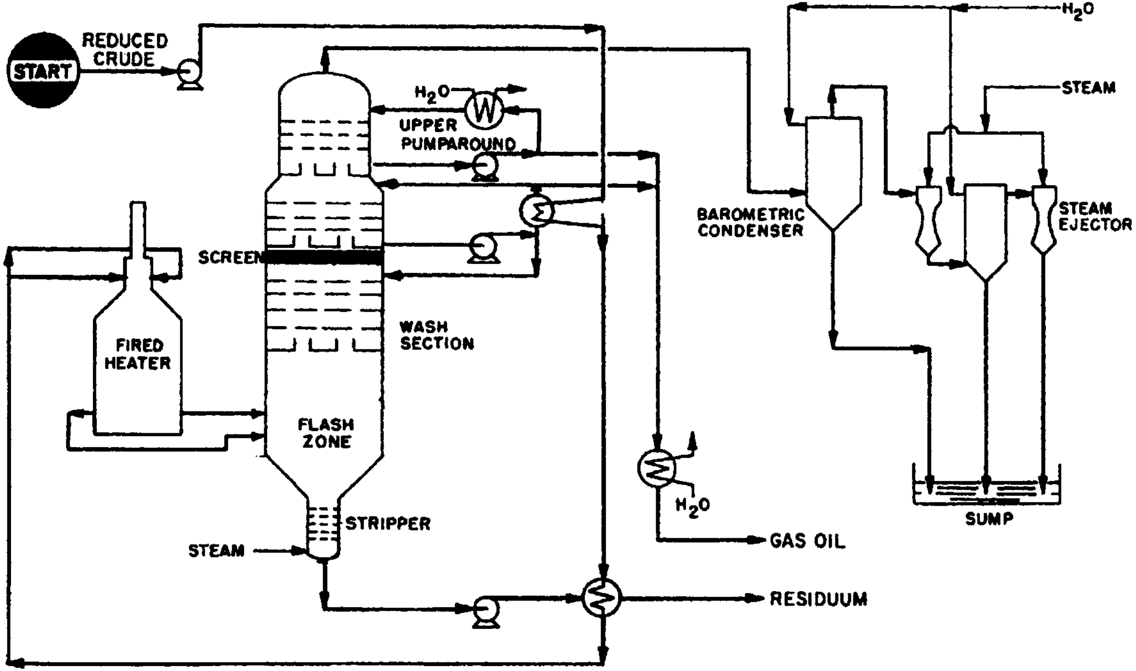


Figure 3. Fuel vacuum pipe stil.

To obtain a low flash zone pressure, the number of plates in the upper section of the vacuum pipe still is reduced to the minimum necessary to provide adequate heat transfer for condensing the distillate with the pumparound streams. A section of plates is included just above the flash zone. Here the vapors rising from the flash zone are contacted with reflux from the product drawoff plate. This part of the tower, called the wash section, serves to remove droplets of pitch entrained in the flash zone and also provides a moderate amount of fractionation. The flash zone operates at an absolute pressure of 60-90 mm Hg.

Another variation in fuel vacuum pipe still designs is to use contacting devices that give lower pressure drop than distillation trays. These permit a lower flash zone operating pressure and thereby allow higher yields of heavy gas oil. The plates in the wash section are replaced with Glitsch grids. These grids are highly efficient de-entrainment devices and give a reasonable fractionation contacting efficiency. The plates in the pumparound sections are replaced with Pall ring packing. This packing competes well with plates in providing the required heat transfer area in the minimum practical tower height.

Since cracking stocks generally do not have to meet the color specifications that lube distillates do, higher flash zone temperatures (up to 800°F) can be tolerated. Fuel units are normally designed to distill material boiling up to 1100°F (at atmospheric pressure) from the feed, and some units have distilled beyond 1200°F at low feed rates.

The fuel vacuum pipe still is also used to recover cracked gas oil from the tar formed in residuum cracking (visbreaking) processes. In this service, it it frequently referred to as a *vacuum, flash unit.* Pipe stills designed for the production of asphalt are usually the fuel type of unit.

REFINERY APPLICATIONS

In addition to the distillation of crude oil coming into the refinery, stills of various designs are used in other types of service throughout the refinery.

Recovery and Separation of Cracked Products

Cracked products are separated in distillation equipment which is very similar to an atmospheric crude pipe still. The principal difference is that these products are hot from the cracking operation, so that a fired heater is not required.

Catalytic cracking units operate at low pressure, and the primary fractionator operates at the same pressure level as an atmospheric pipe still. Two features of the product vapors affect the design of the lower section of the catalytic fractionator:

1. The vapors are superheated and must be cooled about 200°F before an appreciable amount of liquid can be condensed.
2. The vapors from a fluid catalyst unit carry a small amount of fine catalyst particles which might clog the narrow clearances of a conventional bubble cap plate.

For these reasons, the bottom section of a catalytic fractionator contains a number of rows of simple baffles. Cool liquid splashes down over the baffles, cooling the vapors, condensing the small amount of bottoms product, and washing the catalyst dust out of the vapor stream. The major part of the liquid leaving the tower bottom is pumped through heat exchangers (which may be steam boilers) and coolers, then returned to the top of the baffled desuperheating section. The flow rate of this circulating reflux stream is often twice as great as the production rate of all the product streams leaving the unit.

Normally, all of the heat is removed from the fractionator by three or more circulating reflux streams. The proportion of gas and naphtha in the cracked products is much higher than in crude oil, so it is seldom possible to reduce the diameter of the tower top as in atmospheric pipe still design. Due to the low operating pressure, it is necessary to provide expensive compression capacity to permit recovery of these light hydrocarbons in subsequent equipment.

In some situations it is economical to use a single distillation tower to carry out the functions of the cat cracker main fractionator and the fuel vacuum pipe still. This type of tower is called a combination unit. As previously described, a fuel vacuum pipe still needs large quantities of heat to distill heavy gas oil for cat cracking feed stock. On the other hand, the main fractionator of a catalytic cracking unit has a large desuperheating section for removing excess heat from the cracked products. The combination unit has these two functions in a single tower that uses atmospheric pipe still residuum to desuperheat the cracked products and distill out the valuable gas oil from it. Some operating flexibility is sacrificed and the amount of gas oil distilled is not as great as could be obtained in a vacuum pipe still. This must be balanced against the substantial cost savings involved.

Light Ends Processing

In crude distillation, there are thousands of different compounds present having a virtually continuous spectrum of boiling points. It would be impractical to consider each of these compounds in describing the crude or designing the equipment to process it. Instead the crude is treated as if it were composed of a manageable number *(~50)* of psuedo components. These are defined by dividing the crude distillation curve into a series of adjacent boiling cuts.

In light ends distillation, by contrast, the number of compounds present is usually less than 30 and may be as low as 3 or 4. Thus, the feed to a light ends distillation tower can be described in terms of the percentages of the various individual compounds present.

Contiguous cuts from an atmospheric pipe still have boiling ranges which overlap and there are a large number of compounds which are split between two fractions. Usually the value of the residuum produced is not affected by the quantity of lower-boiling materials it contains, as long as gases and other very light materials have been stripped out.

In light ends fractionation it is usually just as important to remove light material from the heavier cut as it is to keep heavy material out of the lighter cut; sidestreams are seldom withdrawn. The desired purity (expressed as per cent of impurity) of the overhead and bottoms is determined by product specifications or by the requirements of subsequent processing units. To meet these purity requirements, higher reflux ratios and greater numbers of plates between cuts are required than in crude distillation units.

Steam stripping is not adequate for the bottoms purity required. More positive stripping is obtained by charging the tower bottom liquid to a heating unit known as a reboiler. In a typical reboiler, 50% of the feed is vaporized and returned to the tower below the bottom plate. A fractionating tower equipped with a steam heated reboiler is shown in Figure 4. The reboiler may also be heated by a hot oil stream, such as a pumparound reflux stream from the primary fractionator of a cracking unit, or by a fired furnace.

Feed to the unit may be all liquid, all vapor, or a mixture of liquid and vapor. Use of a reboiler causes the heavier liquid product to leave the system at a higher temperature than if steam stripping alone were used. For this reason, the use of reboilers is restricted to those systems in which the temperature is well below the level at which cracking takes place. In practice, materials boiling above about 400°F are seldom distilled in a reboiler type unit.

The overhead products from light ends fractionators are frequently gases, or a mixture of gas and liquid, at normal atmospheric pressure and 100°F. (an average temperature for products condensed by cooling water). To avoid the

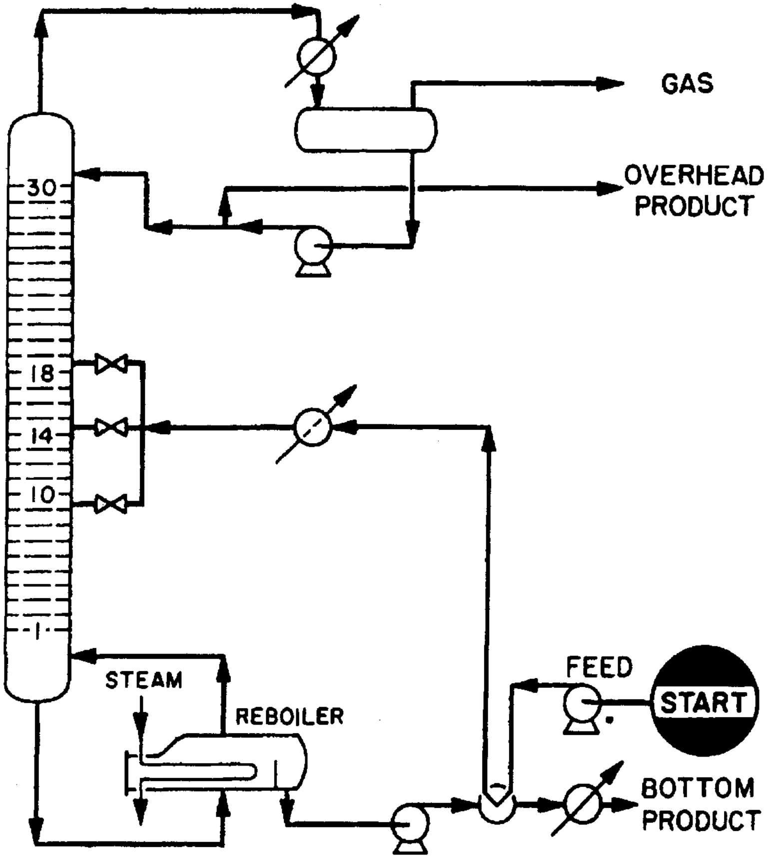


Figure 4. Reboiler type unit.

expense of providing refrigeration to condense product and reflux, operating pressure is maintained at a level that permits the overhead product to be condensed with cooling water. Pressures of 100 to 250 psig are common in refineries; pressures of several hundred pounds are used in oil field plants which recover light naphtha from natural gas. A principal disadvantage of high pressure operation is that the difficulty of separating the products increases with pressure.

As mentioned earlier the ease or difficulty of separating two products depends on the difference in their vapor pressures or volatilities. There are situations in the refining industry in which it is desirable to recover a single valuable compound in high purity from a mixture with other hydrocarbons which have boiling points so close to the more valuable product that separation by conventional distillation is a practical impossibility. Two techniques which may be applied to these situations are *azeotropic* distillation and *extractive* distillation. Both methods depend upon the addition to the system of a third component which increases the relative volatility of the constituents to be separated.

In azeotropic distillation, the added component is a chemical that forms an azeotrope (constant-boiling mixture) which boils at a lower temperature than any of the constituents of the feed. For example, addition of certain ketones to narrow cut mixtures of paraffins, naphthenes, and an aromatic is known to form azeotropes with the paraffins and naphthenes which boil at temperatures low enough to permit relatively easy separation of them from the aromatic.

In extractive distillation, the compounds which it is desired to separate are distilled in the presence of a solvent. This solvent, by reason of its greater affinity for one of the compounds, causes the other material present to exhibit an abnormally high vapor pressure which permits its removal from the system. For example, adding phenol to a mixture of toluene and close-boiling paraffins and naphthenes causes the toluene to behave as if it boiled as much as 40°F above its normal boiling point.

Solvent Removal

Petroleum products may be treated with various solvents for the removal by selective solubility of undesirable constituents or for the recovery of by-products. The solvent and solute must be separated to yield the desired product and to recover the solvent for reuse. The solvents normally boil at a lower temperature than the products from which they are to be removed and so are generally distilled off as overhead products. The pipe stills used for this service may be single-stage or multi-stage units, depending on the service involved. Some solvents can be removed by the use of steam heated stills. In other cases, the high temperature required necessitates the use of fired heaters and vacuum towers.

Rerunning

Rerunning is a second distillation step applied to distillate stocks in order to remove undesirable higher boiling materials from the product. These materials may be present because of poor fractionation in the primary distillation step; more frequently they are heavy polymers formed in treating operations. Rerunning may be combined with solvent removal, as in the case of heavy lube distillates which are diluted with naphtha to lower their viscosity before being chemically treated.

Rerunning operations are characterized by large volumes of distillate products and relatively small residue yields. Frequently, the product is withdrawn as a sidestream with undesirable light fractions passing overhead and polymers being withdrawn from the bottom of the tower. Lube rerun stills usually have several sidestreams which permit close control of flash point and viscosity while producing a wide range of stocks.

DESIGN OF EQUIPMENT

Operating Conditions

The approaches used to define operating conditions for an atmospheric pipe still and for a light ends distillation tower are quite similar. Although the preliminary calculations often involve the use of empirical correlations, which are different for crude and light ends distillation, a final design is arrived at by using a plate to plate type calculation in both cases. In the design of a vacuum pipe still, however, it is often necessary to use empirical crude oil distillation correlations as the primary design tool. This comes about because of the difficulty in characterizing heavy residuum material. These empirical correlations will not be discussed here.

The plate to plate type calculation is a fundamental procedure wherein the tower is assumed to be composed of theoretical equilibrium plates. The actual plates required are determined from the number of theoretical plates using a predicted overall tower efficiency. The starting point for a tower calculation is usually a specified feed composition, feed temperature, and tower operating pressure. The procedure involves defining the compositions and temperatures on each plate in the tower and subsequently the resultant compositions and temperatures of the product streams. The actual computations, which involve trial and error and are quite complex, are carried out on a computer. An individual balance on a tower defines the resultant products for a fixed number of plates and reflux rate in each tower section. By internal computer iterations, plates or reflux can be then varied to meet the controlling product specifications.

As was discussed earlier, in distillation calculations crude oil is represented by a series of adjacent boiling cuts developed from the crude assay distillation while a light ends stream can be characterized in terms of the actual components present. In the case of crude oil, each of the boiling range components are defined by a mid-boiling point and a specific gravity. These pseudo components are treated in the same fashion as actual light ends components in the distillation calculation. The thermodynamic properties required are predicted from mid-boiling point and specific gravity using computerized correlations based on pure component data.

Usually, product specifications for a crude distillation unit are expressed in terms of the products 15/5 or ASTM distillation curves. The prediction of a product 15/5 distillation is accomplished simply by blending the quantities of the pseudo components in the stream so as to form a true boiling point, 15/5 equivalent, distillation curve. This curve can then be converted to an ASTM type distillation using an empirical method. Figure 5 illustrates how a typical ASTM curve compares to the 15/5 curve for the same material.

Selection and Sizing of Equipment

Contacting Devices

A common type of distillation contacting device used in refinery applications is the sieve tray. In the early 50's and for many years before, the bubble cap tray was the mainstay of the distillation field. A sieve tray consists of a flat plate with regularly spaced holes, normally 1/2 to 1 inch in diameter. Liquid flows horizontally across the tray and into a channel, called a downcomer, which leads to the tray below. The sieve tray exhibits good capacity, excellent efficiency, low pressure drop, and good flexibility; i.e., it will operate quite efficiently at tower loadings which are 1/2 to 1/3 of design values.

In some situations special types of trays will show an advantage over the sieve type. For example, in sections of a tower where liquid loadings are quite high, such as in a pipe still pumparound section, a jet type tray may be employed. This is similar to a sieve tray except that the holes are replaced with openings that are partially covered with tabs. These tabs are formed by raising the metal cut out from the openings at a 20 degree angle to the plate. The tabs create a jetlike action that moves the liquid rapidly across the tray.

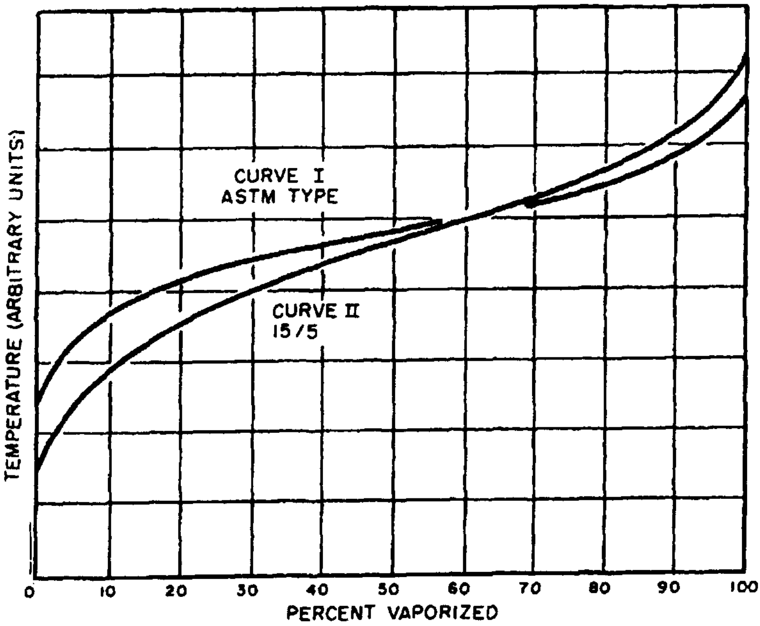


Figure 5. Distilation curves.

Tower packings have limited use as contacting devices in the petroleum industry. The older packing types, such as Berl Saddles, Rasching Rings, Pall Rings have largely been eliminated, but some designs employ Intalox Saddles. In general, these devices do flare high capacity, good efficiency and quite good pressure drop characteristics. In cost, however, they do not compete with trays for most installations. They are used in special situations; e.g., where longer residence time and increased interfacial area are required as in a reaction type separation, and where low pressure drop is particularly important as in a fuel vacuum pipe still.

Tower Sizing and Auxiliary Equipment Considerations

The diameter of a tower is established by the volume of vapors which must be handled and by the maximum allowable vapor velocity which can be tolerated without encountering excessive entrainment of liquid from one plate to the plate above. Entrainment can occur by splashing and/or suspension of small droplets in the vapor as a mist. It tends to defeat the purpose of fractionation; even a small amount may be serious when rigid specifications on color or impurities must be met.

The maximum allowable vapor velocity is predicted from correlations which are largely empirical. The governing factors are liquid and vapor density, liquid surface tension and viscosity, liquid flow rate and certain tray hardware factors such as tray spacing and percent hole area. The tray spacing is chosen by striking an economic balance between spacing and tower diameter. Minimum tray spacing is set by the tray support members, which increase in height with increasing tower diameter.

Another important consideration in tower design is tray downcomers size. At high ratios of liquid flow to vapor flow a proportionally greater area on the tray must be allotted to the downcomer channel opening. Downcomers are designed from basic hydraulic calculations. If the downcomer is inadequately sized and becomes filled with liquid, liquid level will build on the tray above. This unstable situation will propagate its way up to the tower and result in a flooded tower condition. Excessive entrainment can also lead to this same condition and, in fact, is usually the cause of flooding.

The feed to a distillation tower is normally heated either by indirect heat exchange with hot products and/or in a furnace. The products must be condensed and cooled. This is accomplished in part by heat exchange with other petroleum streams and in part by cooling water exchange. The arrangement and relative duties for the equipment are determined by an economic study which considers the investment cost of the furnace, heat exchangers, coolers, condensers, and fractionating trays and the operating cost of incremental fuel and cooling water.

Pumps, instruments and alloys for corrosion protection are also vital parts of a distillation unit and must be carefully selected for maximum efficiency.

7

LIGHT ENDS PROCESSING OPERATIONS

INTRODUCTION

Hydrocarbon fractions in the butane and lighter boiling range are usually referred to as *light ends*. How these materials are recovered, separated and further processed in a refinery are briefly discussed. However, light ends processing also involves the handling of materials in the gasoline boiling range. These heavier fractions are usually recovered together with the light ends from a primary process effluent, such as in catalytic cracking or Powerforming.

Many years ago, little effort was made to recover light ends for anything but refinery fuel with the exception of butane. However, this situation has changed drastically, and the recovery, separation, and further processing of light ends have become major refining operations. For example, by suitable processing - polymerization and alkylation - normally gaseous hydrocarbons can be converted into high octane gasoline components; the recovery of propane and butane for sale as Liquefied Petroleum Gas has become a universal refining operation; and the light olefins, such as ethylene, propylene, and the butylenes, have become the foundations of the petrochemical industry. These processes and outlets not only have made the recovery of light ends from refinery gases essential in order to realize the full economic return for such processes, but also have made it desirable in many instances to operate a process, such as steam cracking, specifically for the production of these components.

As the name implies, light ends are the "light" or volatile components encountered in refinery practice. They are normally gaseous, but may be liquefied under suitable pressure-temperature conditions. In general, light ends equipment is operated under pressure, frequently up to several hundred pounds, which permits handling these fractions as liquids, or as highly compressed gases.

Few light ends, with the exception of some propane and butanes, come into the refinery with the crude. For the most part, they are thermal decomposition by-products of high temperature refinery processes which have as their major aim the production of high- value gasoline and middle distillate fractions. The only exception to this generalization is steam cracking, which is a high temperature process specifically for the production of light ends of particular value in petrochemicals manufacture.

LIGHT ENDS RECOVERY

This section covers the various steps used to recover and separate into useful products the desirable light ends fractions present in the large volumes of effluent gases produced by the various refinery processes. The recovery is usually accomplished by a combination of compression and absorption. The subsequent separation into useful fractions is invariably carried out by distillation, usually in combination with distillates in the gasoline boiling range which are recovered with the light ends fractions.

For illustrative purposes, a typical light ends flow plan for a refinery installation is shown in Figure 1. As is the usual case in most light ends recovery plants, the major source of feed is the distillate drum of the fractionator in the catalytic cracking unit or other primary processing plant. The light ends unit consists of a compression section, an absorption section and a fractionation section. The purpose of the compressor is to bring the gases to a sufficiently high pressure to allow economical operation of the absorber in which the desirable fractions are absorbed from the gas phase into the liquid phase. In subsequent towers in the fractionation section, the recovered materials are separated by distillation into the desired cuts, such as C_2 and lighter as feed to an ethylene plant, propane for LPG, a C_4-cut for alkylation, and C_5 and heavier for motor gasoline blending.

Since this pattern of light ends processing repeats itself over and over again with certain variations from one unit to another, a discussion of the major processing steps is in order here.

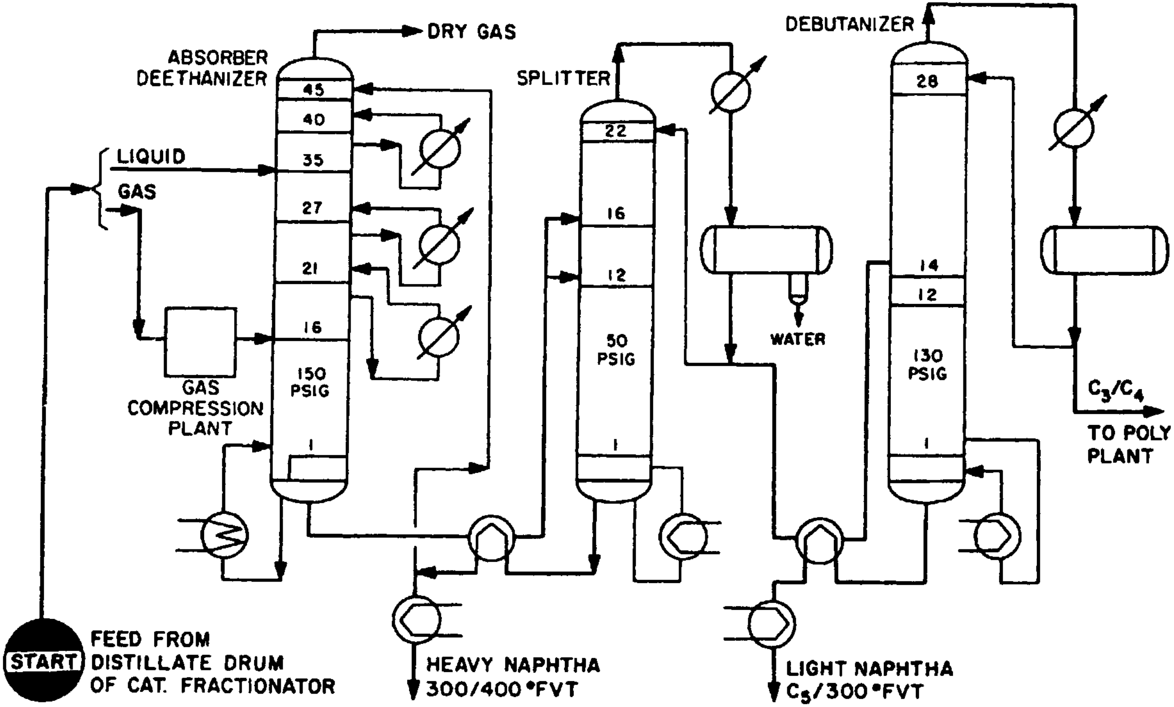


Figure 1. Typical light ends flow-plants (for catalityc cracking unit).

COMPRESSION

Most primary fuel products processes, with the exception of hydrocracking and Powerforming, operate at or near atmospheric pressure. On the other hand the absorber tower in a light ends plant runs most efficiently at elevated pressure. As a matter of fact, the higher the pressure the better it operates. Consequently, a compression plant is usually installed to bring the gases from the primary unit distillate drum to the pressure of the absorber. Since gas compressors are expensive to install and operate, an economic balance usually dictates the optimum pressure to which the gases should be compressed. Both reciprocating and centrifugal machines are used in compressor service; centrifugal compressors generally are employed in the higher horsepower installations. Figure 2 shows roughly how the selection between different types of machines can be made, but this choice will depend somewhat on the particular refinery condition.

The choice of drivers also depends entirely on the particular refinery utilities situation. Drivers range from electric motors, gas engines, and steam turbines in general use, to gas turbines in special high horsepower service (5-10,000 H.P. per machine).

GAS ABSORPTION

Basically, a gas absorption tower is a unit in which the desirable light ends components are recovered from the gas feed by dissolving them in a liquid passing through the tower countercurrently to the gas. The liquid absorbent is called *lean, oil,* and it usually consists of a hydrocarbon fraction in the gasoline boiling range. After the absorption step, the liquid which now contains the desired constituents in solution is referred to as *fat oil.* A similarly descriptive nomenclature is applied to the gas, which is referred to as *wet gas* when it enters the tower and as *dry gas* when it leaves the absorber.

The fat oil is fed to a splitter or stripping tower, where the absorbed light constituents are separated from the oil by distillation. Usually the lean oil is the same material as the heavier part of the absorber feed, so that the bottoms from the stripper are split into lean oil, which is recycled to the absorber, and a stabilized gasoline product, which is passed on to subsequent processing operations.

The absorber tower usually performs a double function. It not only recovers the desirable constituents from the gas, but also rejects undesirable lighter

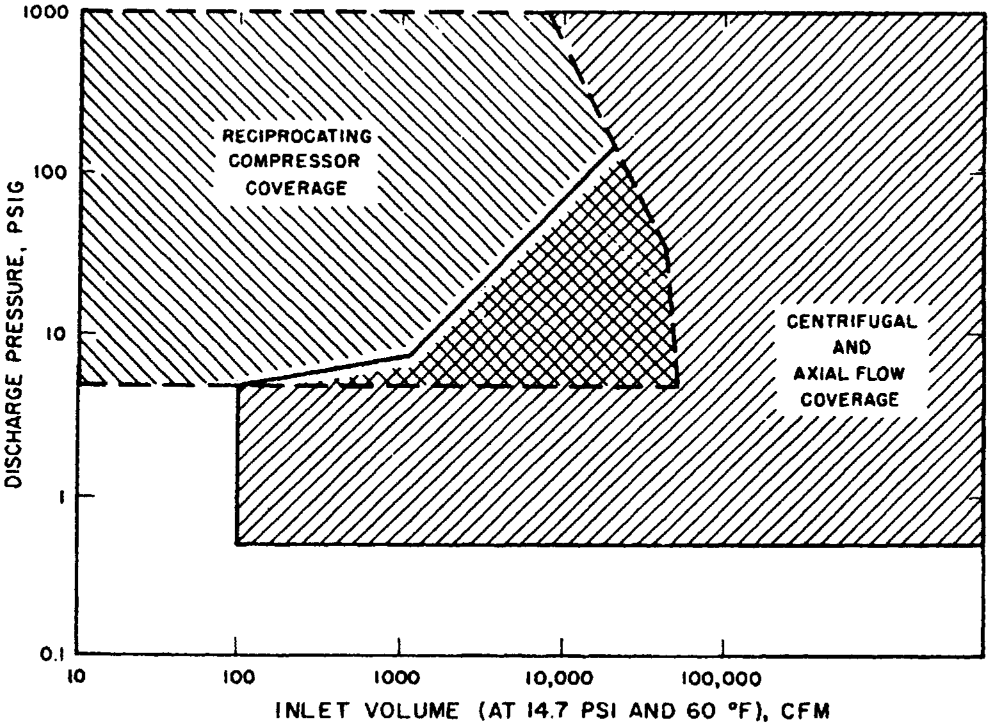


Figure 2. Compressor fields of application.

constituents and prevents them from passing on to the subsequent processing steps. This combined operation is referred to as either absorption-demethanization or absorption-deethanization, depending on which light constituent is being stripped out of the fat oil. The stripping is done by actual reboiling of the fat oil in the bottom of the column and the installation of stripping trays in the tower below the feed point. The operation of an absorber-de(m)ethanizer is analogous to a distillation column with the exception that in the absorber reflux is supplied by introducing extraneous lean oil rather than using overhead product, as in a distillation tower.

LIGHT ENDS FRACTIONATION

Fractionation is a method of separating a mixture into cuts or components according to differences in boiling point. In the distillation of high boiling materials or crude cuts we are usually dealing with a mixture of many compounds; light ends fractionation generally involves mixtures of but a few, well defined compounds or cuts. This has considerable influence on the way light ends towers are designed and operated.

A typical light ends tower is illustrated in Figure 3. In general, light ends fractionation towers are designed to separate between two components, which are referred to as the *"keys"*. The *light key* is the lower boiling, or more volatile component, and the *heavy key* is the higher boiling, or less volatile one. Essentially all the components in the feed lighter than the light key, a specified amount of light key and a small amount of heavy key are withdrawn as "overhead" product from the tower. Consequently, the "bottoms" product consists of a small amount of light key, the bulk of the heavy key, and essentially all the heavier components. This separation is achieved by "refluxing" part of the overhead product back down the tower over a series of plates, and by "reboiling" at the bottom of the tower, i.e., sending part of the bottoms product back up the tower as a vapor. Light ends towers can be designed quite rigorously for a specified split between the keys (i.e., how much of each key will go into the overhead and bottoms) by setting the reflux ratio and the number of plates in the tower. (Reflux ratio is the amount of reflux to the quantity of overhead product.)

In general, light ends towers are named after the light key. Thus, a depropanizer will take overhead the propane and lighter in the feed, and will have butane and heavier as the bottoms product.

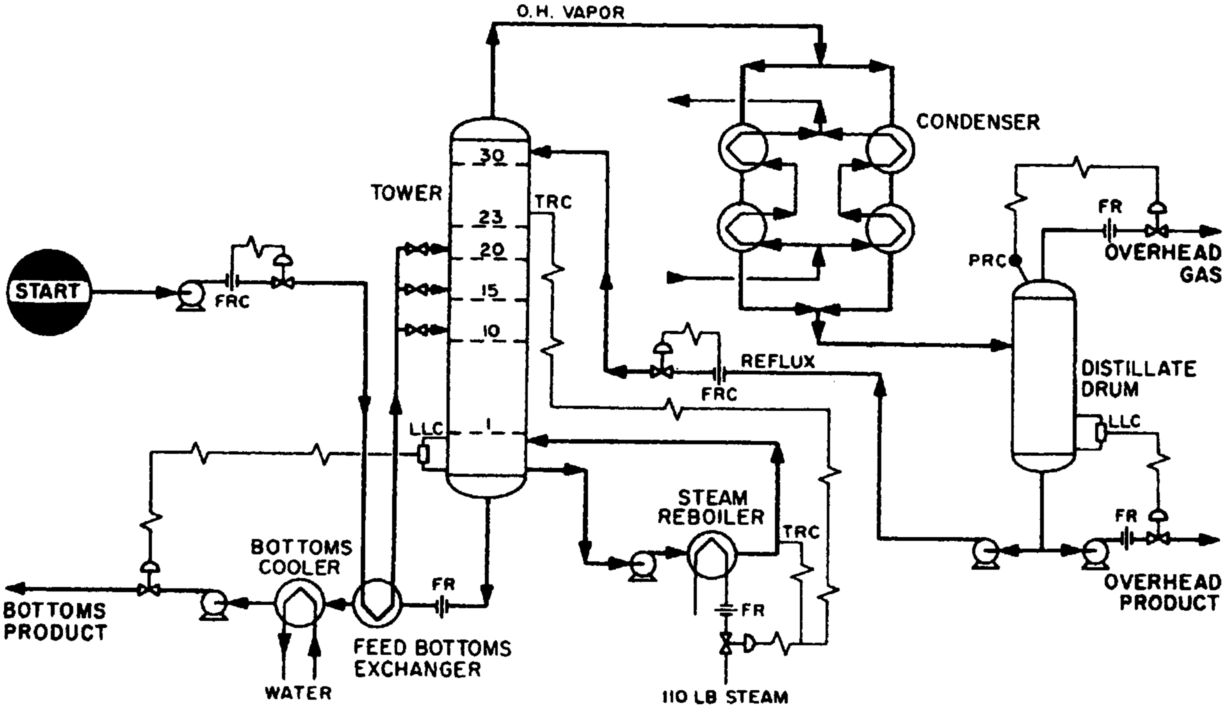


Figure 3. Typical light ends tower.

LIGHT ENDS PURIFICATION

Water and sulfur compounds are the principal non-hydrocarbon impurities present in light ends which frequently require removal. The sulfur compounds of concern are concerned with here are hydrogen sulfide and mercaptans, both of which have to be removed almost quantitatively from any light ends cut which is going to be marketed. There are two reasons for this: First, they have an objectionable odor, even in minute concentrations. Second, they may cause corrosion either by themselves or through their combustion products.

The removal of water from certain light ends streams may be necessitated by the subsequent processing step, such as isomerization or low temperature fractionation, or it may be required to meet product specifications, such as for LPG.

GENERAL TERMINOLOGY FOR LIGHT ENDS

Removal of Sulfur Compounds / ***Hydrogen Sulfide (H_2S)***

The simplest way of removing hydrogen sulfide is to scrub the hydrocarbon mixture with an aqueous solution of a strong base, such as sodium hydroxide. The hydroxide reacts with the H_2S, which is a weak acid, and the hydrogen sulfide removal is quantitative even in a single contact stage. Spent caustic from this operation cannot be regenerated economically.

However, caustic washing is never used except to remove very small amounts of H_2S, or for cleanup only, since the chemical consumption is too high. Consequently, a regenerative operation is usually employed. In this method, an aqueous solution of a weak, non-volatile base is used to extract the H2S by reacting with it at ambient temperatures to form the sulfide. The solution containing the sulfide is then regenerated by heating, which decomposes the sulfide into the original base and H_2S. Amines are generally used for this purpose, particularly diethanolamine (DEA) and monoethanolamine (MEA). The process is equally applicable to liquid and gas scrubbing.

Other regenerative methods are occasionally used to remove H_2S from hydrocarbons, such as the tri- potassium phosphate (TPP) process. Other installations are DEA or MEA, and most TPP units have been converted to DEA since the latter consumes less steam for regeneration.

Mercaptans

Mercaptans are slightly acidic organic sulfur compounds. They can be removed from light ends mixtures by caustic washing in a Merox extraction system. Hydrofining will convert mercaptans to H_2S which can then be removed by amine scrubbing. This process is used primarily for the treating of naphtha and heavier streams. However, it is quite common to treat a total atmospheric pipe still overhead stream, containing material from C_1 to a 400°F final boiling point, in a single hydrofiner process. This type of unit is referred to as a *wide cut* hydrofiner.

When both H_2S and mercaptans are to be removed in the light ends treating plant, the amine scrubbing tower precedes the caustic washing step. If the order were reversed, the large quantity of H_2S present would "fix" the caustic. Also, the caustic will act to remove any final traces of H_2S when it follows the scrubbing tower.

Water Removal (Drying)

Two common methods for drying light ends mixtures are nonregenerative reagent and regenerative adsorption systems. When very low water contents are required, liquids are sometimes dried by either fractionation or countercurrent stripping with a dry inert gas. For gas drying, since the equilibrium water content of a gas stream will decrease with increasing pressure and decreasing temperature, it is desirable to operate at as high a pressure and as low a temperature as economically possible. For example, a typical drying plant for gas streams which are to undergo low-temperature fractionation or cryogenic processing would involve water or air cooling to 95°-105°F, refrigeration to 60°-70°F, separation of the liquid water in a knockout drum and drying with a regenerative adsorption system to remove the last traces of water.

Calcium chloride is the most common nonregenerative reagent used to dry low molecular weight refinery streams to moderately low dew points. Anhydrous potassium or sodium hydroxide have also been used at times to dry liquefied petroleum gas. Sodium chloride is used most commonly to remove entrained and some soluble water from middle distillate streams.

Calcium chloride is a hygroscopic salt which can exist as pure anhydrous $CaCl_2$ or as a series of hydrate forms ranging from a monohydrate ($CaCl_2 - H_2O$) to a hexahydrate ($CaCl_2 - 6\,H_2O$). In the drying process, hydrocarbon passes up through a bed of calcium chloride and the water from the hydrocarbon forms hydrates with the calcium chloride. At the bottom of the bed the hydrate eventually reaches a hexahydrate which turns into a brine and is drawn off.

Periodically, the drier is recharged with fresh calcium chloride.

In refinery operations regenerative adsorption drying systems are used where low dew points or essentially complete drying is desired. Silica gel, activated alumina or molecular sieves are used in dual vessel regenerative drying units. In these operations the desiccant in one vessel is adsorbing water from the liquid or vapor stream at ambient temperatures while the desiccant in the other vessel is being regenerated at high temperatures, around 500°F to desorb water from the desiccant. The regenerated desiccant is cooled to ambient temperatures before switching back to adsorption service. While regenerative driers are sometimes used on LPG streams, they are exclusively used to dry Powerformer recycle gas and streams going to low temperature fractionation or cryogenic processing.

Water Washing

Washing light hydrocarbons with water is a common refinery practice. It finds application on the feed to catalytic polymerization plants. It is used to remove any entrained caustic from the mercaptan removal facilities as well as any other impurities such as amines which tend to poison the polymerization catalyst. Another use for water wash is in alkylation plants to remove salts from streams, where heating would tend to deposit them out and plug up heat exchanger surfaces. Water washing can be carried out in a mixer- settler, or in a tower if more intimate contacting is necessary.

RECOVERY PROCESSING SCHEMES

The arrangement of light ends separation facilities is an important factor in overall refinery economics. The development of the optimum scheme for a particular application often involves postulation of a number of alternatives and comparison of the economics for each.

The feed composition and degree of separation required play a major role in setting the arrangement of the light ends separation steps and the complexity of these steps. Figure 1. presented earlier, illustrates a typical light ends flow plan for a catalytic cracking unit. The first tower is an absorber-deethanizer. The large percentage of C_2 and lighter in the light ends feed stream makes it economical to remove this material as the first step. This allows the downstream towers to run at lower operating pressures. It also permits using the splitter tower to generate the lean oil needed for the absorber-deethanizer. The large quantity of light material in the feed and the high degree of separation required between C_2 and C_3 necessitate the use of an absorber-deethanizer instead of a simple deethanizer. The absorber section of this tower serves to recover C_3 and heavier from the leaving gas. It would not be required if the leaving gas was small in quantity and/or a high degree of C_3 recovery was not justified.

Treating units can be important in defining a light ends separation scheme in that they may alter the light ends stream composition. This happens when we use a hydrofining process for sulfur removal as it requires combining the feed with a hydogen rich gas stream. The light ends separation steps following a hydrofiner must then also handle the hydrogen and associated light ends from this treat gas in addition to the original feed. On the other hand, chemical treating does not significantly alter the feed composition and therefore does not affect the required light ends separation steps. In some cases the choice between hydrofining and chemical treating for C_6 and lighter streams is dictated in part by the differences in the subsequent light ends separation schemes required.

Often it is possible to integrate the separation facilities for different light ends feed streams and thereby achieve significant cost savings. Figure 4 illustrates a scheme where a single absorber-deethanizer is used for separating the light ends from a wide cut hydrofiner unit, a Powerformer and a catalytic cracker. The particular scheme shown is used in cases where a relatively low C_3 recovery is required.

The H_2/400°FVT streams from each system are sent to separate flash drums where the bulk of the C_2 and lighter material is removed. The virgin and cat cracker streams from the flash drums go to separate debutanizers while the Powerformer stream goes to an absorber-deethanizer followed by a debutanizer. The C_4 and lighter overhead streams from the virgin and cat cracker debutanizers are sent to this absorber- deethanizer for final deethanization. In the flow scheme shown this tower does not have a separate lean oil. It is called an absorber-deethanizer because the Powerformer stream serves in part to absorb the C_3 and C_4 components in the streams from the debutanizers. A separate lean oil stream is added in cases where higher C_3 and C_4 recoveries are justified.

LIGHT ENDS MANUFACTURING PLANTS

The following is a description of plants leading to specific light ends cuts. This includes producing LPG propane, and also high purity ethylene.

Propane (LPG) Plant

The sale of C_3 as liquefied petroleum gas (LPG) is a profitable outlet for this material. However, LPG has to meet rigid specifications on low moisture content, low sulfur, and usually a minimum of 95% C_3 content, with the

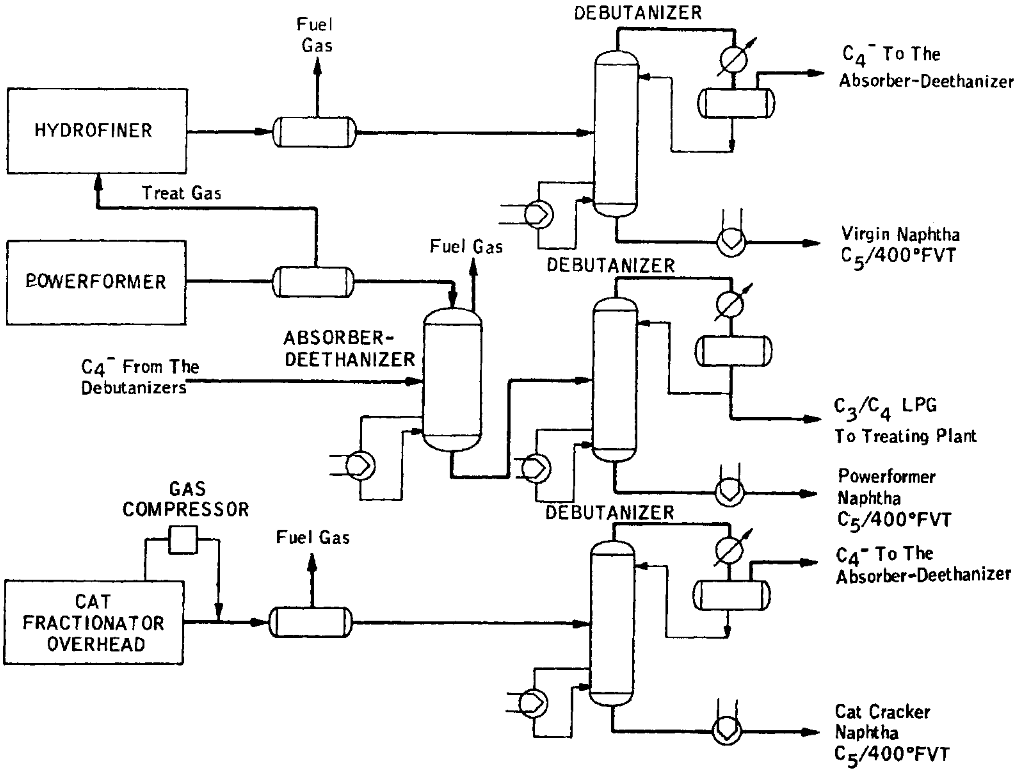


Figure 4. Flow plan for an integrated light ends system.

remaining 5% split half-half between C_3 and C_4.

Figure 5 illustrates a propane LPG recovery unit designed to process the C_3-liquid overhead of a poly plant depropanizer. No sulfur removal facilities such as a caustic scrubber are required here since the poly plant feed has had a thorough sulfur clean-up. The liquid feed is first passed through a coagulator, which is merely a straw-filled drum to coagulate entrained water, which is settled out and withdrawn. The C_2-C_3 liquid now free of entrained water is then fed to the deethanizer which not only deethanizes, but also dehydrates the bottoms due to a peculiar volatility relationship of water in light hydrocarbons. The excess ethane and water are withdrawn from the overhead drum, and the bottoms constitute the dried LPG product.

High Purity Ethylene

Ethylene is one of the principal raw materials in the petrochemical field. There are various schemes used to produce both dilute and concentrated ethylene.

I. Low Temperature - High Pressure Fractionation

In this process the gas containing ethylene is first treated to remove acetylene by selective catalytic hydrogenation. Next, the feed is dried and cooled by refrigeration to a low enough temperature level to condense out the bulk of the ethylene and heavier components in the feed. The mixture of gas and liquid is then fed to a demethanizer where the remaining ethylene is recovered from the gas by going to an extremely low top temperature. The demethanized liquid is withdrawn from the bottom of the tower and fed to the subsequent fractionation section where various high purity cuts are prepared. These are high purity ethylene (98% up to 99.8%), an ethane cut, and in some cases a C_3 and heavier stream.

II. Low Temperature - High Pressure Absorption

Instead of condensing out the ethylene at extremely low temperatures, as described above, the absorption process uses a very light, lean oil, such as C_3, at moderately low temperature to scrub the ethylene out of the gas. By stripping the fat oil, the ethylene and heavier cuts can be recovered as desired.

III. High Pressure Absorption

This is more of a clean-up process than a concentrating operation. Instead of preparing a high purity, high concentration ethylene cut, the absorption

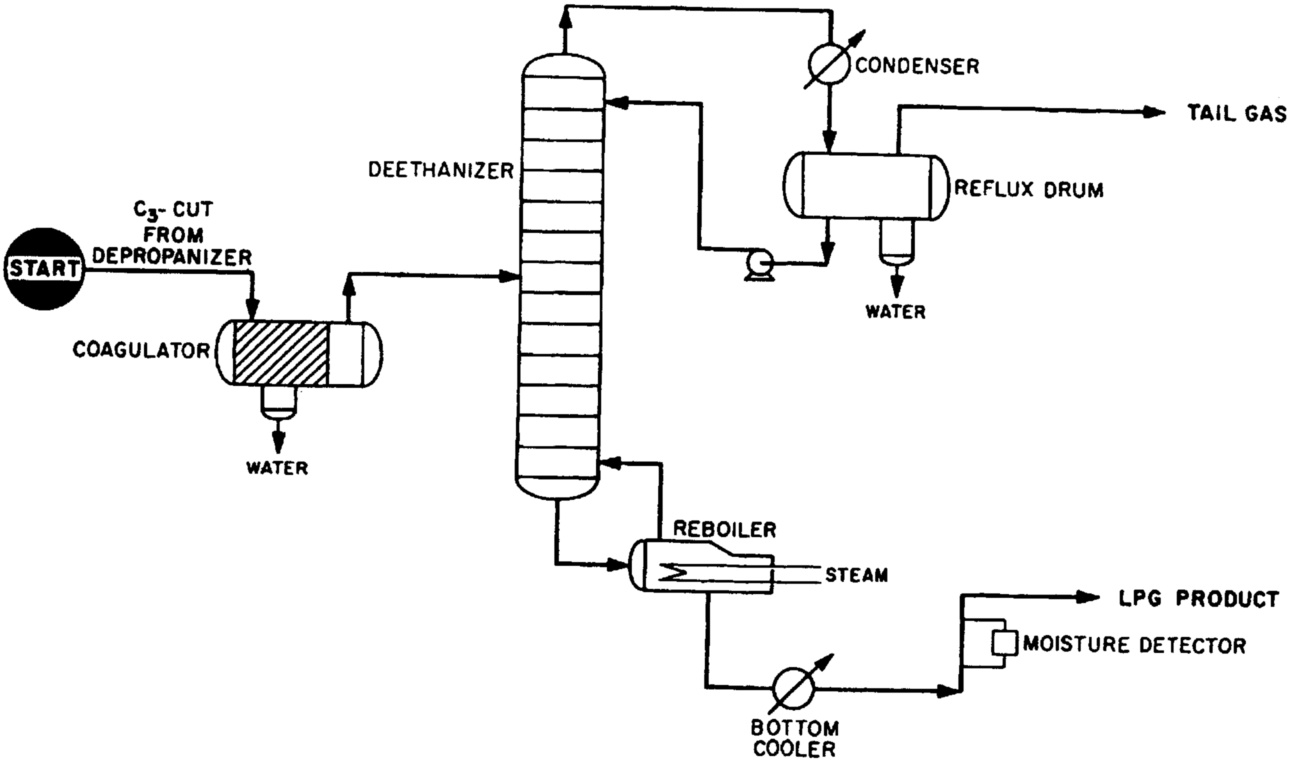


Figure 5. Poly Plant propane LPG recovery unit.

process merely removes C_3 and heavier, down to very slight traces, from the gas stream containing the ethylene. This is done in a conventional absorber-deethanizer tower using a "chilled" lean oil, i.e., cooled to about 60°F.

OLEFINS

Olefins are the basic building blocks for many chemical syntheses. These unsaturated materials enter into polymers, rubbers, and plastics, and react to form a wide variety of chemical compounds such as alcohols, amines, chlorides and oxides.

Steam Cracking is the thermal cracking and reforming of hydrocarbons in the presence of steam at high temperature, short contact time, and rather low pressure in a fired tubular furnace. From the standpoint of both the amount and variety of compounds produced, steam cracking of gas oils and naphthas is one of the most important petroleum process for producing a wide range of chemical raw materials. Ethane and propane cracking are used widely by others but relatively few products other than ethylene result.

In a typical gas oil design, the lighter products overhead from the quench tower/primary fractionator are compressed to 210 psi, and cooled to about 100°F. Some C_5 plus material is recovered from the compressor knockout drums. The gases are ethanolamine and caustic washed to remove acid gases: sulfur compounds and carbon dioxide, and then desiccant dried to remove last traces of water. This is to prevent ice and hydrate formation in the low temperature section downstream.

In the depropanizer tower the propane and lighter gases are taken overhead to become feed to the ethylene and propylene recovery facilities. Separation is accomplished at a relatively low overhead temperature of -25°F to minimize reboiler fouling by olefin polymerization.

Butane and heavier bottoms from the depropanizer flow to the debutanizer where the C_4 stream (almost entirely olefins and diolefins) is taken overhead and sent to butadiene and isobutylene recovery facilities.

Depending upon the refinery needs, the raw C_5 plus steam cracked naphtha may be sent to isoprene extraction, treated to remove gum forming diolefins and sent to the refinery gasoline pool, or else completely hydrogenated and then fed to an aromatics extraction unit.

The principal function of the ethylene recovery facilities is to recover high purity ethylene (Figure 2). Ethylene recovery consists basically of a low temperature, relatively high pressure distillation process to separate ethylene from other hydrocarbons and hydrogen. In addition, acetylene conversion and caustic treating steps are employed to reduce contaminants which would not be adequately removed by the distillation process.

The depropanizer overhead, C_3 and lighter feed is compressed to about 300 psi and then passed over a fixed bed of acetylene removal catalyst, generally palladium on alumina. Because of the very large amount of hydrogen contained in this stream, the operating conditions are critical to selectively hydrogenate the acetylene without degrading the valuable ethylene to ethane.

The gases are again dried and then further compressed to about 550 psi. Separation of hydrogen and methane take place in the demethanizer and in its preflash system. Three successive Golder preflash steps are used in this separation, with propylene as refrigerant, then ethylene, and finally a self-generated methane refrigerant at -200°F.

A high purity hydrogen and a low purity methane stream result. The 95% hydrogen may be used directly to hydrogenate steam cracked naphtha or directly consumed elsewhere in the refinery. The methane stream goes to fuel.

The C_2 plus bottoms from the demethanizer then go to the deethanizer. A propylene-propane bottoms product containing 90-92% propylene is obtained which may either be sold, used directly as propylene- 90, or further purified. The ethylene-ethane overhead from the deethanizer is separated in the splitter tower yielding a 99.8% overhead ethylene product at -25°F. The ethane bottoms at +18°F may either be sent to fuel gas or used as feed to an ethane cracking furnace. Overall ethylene recovery in these facilities is about 98%. The product is of very high purity with less than 50 parts per million of non-hydrocarbon contaminants and a methane plus ethane level below 250 ppm.

Propylene Recovery

The propylene-90 bottoms product from the deethanizer may be upgraded to high polymer grade 99.8% purity by superfractionation. Propane bottoms are used elsewhere in the refinery.

Butenes

Normal butenes and isobutylene are separated by a selective reaction-extraction process which takes advantage of differences in reactivity with dilute sulfuric acid to form butyl alcohols. Because of differences in olefin structure, isobutylene reacts much more rapidly than normal butenes with weak acid. In fact, reaction of normal butenes in acids weaker than 65% is negligible at commercial conditions. Reaction products are soluble in dilute acid. The unreacted feed is only slightly soluble.

The acid extract phase is separated, diluted with water, and heated to regenerate isobutylene. The isobutylene is then caustic and water washed to remove traces of acid, distillation dried, and rerun. The unreacted C_4 stream, containing normal butenes, is also caustic washed before further processing.

C_4 cuts, after extraction of butadiene, are preferred as feed to isobutylene extraction units because the isobutylene concentration (about 30-40%) is higher than in C_4 streams from catalytic cracking. The basic reaction in isobutylene extraction is the reversible hydration of isobutylene to tertiary butyl alcohol in the presence of sulfuric acid.

Polymerization to C_8 and C_{12} olefins is the chief side reaction. Polymerization increases with extraction temperature and with the hold-up time in the extraction section. It limits the temperature used to obtain high extraction rates.

The extraction is carried out in a staged countercurrent system for good recovery of isobutylene. Temperature is maintained by refrigeration, since heat is evolved in the hydration. Normal C_4's are rejected as the raffinate from the lean stage. The stream, typically containing 70 mol% normal butenes, can be used as feedstock for dehydrogenation to butadiene. The rich acid extract is flashed to about 2 psig and blown with a small amount of steam to remove butylenes and butanes physically dissolved in the extract. Isobutylene is then recovered from the acid extract by direct injection of steam in the regenerator tower.

Enough steam is used to reduce the acid concentration from 65% to 45%. The heat supplied by the steam is used in: (a) regeneration of isobutylene from t-butyl alcohol (an endothermic reaction) (b) raising the acid temperature to 250°F and (c) distilling out isobutylene, polymer, and residual tertiary butyl alcohol. High temperature and low acid strength allow regeneration of the isobutylene with minimal polymerization. Acid strength in the regenerator tower is critical. Too low values result in separation of unwanted alcohol while high concentrations increase polymerization rates.

The regenerator overhead is caustic and water washed, yielding a 95-96%

isobutylene product. The 45% acid taken as bottoms from the regenerator is concentrated to 57% for steam cracked C_4 cuts (65% for cat cracked C_4's) and recycled to the lean stage of the extraction section.

High purity 99+% isobutylene can be made by rerunning, with a recovery of over 85% of the isobutylene in the feed.

When high purity isobutylene is not required, the acid extract from the rich stage may be heated for a few minutes to 250-300°F, and then quickly cooled. Under these conditions the isobutylene dimerizes to form largely 2,4,4, trimethyl pentene-1. This is known as the dimer process and may be used to concentrate n-butenes for dehydrogenation feed, the isobutylene dimer being added to the motor gasoline pool. Trimers, as well as codimers with normal butenes are also produced.

Low Molecular Weight Olefins

Low molecular weight olefins ranging from C_6 to C_{15} are produced by the polymerization of propylene or butylenes over a supported phosphoric acid catalyst. The product of this polymerization is a series of highly branched olefins ranging from dimers to pentamers. Some fragmentation of the polymers formed takes place in the reactor, so appreciable quantities of olefins are obtained which are not integral multiples of the monomer units.

Figure 4 shows the flowplan for the production of propylene tetramer (C_{12} olefin). The feed is first caustic-treated, water-washed, and then hydrofined for removal of acetylene, nitrogen and sulfur compounds. Feed is then reacted over a phosphoric acid- Kieselguhr catalyst "UOP Solid Phosphoric Acid" in multiple bed adiabatic reactors at about 375°F and 750 psig. The reaction is strongly exothermic, and is controlled by interbed cooling with (inert) propane quench. Reactor effluent is depropanized and then fractionated into light polymer, tetramer product, and heavy polymer. The light polymer, a C_6 to C_{10} smear, may then be rerun to obtain crude C_7 and C_9 olefin fractions. The relative amounts of C_7, C_9, and C_{12} olefin produced may be shifted by varying the quantity of light polymer recycled to the reactors.

Most of the propylene polymerized by this process is used in motor gasoline ("Polymer Gasoline''). However, an appreciable portion of the C_7, C_9, and C_{12} olefins find use as feedstocks for production of Oxo alcohols.

DIOLEFINS

Conjugated dienes (diolefins) are used in the manufacture of synthetic rubbers, and as starting materials in the synthesis of insecticides and flame proofing paints.

Butadiene yields ranging from 2 to 7 weight percent on feed (usually 4 to 5) are expected in the steam cracking of naphthas and gas oils. This is generally 35 to 45 percent of the total yield of C_4's.

The principal components of the C_4 cut are butene-1, butene-2, isobutylene and butadiene-1,3. Methyl, ethyl, and vinyl acetylenes, butane and butadiene-1,2 are present in small quantities. Butadiene is recovered from the C_4 fraction by extraction with cuprous ammonium acetate (CAA) solution, or by extractive distillation with aqueous acetonitrile (ACN). The former process is a liquid-liquid separation, and the latter a vapor-liquid separation. Both take advantage of differences in structure and reactivity of the various C_4 components to bring about the desired separation.

Acetylenes and easily formed acetylene polymers are troublemakers in both processes, and special purification steps are required. They are also highly unwanted contaminants in the product butadiene.

The butanes and butenes have only limited physical solubility in ammoniacal cuprous acetate solutions. Compounds of higher unsaturation (dienes and acetylenes) form addition complexes, so their effective solubilities are much higher.

Acetylenes are most readily absorbed by CAA, and hence are generally removed in a prewash step. Strict control of the acetylene content of the recirculating solvent is necessary and a charcoal adsorbent bed is used to maintain the polymer content of the solution at low levels.

A series of refrigerated counterflow mixers and settlers can be used in the main extraction train. A temperature profile is maintained between stages. For high butadiene recovery, lean solvent enters at the Goldest stage where spent C_4's are rejected, and flows countercurrent to the less soluble components. Typically, seven mixer-settler stages are used. Four serve as absorption stages above the feed entry, two serve as stripping stages, and one as a heated rejection stage. In this last stage some butadiene is removed from the rich solution and returned to the stripping stages. This is analogous to reboiling bottoms product in distillation, and controls product purity in the same way. The bulk of the butadiene remaining in the rich solution is then desorbed by heating, scrubbed to recover ammonia evolved from the CAA, and rerun to meet product specifications. A high boiling flux oil is added to the rerun tower for ease in processing the low boiling product. Impurities are generally acetylenes, carbonyls, peroxides, sulfur compounds, butadiene dimer, and an isomer: butadiene-1,2. The lean CAA is returned to the main extraction train, with a slipstream to charcoal treatment.

Acetonitrile serves to greatly enlarge the spread of relative volatilities so that reasonably sized distillation equipment can be used to separate butadiene from the other components in the C_4 fraction. The polar ACN acts as a very heavy component and is separated from the product without much difficulty.The feed stream is carefully hydrogenated to reduce the acetylene level rerun, and then fed to the single stage extractive distillation unit. Feed enters near the middle of the extractive distillation tower, while (lean) aqueous ACN is added near but not at the top. Butenes and butanes go overhead as distillate, with some being refluxed to the tower and the rest water washed for removal of entrained ACN.

Tower bottoms-ACN, butadiene, with some butenes and acetylenes - are fed to a recovery/stripping column. The hydrocarbons are taken overhead and then rerun to meet product specifications. The stripping column bottoms, (ACN) is then returned near the top of the extractive distillation tower. A small slipstream goes to the ACN recovery tower, where solvent is also recovered from the water wash streams.

The spent C_4 streams from either process can be sent to isobutylene extraction, butene dehydrogenation, or used in further refinery processing.

Isoprene

The demand for isoprene for Butyl rubber led to the development of a recovery process for this C_5 diolefin. Extractive distillation with acetone was the first process used but it has been replaced with acetonitrile (ACN;). The first step in the process is the fractionation of steam cracker debutanizer bottoms in a conventional two tower system to produce a C_5 cut containing 30% isoprene. The first tower rejects C_6 and heavier while the second rejects C_4 and lighter materials.

Since most of the C_5 hydrocarbons boil extremely close to isoprene, simple distillation of the C_5 cut would not produce high purity isoprene. Two stages of extractive distillation with acetonitrile are used where the ACN modifies the relative volatilities of the hydrocarbons and thus enables the separation to be made.

The first stage does the bulk of the separation, and the second is used to remove other dienes and acetylenes from the isoprene. In the extractive distillation tower of each stage, the ACN solvent is introduced near the top, and being the highest boiling component, travels downward. The effective relative volatility of the less unsaturated hydrocarbons is increased with respect to isoprene. Thus most of the impurities go overhead and the isoprene is carried down with the solvent.

Overhead from the first stage extractive distillation tower consists of the bulk of the other C_5's: pentenes and pentanes. Tower bottoms are fed to a recovery/stripping column. The stripping column bottoms (ACN) is then returned near the top of the distillation tower. Stripper overhead containing 80% isoprene and consisting almost entirely of dienes and acetylenes is then fed to the second stage of extractive distillation. The isoprene goes overhead where it is water washed and rerun to 99.6 mol% purity. The other dienes and acetylenes go out in the extractive distillation tower bottoms to a second recovery/stripping column.

All streams leaving the extractive distillation sections are water washed to remove entrained ACN, and the ACN is recovered by distillation. Spent C_5's from the first stage distillation tower overhead may be recycled to a steam cracking unit. This material gives excellent butadiene yields.

Cyclopentadiene and Methylcyclopentadiene

The recovery of cyclopentadiene is based on the rapid dimerization rate of this compound to dicyclopentadiene, and the ease of separating heavy dimer from the feedstock.

A C_5 to C_8 cracked naphtha cut is heated under pressure in a soaking drum. Dimers and codimers of cyclopentadiene, methylcyclopentadiene, and isoprone are formed. This is followed by distillation, leaving a dimer concentrate bottoms.

This crude dimer concentrate may be sold "as is" for some uses. But generally it is upgraded by further processing. Here the dimers are partially cracked at about 400°F in a cracking drum and fed to a distillation tower. A 95+% purity cyclopentadiene monomer stream is taken off overhead. A C_6 monomer concentrate is taken off as a sidestream below the feed point. Dimers and higher polymers are recycled to the cracking drum, where inert materials and polymers are removed in a liquid purge steam. The purified monomers are redimerized before storage and sold as dimer.The cyclopentadiene unit can also produce a 90+% methylcyclopentadiene fraction in blocked operation.

ACETYLENES

C_2 and C_3 Acetylenes

Significant quantities of C_2 and C_3 acetylenes are produced in cracking. They can be converted to olefins and paraffins. For the production of high purity ethylene and propylene, the contained C_2 and C_3 acetylenes and dienes are catalytically hydrogenated leaving only parts per million of acetylenes in the products. Careful operation is required to selectively hydrogenate the small concentrations of acetylenes only, and not downgrade too much of the wanted olefin products to saturates.

One approach is to uses solvent extraction with dimethyl formamide (DMF) to remove C_2 acetylene and a C_3 acetylene-propadiene mixture from their steam cracked ethylene and propylene streams. The simple acetylene is sold as welding gas, and the C_3 stream is sold as starting material for chemical synthesis.

C_4 and C_5 Acetylenes

Troublesome amounts of C_4 and C_5 acetylenes are also produced in cracking. In the butadiene and isoprene recovery processes, the acetylenes in the feed are either hydrogenated, polymerized, or extracted and burned. Acetylene hydrogenation catalyst types include palladium on alumina, and some non-noble metals.

AROMATICS

Aromatics are petrochemicals. Fixed-bed reforming of virgin naphthas is one source of these materials. Aromatics from the high temperature of coking of coal, the main source prior to 1940, now only account but for a small proportion of the total production.

The C_6-C_8 aromatics are important large volume chemicals to the entire petroleum industry.

Powerforming is one tecnique used for aromatics chemical production. Powerforming uses a platinum catalyst to reform virgin naphthas. The principal reaction is the conversion of naphthenes in virgin naphthas to aromatics; e.g., isomerization and dehydrocyclization reactions also occur in catalytic reforming.

The virgin naphtha feed after having been hydrofined over a cobalt molybdenum catalyst to remove sulfur compounds passes through the Powerformer. After stabilization, the aromatics are recovered by extraction with either aqueous glycols or sulfolane. The extract is then clay treated and fractionated into benzene, toluene and mixed C_8 aromatics (ethylbenzene, ortho, para, and metaxylene).

Benzene

Benzene is surpassed only by ethylene as a primary building block for organic chemicals.

Over eighty percent of the benzene produced is used in manufacturing only three derivatives: styrene, phenol and cyclohexane.

Benzene can be produced by hydrodealkylation of toluene and mixtures of toluene-xylenes. The hydrodealkylation process, operates at 1175-1330°F and 600 psig. The feed is preheated to the reaction temperature in a furnace and is passed through an open reactor ($L/D > 20$) where thermal demethylation of toluene and xylenes occurs. Due to the high heat release of the dealkylation reaction (about 22,000 BTU/lb- mole of hydrogen consumed), hydrogen is injected at a number of points in the reactor in order to control the temperature. A small amount of hydrogen sulfide or carbon disulfide is added to prevent catastrophic corrosion of the furnace tubes. The hydrodealkylation process is quite selective, e.g. 95 mole% selectivity to benzene at a conversion level of 90%. A small amount of polymer is formed, mainly diphenyl, which may be recycled to the reactor to forestall further diphenyl formation, thereby resulting in even higher selectivity.

Benzene is a flammable liquid and its vapors are *toxic* and explosive. Low concentrations are dangerous on continued inhalation because benzene affects the blood forming function of the bone marrow and it is a cancirogen. Dermatitis may result from repeated skin contact. Alkyl derivatives such as toluene and xylenes are far less toxic and are, therefore, much safer than benzene for use in solvents. Some of the symptoms of benzene poisoning are: dizziness, constriction of the chest, and tightening of the leg muscles.

Toluene

Toluene is recovered as a high purity product by fractionating the mixed aromatics obtained from the extraction of catalytic reformate or powerformate. About 70 fractionation trays are required to produce toluene having a purity of 99.7 percent. Toluene is consumed principally as a feedstock for hydrodealkylation plants. Toluene is used in a number of solvent applications.

One large outlet is its use as a solvent for nitrocellulose lacquers. Toluene is used to make phenol, benzaldehyde, benzole acid and a variety of other derivatives.

Ethylbenzene

Ethylbenzene is separated from mixed xylenes by fractionation using 360 trays and a high reflux ratio. Ethylbenzene is separated from the closest isomer paraxylene whose normal boiling point is only 3.90°F higher. The average relative volatility between ethylbenzene and paraxylene in the fractionation is about 1.06. The fractionator feed is entirely C_8 aromatics which are prepared by the extraction of powerformate by the sulfolane process and by fractionation of the aromatic extract.

Ethylbenzene can also be produced by catalytic alkylation of benzene with ethylene. Benzene is alkylated with ethylene in a fixed bed alkylator. An excess of benzene is used to suppress the formation of di- and triethyl- benzenes. The excess benzene is removed from the alkylate by fractionation and recycled to the alkylator. The ethylbenzene is separated from the polyalkylated benzenes which are in turn fed to a separate reactor. Here benzene is added to convert the polyalkylated benzenes to monoethylbenzene by transalkylation.

Styrene from Ethylbenzene by Dehydrogenation

Ethylbenzene is dehydrogenated to styrene over a fixed bed of catalyst and in the presence of a large excess of steam at 1150-1200°F and 1 atmosphere.

Superheated steam is used to bring the feed to reaction temperature. Reactor effluent is quenched, distilled to remove unreacted feed plus benzene and toluene made during the reaction, and the crude styrene finished by vacuum distillation. Inhibitors are added during the distillation steps to prevent polymerization of the styrene monomer.

Paraxylene

Paraxylene is recovered from C_8 aromatics by crystallization. Fortunately, the solidification point of the para isomer is unusually high, +55.9°F, considerably above the meta and orthoxylenes which freeze at -54.2 and -13.3°F, respectively. The separation of para from meta by distillation is impractical because the spread in their normal boiling points is only 1.4°F.

Paraxylene is used to make terephthalic acid, the raw material for manufacturing polyester fibers, such as Dacron, Kodel, Fortrel and Terylene.

In order to produce more paraxylene than is available in catalytic reformate, a xylenes-isomerization plant is sometimes included in the processing scheme. The isomerization step uses the effluent (filtrate) from the paraxylene crystallization step as feed. The filtrate contains about 7-9 percent of paraxylene. The isomerization unit brings the concentration back to its equilibrium value of about 20 percent.

Orthoxylene

Orthoxylene (the highest boiling xylene isomer) is separated from the other xylenes and the heavier C_9 aromatics by fractionation. The meta and lighter xylenes are taken overhead in a xylenes splitter containing 160 trays. Orthoxylene is then separated from the C_9 aromatics in a 50-plate rerun column. Product purity from such a fractionation is typically 99+%.

High purity orthoxylene is used almost solely in the manufacture of phthalic anhydride. The anhydride is an intermediate for the manufacture of plasticizers, alkyd resins, polyester resins and other derivatives, such as phenolphthalein.

Metaxylene

Processes have been developed separating pure metaxylene from other C_8 aromatics.Metaxylene is a raw material for the manufacture of isophthalic acid. The major outlets for the acid are in the synthesis of unsaturated polyester and alkyd resins, and for the production of isophthalic esters (plasticizers).

NORMAL PARAFFINS

Normal paraffins in the C_{10} - C_{15} range are recovered from petroleum fractions by adsorption-desorption using molecular sieves. Ammonia can be used to desorb the n-paraffins. By employing two beds of sieves, one on adsorption and one on desorption at all times, a continuous flow of the feed and ammonia is maintained.

In the adsorption step, the hydrocarbon feed is vaporized in a furnace and is passed upward through the sieves which have been previously desorbed by ammonia. The n-paraffins in the feed are adsorbed by the sieve and displace the ammonia. The amount of ammonia removed depends largely on the molecular weight of the feed and the amount of ammonia in the feed. The effluent, which is depleted in n-paraffins, is condensed and separated from the ammonia. Desorption by ammonia displacement overcomes the high heat of adsorption of n-paraffina. Ammonia is absorbed on the sieve with a high heat of adsorption, about 20,000 BTU per mole.

In the desorption step, ammonia is passed downflow through the bed which has completed the adsorption cycle. The ammonia is heated to approximately the same temperature as that of the feed in the adsorption step in order to maintain a nominally isothermal operation. The first portion of the desorbate, although rich in n-paraffins, contains impurities and is recycled to the second bed which is simultaneously operating on the adsorption cycle. The remaining product is condensed and separated from ammonia. The product is freed of dissolved ammonia by distillation.

The sieves slowly lose capacity and are regenerated about by burning with dilute oxygen. The process is capable of recovering n-paraffins up to C_{20}.

CYCLOHEXANE

High purity cyclohexane is manufactured by hydrogenating benzene at 400-500°F and 500 psig. Some cyclohexane was earlier produced by fractionating naphtha but its purity of 85-90% was too low to compete with 99+ percent purity from benzene hydrogenation. A number of cyclohexane processes based on benzene hydrogenation are available.

The hydrogenation of high purity benzene is carried out in three or four adiabatic reactors over a supported platinum catalyst. External cooling between the reactors is used to control the highly exothermic reaction which liberates about 90,000 BTU per pound mole of benzene converted to cyclohexane. The effluent from the final reactor is cooled to about 100°F and sent to a separator drum. The vapor from the drum is refrigerated to remove the cyclohexane before the gas is split into a recycle hydrogen stream and purge gas stream. The recycle gas is compressed and mixed with makeup hydrogen. The mixture is caustic and water washed to prevent sulfur buildup. A part of the cyclohexane liquid from the separator drum is recycled to the reactors to absorb a part of the heat of reaction. The remainder of the separator liquid, which is the net production of cyclohexane, is stabilized in a tray column and sent to product tankage. Cyclohexane is used primarily to make adipic acid which is used to synthesize Nylon 66, and to produce caprolactam, which is converted to Nylon 6.

8

PLANT SAFETY AND PRESSURE RELIEVING OPERATIONS

INTRODUCTION

This chapter describes the basic principles and procedures for the evaluation of overpressure potential in plant equipment, and for the selection, design and specification of appropriate pressure relieving facilities. The design of closed safety valves and flare headers is included in this chapter, but blowdown drums and flares are covered separately. To properly discuss this subject, the reader should become familiar with the following terminology.

Contingency - An abnormal event which is the cause of an emergency condition.

Emergency - An interruption from normal operation in which personnel or equipment is endangered.

Single Contingency - A single abnormal event causing an emergency.

Single Risk - The equipment affected by a single contingency.

Relief Valve - A relief valve is an automatic pressure-relieving device actuated by the static pressure upstream of the valve, and which opens in proportion to the increase in pressure over the opening pressure. It is used primarily for liquid service.

Safety Valve - A safety valve is an automatic pressure-relieving device actuated by the static pressure upstream of the valve and characterized by rapid full opening or pop action. It is used for gas or vapor service. (In the petroleum industry it is used normally for steam or air.)

Safety Relief Valve - A safety relief valve is an automatic pressure-relieving device suitable for use as either a safety valve or relief valve, depending on application. (In the petroleum industry it is normally used in gas and vapor service or for liquid.) Safety relief valves are classified as "Conventional" or "Balanced", depending upon the effect of back pressure on their performance.

Conventional Safety Relief Valve - A conventional safety relief valve is a closed-bonnet pressure relief valve that has the bonnet vented to the discharge side of the valve and is therefore unbalanced. The performance characteristics, i.e., opening pressure, closing pressure, lift and relieving capacity, are directly affected by changes of the back pressure on the valve.

Balanced Bellows Safety Relief Valve - A balanced safety relief valve incorporates means for minimizing the effect of back pressure on the performance characteristics; opening pressure, closing pressure, lift and relieving capacity. This is usually achieved by the installation of a bellows.

Pressure Relief Valve - This is a generic term applying to relief valves, safety valves or safety relief valves. It is commonly abbreviated to "PR Valve".

Pilot-Operated Pressure Relief Valve - A pilot-operated pressure relief valve is one that has the major flow device combined with and controlled by a self-actuated auxiliary pressure reliefs valve. This type of valve does not utilize an external source of energy.

Rupture Disc Device - A rupture disc device is actuated by inlet static pressure and is designed to function by the bursting of a pressure-retaining diaphragm or disc. Usually assembled between mounting flanges, the disc may be of metal, plastic, or metal and plastic. It is designed to withstand pressure up to a specified level, at which it will fail and release the pressure from the system being protected.

Maximum Allowable Working Pressure -As defined in the construction codes for pressure vessels, the maximum allowable working pressure depends on the type of material, its thickness, and the service conditions set as the basis for design. The vessel or equipment may not be operated above this pressure, or its equivalent stress at any other metal. temperature. However, ANSI B31.3, Petroleum Refinery Piping Code, permits variations above the maximum allowable working pressure, as follows:

(a) When the increased operating condition will not exceed 10 hours at any one time or 100 hours per year, it is permissible to increase the pressure rating at the temperature existing during the increased operating condition, by a maximum of 33%.

(b) When the increased operating condition will not exceed 50 hours at any one time or 500 hours per year, it is permissible to increase the pressure rating at the temperature existing during the increased operating condition, by a maximum of 20%.

While the above permitted variations are not used for selecting the pressure setting of pressure relief devices, these variations may be used for piping considerations such as startup, shutdown and emergency conditions, provided that the pressure vessel limits are not exceeded and that such variations are permitted by local codes.

Design Pressure - The design pressure is the pressure used as a basis for determining the minimum metal thickness of a vessel or equipment. The design pressure is never greater than the maximum allowable working pressure In the case where the actual metal thickness available for strength is not known, the design pressure is assumed to be equal to the maximum allowable working pressure. When the design pressure is being set, consideration must be given to pressure drop required for the fluid to reach the pressure relief device, liquid head and similar factors, as well as a suitable margin between operating pressure and pressure relief device set pressure. In final vessel construction, the maximum allowable working pressure may be slightly higher than the design pressure, due to the use of thicker plates or plates with higher than minimum specified tensile or yield stress.

Operating Pressure - The operating pressure is the gauge pressure to which the equipment is normally subjected in service. A process vessel is usually designed for a pressure which will provide a suitable margin above the operating pressure, in order to prevent leakage of the relief device.

Set Pressure - The set pressure (expressed as kPa gage or other increment above atmospheric pressure) is the inlet pressure at which the pressure relief valve is adjusted to open under service conditions. For a relief or safety relief valve in liquid service, the set pressure is to be considered the inlet pressure at which the valve starts to discharge under service conditions. For a safety or safety relief valve in gas or vapor service, the set pressure is to be considered the inlet pressure at which the valve pops under service conditions.

Cold Differential Test Pressure - The cold differential test pressure (in kPa gage) is the pressure at which the valve is adjusted to open on the test stand. This cold differential test pressure includes the corrections for service conditions of back pressure and/or temperature.

Accumulation - Accumulation is the pressure increase over the maximum allowable working pressure or design pressure (in kPa gage) of the vessel during discharge through the pressure relief valve, expressed as a percent of that pressure.

Overpressure - Overpressure is the pressure increase over the set pressure of the primary relieving device during discharge. It is the same as accumulation when the relieving device is set at the maximum allowable working pressure of the vessel. It is also used as a generic term to describe an emergency which may cause the pressure to exceed the maximum allowable working pressure.

Slowdown - Slowdown is the difference between the set pressure and the reseating pressure of a pressure relief valve, expressed as percent of set pressure.

Lift - Lift is the amount of rise of the disc in a pressure relief valve.

Back Pressure - Is the pressure on the discharge side of a pressure relief valve. Total back pressure is the sum of superimposed and built-up back pressures.

Superimposed Back Pressure - Is the pressure at the outlet of the pressure relief valve while the valve is in a closed position. This type of back pressure comes from other sources in the discharge system; it may be constant or variable; and it may govern whether a conventional or balanced bellows valve should be used in specific applications.

Built-up Back Pressure - Is the frictional pressure drop that develops as a result of flow through the discharge system after the pressure relief valve opens.

Spring Pressure - The spring pressure is equal to the set pressure minus the superimposed back pressure for a conventional PR valve. For a balanced bellows safety relief valve, the spring pressure equals the set pressure.

Open Disposal System - This is discharge piping of a PR valve which releases directly to the atmosphere.

Closed Disposal System - This is the discharge piping for a PR valve which releases to a collection system, such as a blowdown drum and flare header. However, a closed system can also be a process vessel or other equipment at a lower pressure.

Fire Risk Area - A process plant is subdivided into fire risk areas, each of which is the maximum area which can reasonably be expected to be totally involved in a single fire. This is used to determine the combined requirement for pressure relief due to fire exposure. This should not be confused with the areas used to determine fire water and sewer capacities, which are defined as plot subdivision areas.

DESIGN CONSIDERATIONS AND CAUSES OF OVERPRESSURE

This section discusses the principal causes of overpressure in refinery equipment and describes design procedures for minimizing the effects of these causes. Overpressure is the result of an unbalance or disruption of the normal flows of material and energy that cause material or energy, or both, to build up in some part of the system. Analysis of the causes and magnitudes of overpressure is, therefore a special and complex study of material and energy balances in a process system.

Although efforts have been made to cover all major circumstances, the reader is cautioned not to consider the conditions described as the only causes of overpressure. Any circumstance that reasonably constitutes a hazard under the prevailing conditions for a system should be considered in the design.

Overheating above design temperature may also result in overpressure, due to the reduction in allowable stress. A pressure relief valve cannot protect against this type of contingency.

The cost of providing facilities to relieve all possible emergencies simultaneously would be prohibitive. Every emergency arises from a specific cause or contingency. The simultaneous occurrence of two or more contingencies is unlikely. Hence, generally an emergency which can arise only from two or more unrelated contingencies (e.g., the simultaneous failure of both a control valve in the open position and cooling water along with the failure of an exchanger tube at the same time a control valve fails closed) is normally not considered for sizing safety equipment. Likewise, simultaneous but separate emergencies are not considered if the contingencies causing them are unrelated. Contingencies including external fire, are considered as unrelated. If there is no process, mechanical or electrical inter-relationship between them, or if the length of time elapsing between possible successive occurrences of these causes is sufficient to separate their effects, they are considered unrelated.

Every unit or piece of equipment must be studied individually and every contingency must be evaluated. The safety equipment for an individual unit is sized to handle the largest load resulting from any possible single contingency. When analyzing any single contingency, one must consider all directly related effects which may occur from that contingency. For example, should an air failure also cause a control valve in a cooling circuit to close, then both the air failure and the loss of cooling in that circuit are considered as part of the same contingency.

Likewise, if a certain emergency would involve more than one unit, then all affected units must be considered together. An example of this is the use of a stream from one unit to provide cooling in a second unit. Loss of power in the first unit would result in loss of this cooling in the second unit, and thus must be considered as part of the same contingency.

The equipment judged to be involved in any one emergency is termed a *single risk*. The single risk which results in the largest load on the safety facilities in any system is termed the *largest single risk* and forms the design basis for the common collection system, such as the flare header, blowdown drum and flare. The emergency which results in the largest single risk on the overall basis may be different from the emergency which form the basis for each individual piece of equipment.

While generally only a single contingency is considered for design purposes, there may be situations where two or more simultaneous contingencies should be taken into account; e.g., if there is some remote interrelationship between them, and pressures or temperatures developed could result in catastrophic failure. Such contingencies are also considered, and the *1.5 Times Design Pressure* rule may be applied in this situation.

Overpressure which may occur at normal or below normal pressures, as a result of reduced allowable stresses at higher than design temperatures, are also evaluated and appropriate protective features applied in the design. For example, such conditions may result from chemical reactions, startup or upset conditions. Likewise, low metal temperature must be considered, such as from autorefrigeration, to make sure that brittle fracture conditions do not develop.

The basis for design overpressure described in this section is related to the ASME Boiler and Pressure Vessel Codes and ANSI B31.3, Code for Petroleum Refinery Piping. Compliance with these codes is a requirement, or is recognized as the equivalent of a requirement in many locations. Where more stringent codes apply, the local requirements must be met. Therefore, local codes must be checked to determine their requirements. For example, some countries do not permit the use of block valves underneath pressure relief valves, unless dual valves with interlocks are installed. Also, in some cases, 20% accumulation under fire exposure conditions is not permitted, and accumulation allowed may be lower than the ASME Codes. In the United States, the ASME Code is mandatory, since it is a requirement under the Occupational Safety and Health

Act.

The essential steps in the design of protection against overpressure can be summarized as follows:

Consideration of Contingencies - All contingencies which may result in equipment overpressure are considered, including external fire exposure of equipment, utility failure, equipment failures and malfunctions, abnormal processing conditions, thermal expansion, startup and shutdown, and operator error. For each contingency, the resulting overpressure is evaluated and the need for appropriately increased design pressure (to withstand the emergency pressure) or pressure relieving facilities to prevent overpressure (with calculated relieving rates) is established.

Selection of Pressure Relief Device - From the range of available pressure relief valves and other devices, selection is made of the appropriate type for each item of equipment subject to overpressure. Instrumentation, check valves, and similar devices are generally not acceptable as means of overpressure protection.

Pressure Relief Valve Specification - Standard calculation procedures are applied to determine the size of the pressure relief valve required for the maximum relieving rate, together with other information necessary to specify the valve.

Design of Pressure Relief Valve Installation - Finally, the pressure relief valve installation is designed in detail, including location, sizing of inlet and outlet piping, valving and drainage, selection of open or closed discharge, and design of closed discharge system to a flare or other location.

Summation and Documentation of Contingencies - The Design Specification should include a tabulation of all contingencies considered, together with their relief requirements. Such a tabulation is helpful to assure that all contingencies have been considered and for selecting the contingency which sets the design of the collection system.

DETERMINATION OF CONTINGENCIES AND RELIEVING RATES

The first step in the design of protection against overpressure is to consider all contingencies which may cause overpressure, and to evaluate them in terms of the pressures generated and/or the rates at which fluids must be relieved.

All unfired pressure vessels must be protected by pressure relieving devices that will prevent the pressure from rising more than 10% above the maximum allowable working pressure (16% with multiple valves), except that when the excess pressure is caused by fire exposure a 21% rise in pressure is permitted.

The ASTM fired pressure vessels code requires pressure-relief devices to prevent pressures from rising more than 6% above the maximum allowable working pressure.

Design for overpressure protection in most cases consists of providing pressure-relief devices sized to handle the calculated relieving rates necessary to prevent emergency pressures from rising above the design pressure (plus allowable accumulation).

As an alternative means of protection, it is economical in some cases to specify an increased equipment design pressure which will withstand the maximum pressure that can be generated, without relieving any contained fluids. Also, in some cases, the cost of the collection system can be reduced by specifying higher design pressures which will permit a higher back pressure in the collection system.

In some cases where the ASME Code would not require pressure relief protection, the 1.5 Times Design Pressure Rule is applicable. This rule is stated as follows: *Equipment may be considered to be adequately protected against overpressure from certain low-probability situations if the pressure does not exceed 1.5 times design pressure. This criterion has been selected since it generally does not exceed yield stress, and most likely would not occur more frequently than a hydrostatic test. Thus, it will protect against the possibility of a catastrophic failure. This rule is applied in special situations which have a low probability of occurrence but which cannot be completely ruled out.*

In applying this rule, the capacity of the pressure relief system must also be sized to handle the quantity of fluid released at this pressure (together with other expected loads during this contingency), so that the built-up back pressure will not result in exceeding 1.5 times the design pressure. This additional load need not, however, be considered in calculations of flare and PR valve radiant heat levels.

Examples where the 1.5 Times Design Pressure Rule applies are:

1. Double contingencies which have some remote interrelationship and which could develop pressures or temperatures sufficient to cause catastrophic failure.
2. Remote contingencies such as heat exchanger tube failure or in case of closure of a CSO valve.
3. Plugging of a hydrofiner catalyst bed (some local codes consider catalyst

bed plugging as a contingency that requires normal pressure relief valve protection).

Fire as a Cause of Overpressure - Equipment in a plant area handling flammable fluids is subject to potential exposure to external fire, which may lead to overpressure resulting from vaporization of contained liquids. This hazard may exist even in items of equipment containing nonflammable materials.

Equipment to be Protected - All vessels subject to overpressure by fire must be protected by PR valves, with the following exceptions:

1. A PR valve is not required for protection against fire on any vessel which normally contains little or no liquid, since failure of the shell from overheating would occur even if a PR valve where provided. Examples are fuel gas knockout drums and compressor suction knockout drums. (Note: Some local codes require pressure relief valve protection for "dry drum" situations.)
2. Drums and towers, 6 m and less in diameter, constructed of pipe, pipe fittings or equivalent, do not require PR valves for protection against fire, on the basis that piping is not provided with protection against overpressure from this contingency. PR valves are required on such vessels, however, if overpressure can result from contingencies other than fire.
3. Except for special situations, pressure relief devices are not provided for fire exposure of heat exchangers, air fins, or piping, nor are the exposed surfaces of such items included for calculating the fire exposure heat input. Special situations may be congestion and substandard spacing, or unusually large equipment with normal liquid holdup over about 4 m^3 and/or which represents over 15% of the total wetted surface of the system to which it is directly connected for pressure relief.

Determination of Relieving Rate and Risk Area - In calculating fire loads from individual vessels, one assumes that vapor is generated by fire exposure and heat transfer to contained liquids at operating conditions. For determining PR valve capacity for several interconnected vessels, each vessel should be calculated separately, rather than determining the heat input on the basis of the summation of the total wetted surfaces of all vessels. Vapors generated by normal process heat input or compression, etc., are not considered. No credit is taken for any escape path for fire load vapors other than through the PR valve (which may be a common relief valve for more than one connected vessel), nor is credit allowed for reduction in the fire load by the continued functioning of condensers or coolers.

In order to determine the total vapor capacity to be relieved when several vessels are exposed to a single fire, a processing area is considered as being divided into a number of single fire risk areas. Plant layout should be designed in accordance with the spacing standards and must include accessibility for firefighting. Fire risk areas are established by the provision of accessways or clear spacing at least 6 m wide on all sides, which permit firefighting attack into all parts of the area and which limit the spread of fire. Clear space under pipebands, if more than 6 m wide, is considered as acceptable separation between fire risk areas. A single fire risk area need not, however, be considered to extend over an area greater than 465 m^2, provided that such an area includes the most severe combination of fire loads within a reasonable area pattern which might be involved in an actual fire. Use of unusual shapes such as "T"s should be avoided. The basis for this is that a fire involving a larger area would be of such magnitude that piping or other equipment failure would provide additional means of pressure release.

When a fire occurs, it is assumed that all fluid flow to and from the fire risk area has been stopped. Therefore, flow loads such as control valve failure or incoming feed streams are not additive to the fire load. Credit is not generally given to flow out through normal channels, since they could also be blocked during the fire emergency.

The total fire load is calculated for each fire risk area and used in the determination of the largest single risk release. Where the size of a closed discharge header is set by the fire relieving capacity requirement, advantage may be taken of reducing the requirement by providing fireproofing or insulation on vessels with large liquid inventories,

Protection from Fire Exposure and Pressure Relief Considerations - Pressure relief valves cannot protect a vessel that becomes locally overheated on an unwetted surface, although they do prevent the pressure from rising beyond the accumulation pressure of the valve. However, in such a case the vessel may be effectively protected against failure by either one of two methods for mitigating the effect of fire:

1. The Reduction of the Pressure by Depressuring - The reduction of pressure in a vessel exposed to fire has the advantages of not only reducing the metal stress to a value that will not result in failure, but also of reducing markedly the quantity of fuel that might feed the fire.
2. An Effective Limitation of the Heat Input - Application of firewater from fixed and mobile firefighting facilities is the primary method of cooling equipment which is exposed to fire. Further protection by fixed water deluge or spray systems, or fireproofing, is applied in areas of particularly high fire risk. However, in sizing PR valves, no credit is taken for reduced heat input due to application of cooling water, since it cannot be considered 100% effective in all fire conditions.

If a situation occurs which involves more than one fire risk area simultaneously (such as an entire Refinery or Chemical complex), it would be classed as a remote contingency event, and the 1.5 Time Design Pressure Rule may be applied.

Now let us consider utility failure as a cause of overpressure. Failure of the utility supplies (e.g., electric power, cooling water, steam, instrument air or instrument power, or fuel) to refinery plant facilities will in many instances result in emergency conditions with potential for overpressuring equipment. Although utility supply systems are designed for reliability by the appropriate selection of multiple generation and distribution systems, spare equipment, backup systems, etc., the possibility of failure still remains. Possible failure mechanisms of each utility must, therefore, be examined and evaluated to determine the associated requirements for overpressure protection. The basic rules for these considerations are as follows:

1. Interruptions of utility supply are considered only on a single contingency basis; i.e., corresponding to a failure of a single component of the generation or distribution system of one utility. Consideration must, however, be given to the direct effect of one utility on another. If a supply failure in one utility system, as a result of a single contingency, results in a complete or partial loss of another inter-related utility, then the dual failure must be considered. For example, in a plant where electricity is generated by steam turbine generators, loss of steam production may cause direct loss of power.
2. Failure is considered both on a local basis; i.e., loss of utility supply to one item of equipment (e.g., electric power to a pump motor) and on a general basis; i.e., loss of supply to all consuming equipment in a process unit (e.g., cooling water to all coolers and condensers). For the purpose of these pressure relief design considerations, a process unit is defined as one which meets all the following criteria:
    a. It is segregated within its own clearly identifiable battery limits boundary.
    b. It is supplied with each utility through one or two independent laterals from an offsite header.
    c. It constitutes a complete processing function.
3. For a process unit with its own segregated and self-contained closed PR disposal system, only a utility failure to that unit need be considered for the purpose of sizing the safety facilities. However, when two or more units share a closed disposal system (e.g., a common blowdown drum and/or flare), the design procedure must include consideration of the potential for single contingency utility supply failures to more than one of the units. Although such refinery-wide or plant-wide utility failures are not normally used as a basis for sizing the safety facilities, they must be evaluated. They need not be used as the

basis for sizing, provided that they can be realistically assessed as of very low probability. This entails the evaluation of reliability of generation and distribution systems and requires good engineering design and the inclusion of adequate backup features.

Examples of such backup features are given for each utility below. In designs where all pressure relief valves discharge into a closed collection system, because of environmental restrictions, a total failure of one utility deserves more careful consideration since there are no atmospheric releases which would tend to relieve the load on the closed system.

4. Evaluation of the effects of overpressure attributable to the loss of a particular utility supply must include the chain of developments that could occur and the reaction time involved. In situations where fluid flow stops due to failure of its utility supply, but is in parallel with equipment having a different energy source, credit may be taken for the unaffected and functioning equipment to the extent that operation is maintained and the operating equipment will not trip out due to overloading.

For example, consider a cooling water circulating system consisting of two parallel pumps in continuous operation, with drivers having different and unrelated sources of power. If one of the two energy sources should fail, credit may be taken for continued operation of the unaffected pump, provided that the operating pump would not trip out due to overloading. Similarly, credit may be taken for partial continued operation of parallel, normally operating instrument air compressors and electric power generators which have two unrelated sources of energy to the drivers.

Backup systems which depend upon the action of automatic cut-in devices (e.g., a turbine-driven standby spare for a motor-driven cooling water pump, with PLCI control) would not be considered an acceptable means of preventing a utility failure for normal pressure relief design purposes, even though their installation is fully justified by improved continuity and reliability of plant operations.

When the utility can realistically be classed as "Normally Reliable" (fewer than one failure in 2 years) then its failure may be classed as a Remote Contingency event; and the "1.5 Times Design Pressure Rule" may be applied.

Utility Failure Contingencies

In some cases, the loss of utility supply is not a direct cause of overpressure, but it initiates a plant upset or emergency (e.g., power failure leading to loss of tower reflux), which in turn may result in overpressure. Where necessary, reference should be made to the procedures for evaluating such upset or emergency situations and for determination of relieving rates, described later.

Electric Power

1. *Normal Individual and Process Unit Basis for PR Sizing Considerations* - The following single contingencies should be considered as the normal basis for evaluating overpressure that can result from electric power failures:

a. Individual failure of power supply to any one item of consuming equipment, such as a motor driver for a pump, fan or compressor.

b. Total failure of power to all consuming equipment in a process unit.

c. General failure of power to all equipment supplied from any one bus bar in a substation servicing one or more process units.

2. *Consideration of Plant-wide or Refinery-wide Power Failure* - Although not normally used as a basis for sizing pressure relieving facilities, the following general power failures on a plant-wide scale must be considered.

a. Failure of purchased power supply to the plant or refinery.

b. Failure of own generated power supply to the plant or refinery.

c. Total power failure in any one substation.

As discussed above, adequate backup features must be included to reduce the probability of such a major failure to an acceptably low level and balanced against the consequences, should such a failure occur. The following backup arrangements are normally required as a minimum:

d. Two or more feeders for purchased power.

e. Two or more parallel generators with a spare backup where power is generated in the plant or refinery.

f. More than one fuel to boilers generating steam for turbine generators.

g. Load shedding arrangements to maintain supply preferentially to critical consumers.

h. Secondary selective power distribution systems.

Cooling Water

1. *Normal Individual and Process Unit Basis for PR Sizing Considerations* - The following single contingencies should be considered as the normal basis for evaluating overpressure that can result from cooling water failures:

a. Individual failure of water supply to any one cooler or condenser.

b. Total failure of any one lateral supplying a process unit which can be valved off from the offsite main.

2. *Consideration of Plant-wide or Refinery-wide Failure* - Although not

normally used as a basis for sizing pressure relieving facilities, the following general cooling water failures must be considered:

a. Failure of any section of the offsite cooling water main.

b. Loss of all the cooling water pumps that would result from any single contingency in the utility systems supplying or controlling the pump drivers.

c. Loss of all the fans on a cooling tower that would result from any single contingency in the utility systems supplying or controlling the fan drivers.

As discussed above, adequate backup features generally are included to reduce the probability of these major failures to an acceptably low level. As a minimum, the following should normally be provided:

d. Multiple cooling water pumps with different drivers and automatic cut-in of the spare pump.

e. At least 30 minutes useful holdup in the sump of a cooling tower, based on loss of makeup watery with an independent lower water level alarm set at the 30-minute holdup level. This should provide sufficient warning so that corrective steps can be taken by operating personnel. Without such an alarm, makeup water failure may not be apparent.

f. Secondary selective power supply to cooling tower fan motors.

g. Adequate instrumentation and alarms to give warning of potential cooling water system failures, such as a low flow alarm on the makeup cooling water supply.

Application of the following should also be considered, in appropriate cases:

h. Multiple cooling towers.

i. Cross-connected or looped distribution headers, to gain the added spare capacity of multiple cooling towers.

Steam

1. *Normal Individual and Process Unit Basis for PR Sizing Considerations* - The following_single contingencies should be considered as the normal basis for evaluating overpressure that can result steam failures:

a. Individual steam failure to any one item of consuming equipment (e.g., turbine drivers, reboilers, strippers, ejectors, etc.) b. Total failure of any one lateral supplying a process unit from the offsite main.

2. *Consideration of Plant - wide or refinery wide failures* - Although not normally used as a basis for sizing pressure relieving facilities, the following general steam failures must be considered:

a. Failure of any section of the offsite steam main.

b. Loss of any one steam generator

c. Loss *of* purchased steam in any one supply line.

As discussed above, the probability of these major failures must be reduced to an acceptable low level by backup features, such as the following:

d. Multiple boilers with spare capacity fired by multiple fuels.

e. Adequate control and alarm systems, load shedding arrangements, etc.

f. Looped and valved distribution headers, so that only a portion of the system need be shutdown if a major leak occurs.

Instrument Air

1. *Normal Individual Process Unit Basis for PR Sizing Considerations* - The following single contingencies should be considered as the normal basis for evaluating overpressure that can result from an instrument air failure:

a. Loss of instrument air supply to any one individual control instrument or control valve. It is assumed that the correct air failure response occurs. In the case of "remain stationary" control valves, it is assumed that the drift action to the open or closed position takes place.

b. Total failure of any one valved lateral supplying a process unit from the offsite main. The correct air failure response of instruments and control valves is assumed (consider "remain stationary" valves as drifting to either the open or closed position, depending on the direction of the actuator).

2. *Consideration .of Plant-wide or Refinery-wide Failures* - Although not normally used as a basis for sizing pressure relieving facilities, the following general instrument air failures must be considered:

a. Failure of any section of the offsite instrument air main.

b. Loss of flow through any one set of instrument air driers.

As discussed above, the probability of these major failures must be reduced to an acceptable low level by appropriate backup features. "The following should be considered as minimum requirements:

c. Multiple air compressors with different drivers and automatic cut-in of the spare machine.

d. Multiple instrument air driers.

e. Automatic cut-ins from the maintenance air system.

f. Looped distribution headers.

Instrument Power - Instrument power failures are evaluated on a basis similar to that described for a power failure. Included in the normal considerations for PR sizing should be the failure of power supply to all instruments in and controlled from a single bus bar. Reliability features should include secondary selective power supply to control rooms, with emergency generator or battery backup for critical instruments and control computers. Critical controls should be able to continue operation independently of control computers.

Fuel - Fuel supplies to boilers, furnaces, gas turbine and engine drivers, etc., are designed with features such as multiple fuels, propane vaporizer backup, and a liquid fuel surge tank, to promote reliability. The failure of any one fuel to a process unit or utility generation facility is used as the basis for evaluating a potential overpressure.

Other Utilities - Failure of other utilities, such as inert gas to seals and purge systems, or compressed air (when used by the process) may in some cases determine pressure relief requirements. These cases are evaluated on a single contingency failure basis similar to the above.

In addition to failure as a result of their utility supply, items of equipment are subject to individual failure through mechanical malfunction. Such items include pumps, fans, compressors, mixers, instruments and control valves. The process upset resulting from such malfunctions (e.g., loss of a reflux pump) may in turn result in emergency conditions and the potential for overpressure. These contingencies should be examined and evaluated.

Operator error is considered as a potential cause of overpressure, although contingencies of gross negligence or incompetence are generally not considered. Gross negligence type items which generally are not included are a failure to remove blinds, vacuum due to blocking in of a steamed vessel during unit turnaround, bypassing of emergency devices, operating with closed block valve under a pressure relief valve and gross misalignment of process flow during startup. While closing and opening of CSO and CSC valves is also considered as gross negligence, the 1.5 times design pressure rule is normally applied in such cases.

We can consider a range of typical plant emergency situations which may result from utility failures, equipment malfunctions, or plant upsets, and which may result in equipment overpressure along with some guidelines for the evaluation of these emergency conditions and determination of relieving rates.

Failure of Automatic Control - Automatic control devices are generally actuated directly from the process or indirectly from a process variable; e.g., pressure, flow, liquid level, or temperature. When the transmission signal or operating medium fails, the control device will assume either a fully open or fully closed position according to its basic design, although some devices can be designed to remain stationary in the last controlled position. Such "remain stationary" control valves will, however, drift in the direction of drive and this must be considered. The failure of a process-measuring element in a transmitter or controller without coincidental failure of the operating power to the final controlled element should also be reviewed to determine the effect on the final controlled element. However, when examining a process system for overpressure potentials, one assumes that any one automatic control device could fail either open or closed, regardless of its action under loss of its transmission signal or operating medium.

When the control valve size is established by the designer, for example selecting a CV, it is assumed that this size valve is installed. If the control valve size is later increased it may be necessary to recheck the PR valve relief considerations.

1. Instrument Air Failure or Power Failure Action - Instrument air or power supply should be designed to minimize the hazards of the resulting emergency situation. Generally, this is achieved by specifying the closure of control valves in sources of heat input, water drawoffs and feed and product streams. Boxing in the plant equipment in this way, on the basis that any resulting overpressure is relieved by properly designed PR valves, is considered safer than the uncontrolled dumping of plant contents into tankage and other units.

A control valve in a heating system should generally be set to fail closed, to eliminate heat input. However, a control valve in a heating circuit, such as a furnace inlet control valve, should generally be set to fail open, to prevent overheating. Likewise, heat integrated circuits need careful scrutiny to determine the effect of the loss of cooling or heating. For example, a closed circuit system which is used for both heating and cooling could result in increased heat input if the stream which removes heat should fail.

2. Control Valve System Analysis - To evaluate system relieving capacity requirements for any single contingency (other than failure of the utility which affects valve movement), such as opening or closing of a single valve or utility failure, it is assumed that all control valves in the system which open under the contingency under consideration remain in the position required for normal processing flow. Therefore, credit may be taken for the normal capacity of these valves, corrected to relieving conditions, provided that the downstream system is capable of handling any increased flow. While some controllers may respond correctly by increasing valve openings, capacity credit should be taken only to the extent corresponding to their normal operating position. This will avoid subjective decisions involved in evaluating response times and effects of controller settings, such as proportional band, reset, and rate action. It is also compatible with the basic philosophy that instruments may not necessarily operate in an emergency situation. Should the valve close under the contingency considered, no credit should be taken for relief through this valve.

3. Failure of Individual Control Valve - The following individual control valve failures should be included in the analysis of control systems for determination of pressure relief requirements :

a. Failure in the wide open position of a control valve admitting fluid from a high-pressure source into a lower pressure system.

b. Failure in the wide open position of a control valve which normally passes liquid from a high-pressure source into a lower pressure system, followed by loss of liquid level in the upstream vessel and flow of only high-pressure vapor.

c. Failure in the closed position of a control valve in an outlet stream from a vessel or system.

d. A control valve failing open with its bypass open at the same time. In this case, the relieving rate and PR orifice should be based on the control valve failing with the bypass *50 %* open (based on areas) at 110 % of design pressure.

When analyzing such individual control valve failures, one should consider the action of other control valves in the system. In the first two cases above, credit may be taken, where applicable for the reduction in pressure of a high-pressure source due to net inventory depletion during the period that the downstream equipment pressure is rising to relieving pressure. However, the pressure relieving facilities must be sized to handle the calculated peak flow conditions.

4. Special Capacity Consideration - Although control devices, such as diaphragm-operated control valves, are specified and sized for normal operating conditions, they are also expected to operate during upset conditions, including periods when pressure relieving devices are relieving. Valve design and valve-operator capability must be selected to insure operation of the valve plug in accordance with control signals during abnormal pressure conditions. When wide discrepancies exist between normal and emergency conditions, the higher valve-operator pressure requirements should be considered.

When determining pressure relief requirements, one should calculate capacities of control valves for the relieving conditions of temperature and pressure, since these are in many cases significantly different from capacities at normal operating conditions. Downstream equipment must be analyzed under relieving conditions.

Cooling Failure in Condenser - In addition to the general failure of cooling water, failure of cooling water flow to each individual condenser or cooler should be considered. Normally, no credit is taken for the effect in a shell and tube condenser after the cooling stream fails, because it is time-limited and dependent on physical configuration of piping. However, if the process piping system is unusually extensive and uninsulated, the effect of atmospheric heat loss may be considered.

Total Condensing - The relief requirement is the total incoming vapor rate to the condenser. Credit may be taken for reduced relieving rate when recalculated at a temperature corresponding to the new vapor composition at the PR valve set pressure plus overpressure, and heat input prevailing at the time of relief. The overhead accumulator surge capacity at normal liquid level is typically limited to less than 10 minutes. If the duration of cooling failure exceeds normal liquid holdup time, reflux is lost and the overhead composition, temperature, and vapor rate may change significantly.

Also, the vapor load at the time of relief may be reduced below the normal design rate, due to the higher pressure, which may suppress vaporization at the time of the overpressure. Pinchout of a reboiler is such a situation. In such a case, steam pressure design conditions may be used, rather than the maximum steam pressure which could exist under pressure relieving conditions of the steam system. These changes can be taken into account, where appropriate, both for the equipment involved and for downstream equipment.

Partial Condensing - The relief requirement is the difference between the incoming and outgoing vapor rates at relieving conditions. The incoming vapor rate should be calculated on the same basis as stated earlier. If the reflux is changed in composition or rate, the incoming vapor rate to the condenser should be determined for the new conditions.

Air Fin Exchanger Failure - Loss of air fin exchanger capacity may result from fan failure or inadvertent louver closure.

1. Fan Failure - The effect of fan failure on heat transfer capacity will depend on the configuration of fans and tube bundles. For pressure relief design, the single contingency failure of one fan is considered. (Failure of all fans, that would result from a general power failure, would be included under utility failure considerations.) Credit may be taken for continuing heat transfer as a result of natural convection. As a guide, this may be taken as 10% of design capacity for condensing service and 30% of design capacity for cooling service, applied over the area of tube bundle affected by the fan failure.

2. Louver Failure - Inadvertent louver closure may result from automatic control failure or mechanical linkage failure. The effect on heat transfer will depend on the degree of restriction to air flow in the closed position and the configuration of louvers in relation to tube bundles (e.g., louvers may not be installed over all tube bundles in a given service).

For pressure relief design, the single contingency closure of one set of louvers is generally assumed. (Failure of all louvers, that would result from a general instrument air or power failure, would be included in utility failure considerations.) Credit may be taken for continuing heat transfer as a result of restricted air flow. As a guide, this may be taken as 10 %of design capacity for condensing service and 30 % of design capacity for cooling service, applied over the area of tube bundle affected by the failed set of louvers.

Special Conditions in Closed Circuit - Where heating or cooling is used in a closed loop circuit (e.g., hot oil and refrigeration system), consideration must be given to overpressure conditions that might occur on loss of fluid flow, loss of heat input or loss of heat removal.

Pumparound Flow Failure - The relief requirement is the vaporization rate caused by an amount of heat equal to that removed in the pumparound circuit. The latent heat of vaporization would correspond to the temperature and pressure at PR valve relieving conditions. "Pinchout" of steam heaters may be considered.

Absorbent Flow Failure - For lean oil absorption generally, no relief requirement results from lean oil failure. However, in a unit where large quantities of inlet vapor may be removed in the absorber, loss of absorbent could cause a pressure rise to relief pressure, since the downstream system may not be adequate to handle the increased flow. In such cases, the effect of this additional vapor flow into downstream equipment must be analyzed.

Loss of Heat in Series Fractionation System - In series fractionation, i.e., where the bottoms from the first column feeds into the second column and the bottoms from the second feeds into the third, it is possible for the loss of heat input to a column to overpressure the following column. Loss of heat results in some of the light ends remaining with the bottoms and being transferred to the next column as feed. Under this circumstance, the overhead load of the second column may consist of its normal vapor load, plus the light ends from the first column. If the second column does not have the condensing capacity for the additional vapor load, excessive pressure could occur.

Abnormal Process Heat Input - The required capacity is the maximum vapor generation rate at PR valve relieving conditions, including any noncondensibles produced from overheating, less the normal condensation or vapor outflow rate. In every case, one should consider the potential behavior of a system and each of its components. For example, the fuel or heating medium control valve or the tube heat transfer may be the limiting consideration. Consistent with the practice in other causes of overpressure, design values should be used for an item such as valve size. Where limit stops are installed on valves, the wide-open capacity should be used rather than the capacity at the stop setting. In shell and tube heat exchange equipment, heat input should be calculated on the basis of clean, rather than fouled, conditions.

Emergency Conditions in Integrated Plants - In integrated plants, a process upset in one unit may have an effect on other units (e.g., loss of flow of a pumparound which is used as a source of heat for reboiling other towers). All possibilities such as these must be carefully considered and the potential for resulting overpressure evaluated.

Abnormal Heat Input From Reboiler - Reboilers are designed with a specified heat input. When they are new or recently cleaned, additional heat input above the normal design can occur. In the event of temperature control failure, vapor generation may exceed the process system's ability to condense or otherwise absorb the buildup of pressure, which may include noncondensibles due to overheating.

Accumulation of Noncondensibles - Noncondensibles do not accumulate under normal conditions since they are released with the process vapor streams. However, with certain piping configurations, it is possible for noncondensibles to accumulate to the point that a condenser is "blocked". Such a condition could occur if an automatic vent control valve failed closed for a period of time. This effect is equal to a total loss of coolant, and thus need not be considered separately.

Water or Light Hydrocarbon Into Hot Oil - In limited situations, if the quantity of water present and the heat available in the process stream are known, the PR valve size can be calculated as a steam valve. For example, in the case of a hot feed accumulator operating above 100°C, it may be possible to estimate the pressure that would be developed if water were pumped into the vessel at normal feed rate. In most cases, however, the quantity of water is not known, even within broad limits. Also, since the expansion in volume from liquid to vapor is so great (approximately 1,400-fold at atmospheric pressure) and the speed of vapor generation is very great, it is questionable whether the PR valve could open fast enough to be of value, although a rupture disc could provide relief. Since rupture discs are not too practical, normally no pressure-relieving device is provided for this contingency. As such, proper process system design and operation, including startup conditions, are very important requirements to eliminate this possibility. Avoidance of water-collecting pockets, proper steam condensate traps, and double blocks and bleeds on water connections to hot process lines are some precautions that can be taken. Likewise, vessel bottom withdrawal connections should not be installed with internal extensions, which could result in water being trapped in the bottom of the vessel.

Internal Equipment Blockage by Collapsed Internals - Contingencies such as collapsed reactor bed vessel internals (e.g., fixed-bed reactor grids, coked catalyst beds, accumulation of catalyst fines, plugging of screens and strainers, lines blocked with sediments, etc.) should be considered to identify any overpressure situations that could result. The use of the "1.5 Times Design Pressure Rule" is applicable in such cases, if this is a remote contingency.

Manual Valve - Inadvertent operation of a block valve while the plant is onstream may expose equipment to a pressure that exceeds the maximum allowable working pressure. For the purpose of design, it is considered that only one manual valve is opened or closed and control devices are in their normal design position. A PR valve is required if the block valve is not locked or car sealed in the open position and if closure of such valve can result in overpressure.

The quantity of material to be relieved should be determined at conditions corresponding to the PR valve set pressure plus overpressure, not at normal operating conditions. Frequently, there is an appreciable reduction in required PR valve capacity when this difference in conditions is considerable. The effect of friction pressure drop in the connecting line between the source of overpressure and the system being protected should also be considered in determining the capacity requirement. If the valve passes a liquid which flashes or the heat content causes vaporization of liquid, this should be considered in determining PR valve size.

Startup, Shutdown and Alternate Operations - Not only design steady-state conditions, but also startup, shutdown, washout, regeneration, alternate feed stocks, blocked operations and other possible different conditions must be evaluated for overpressure protection.

Increased Plant Capacity - When an existing plant capacity is increased, the entire pressure relieving system should be reevaluated, even though new equipment has been added. For example, a unit operating at 120 % of design capacity may require additional pressure relieving capacity.

Overpressure in Different Equipment

In addition to equipment malfunctions which can cause process overpressure in associated equipment (e.g., overpressure in a fractionator due to failure of cooling water or reflux pump), certain support items to major processing equipment are themselves subject to overpressure due to mechanical reasons. Such items include heat exchangers, pumps, compressors, turbines and furnaces. The design of appropriate protection for them is discussed below.

Heat Exchanger Tube Failure - In a shell and tube heat exchanger, the tubes are subject to failure from a number of causes, such as thermal shock, vibration or corrosion. Whatever the cause, the possibility of the high-pressure stream overpressuring equipment in the low-pressure side of the exchanger is the result. Economic design usually dictates that the higher-pressure stream flow through the tubes, since this can result in a thinner shell; but this is not always the case. The pressure relationships must be known to permit proper evaluation of the results of tube failure. The ability of the low-pressure system to absorb this release should be determined. The possible pressure rise must be ascertained, to determine whether additional pressure relief will be required if flow from the tube rupture were to discharge into the lower-pressure stream.

The overall basis of design is that the low-pressure side of an exchanger must be protected by pressure relieving facilities if the design pressure on the high-pressure side is more than 1.5 times the design pressure on the low-pressure side and the piping on the low side cannot handle the discharge from a split tube without exceeding 1.5 times the low-pressure side design pressure. The entire low-pressure system must be examined for potential overpressure by a split tube. If the decision is to uprate the low-pressure side of the exchanger to eliminate the need for a PR valve, it is imperative that the entire low-pressure side be checked accordingly within the "1.5 Times-Design Pressure Rule".

The effect of a temperature change on the low-pressure side as a result of tube leakage is generally neglected when the design temperature of the low-pressure side is specified, since an increase in temperature is usually considered as a case of short-term allowable stress. However, where brittle fracture conditions might occur in the low-pressure side, various procedures to minimize the chance of tube failure (such as welding the tubes into the tube sheet and upgrading of tube material to obtain corrosion resistance) are acceptable alternative to that of specifying brittle fracture-resistant materials throughout the low-pressure side.

Pump and Downstream Equipment - A PR valve is required for a pump when the shutoff pressure of the pump is greater than the design pressure of the discharge piping, downstream equipment, or pump casing. Positive displacement pumps normally require such protection, while in most cases centrifugal pumps do not.

The capacity of a pump discharge PR valve should equal the capacity of the pump, except that for a centrifugal pump with an available pump curve, advantage can be taken of the reduction in pump capacity as it backs up on its performance curve, to reduce the PR valve size. Centrifugal pump shutoff pressure for determining whether downstream equipment needs PR valve protection is estimated as described below, unless higher design pressures are required by considerations other than pressure developed by the pump.

1. For a centrifugal pump, blockage downstream usually causes the suction pressure to increase; so the design pressure for the pump casing and downstream equipment is normally maximum suction pressure plus maximum pump differential pressure, which is developed at shutoff with no flow. When downstream blockage does not cause suction pressure to increase, design pressure is the greater of (a) normal suction pressure plus maximum pump differential pressure, or (b) maximum pump suction pressure plus normal pump differential pressure. The most accurate determination of maximum pump differential pressure is made from the curve for the specific pump. This may not be available during the design phase, so actual curves for similar pumps in similar service may be consulted. If this information is not available, the maximum differential may be approximated by using at least 120% of normal pump differential pressure. In such cases the design should be rechecked when the actual curve is available.

2. The design pressure of a steam-driven reciprocating pump and downstream equipment may be set by the maximum process pressure which the steam cylinder is able to produce at maximum steam pressure, in which case no pressure reliefs facilities are required. However, in most cases it is not economical to set the design pressure of downstream equipment as high as this maximum stalling pressure. In these cases, a PR valve would be required to protect the downstream equipment against overpressure. For reciprocating pumps driven by electric motors, PR valves serve the dual purposes of protecting the pump and downstream piping from overpressure, and protecting the driver from overload.

Other positive displacement pumps, such as rotary, gear, and diaphragm pumps, normally require PR valve protection for both the pump and downstream equipment. PR valves for all positive displacement pumps should have a capacity at least equal to the pump capacity.

For any pump requiring a PR valve for its protection or for protection of downstream equipment, the PR valve set pressure should be higher than the normal pump discharge pressure by 170 kPa or 10% of the set pressure, whichever is greater. Note, however, that in some cases a higher PR valve set pressure may be desirable to assure a sufficient differential when the pump is to be operated under lower than normal design pumping rate. This will recognize the higher pump discharge pressure under low flow conditions. In the case of reciprocating pumps, a greater differential than 10 % is desirable due to pressure surges.

Pump PR valves should be discharged to a closed system. In many cases they may be conveniently routed to the suction line or suction vessel.

Compressor and Downstream Equipment - PR valves are required for any compressor where the maximum pressure which can be generated during surge or restricted discharge conditions exceeds the design pressure of the discharge piping, downstream equipment or compressor casing.

For centrifugal compressors, it is usually economical to set the design pressure lower than the maximum possible pressure that the compressor can develop, and to provide appropriate PR valve protection on the discharge. In some cases (e.g., where the flow through a PR valve would be the largest release, or determining the size of a closed disposal system), it may be advantageous to set the design pressure of the compressor casing and downstream equipment equal to the maximum pressure that can be generated at the surge point, assuming the most severe combination of speed, molecular weight, suction pressure and temperature conditions that can occur as the result of a single contingency. However, this is generally not desirable.

For positive displacement compressors, discharge PR valves are nearly always required. Reliance on stalling of a reciprocating compressor is generally not economically attractive, since driver stalling pressures are usually quite high in comparison to operating pressure.

Low-pressure stage casings and interstage circuits on both centrifugal and positive displacement multi-stage compressors are not normally designed for full discharge pressure and must also be provided with overpressure protection.

Pressure relief valves for centrifugal compressors should be set higher than the normal operating pressure by 170 kPa or 10% of set pressure, whichever is greater. However, in the case of reciprocating compressors, a greater differential than 10% may be desirable because of pressure surges. Interstage PR valves should be set at least as high as the compressor settling-out pressure to avoid valve lifting during compressor shutdowns.

PR valve capacity should be equal to the compressor capacity at the emergency conditions. For centrifugal compressors, the combination of PR valve set point and relieving capacity should be such as to avoid surge conditions over the anticipated combinations of emergency conditions and operating variables. Compressor PR valves should discharge to an appropriate atmospheric or closed system, and not to the suction of the machine.

Steam Turbine - A PR valve is required on the steam inlet to any steam turbine if the maximum steam supply pressure is greater than the design pressure of the casing inlet. The PR valve should be set at the casing inlet design pressure and sized such that overpressure of the casing is prevented, under conditions of wide open steam supply and normal exhaust flow.

Protection of the exhaust end of steam turbines is dependent on whether they are in condensing or noncondensing service, as follows:

1. Condensing Turbine - The condenser and the exhaust end of a condensing turbine casing are not normally designed for full steam supply pressure. Therefore, in such cases, protection must be provided against overpressure which could result from loss of cooling water or other operating failure. The special type of pressure relief valve which is normally installed on the turbine exhaust for this purpose is illustrated in Figure 1. It has no spring end and is normally held closed by the vacuum conditions in the condenser but a supply of fresh water (not salt or brackish water) is required for the sealing system on older designs. The condenser vendor usually specifies and provides the PR valve.

The required relieving rates for these PR valves are based upon the steam rate to the turbine As a rough guide, the appropriate size can be checked against the following tabulation, which are recommendations of the "Standards of the Heat Exchanger Institute, Surface Condenser Section."

Maximum Steam to Turbine, kg/s	PR Valve Sze, mm	Maximum Steam to Turbine, kg/s	PR Valve Sze, mm
< 0.95	150	5.95 to 7.81	350
0.95 to 1.49	200	7.81 to 8.56	400
1.50 to 2.14	200	8.57 to 10.33	400
2.15 to 2.52	200	10.34 to 13.3	450
2.53 to 2.91	250	13.4 to 15.1	450
2.92 to 5.80	250	15.1 to 21.4	500
3.81 to 4.81	300	21.5 to 31.5	600
4.82 to 5.67	300	31.6 to 47.9	750
5.68 to 5.94	350		

2. Noncondensing Turbines - The casing of a noncondensing turbine is

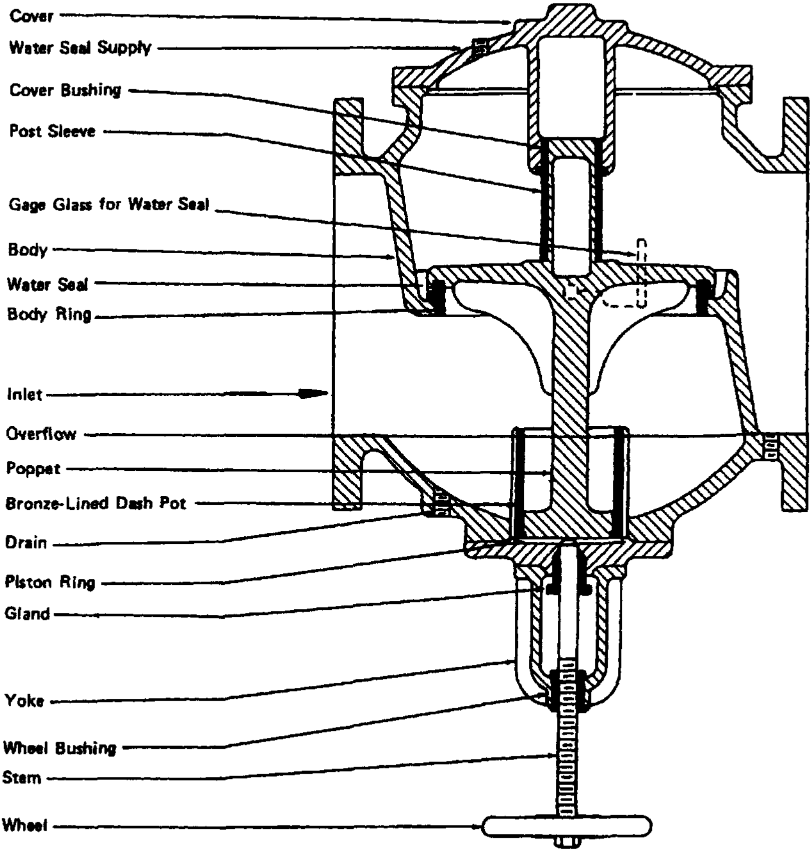


Figure 1. Bottom and top-guided low-lift valve for turbine exhaust.

usually not equipped for full steam supply pressure at the exhaust end. A pressure relief valve is therefore required at the turbine exhaust if the pressure in any part of the casing can exceed its design pressure as a result of exhaust valve closure, backpressure fluctuations or similar contingency. The PR valve should exhaust to the atmosphere, and its set pressure should be higher than the normal exhaust pressure by 170 kPa or 10% of set pressure, whichever is greater.

It is also important to note that the lowest design pressure of any section of the casing must be specified to be no lower than the pressure which it may be subjected to under the PR valve relieving conditions. This is necessary to recognize pressure drop within the casing. The PR valve should be sized to pass the normal steam flow to the turbine, but credit may be taken for steam flow which is withdrawn from an intermediate turbine stage if it would not be blocked by the same contingency as closure of the exhaust.

Exceptions to the foregoing requirements apply in the following cases:

a. Turbines exhausting to the atmosphere through open piping without valves do not require exhaust overpressure protection.

b. Turbines exhausting into an exhaust steam main through a top or side connection do not require exhaust overpressure protection, provided that the maximum inlet steam pressure does not exceed 1000 kPa gage; but a warning notice must be provided at the exhaust valve. A sentinel valve must also be installed on the turbine casing, to give an audible warning.

Furnaces and Boilers - Two potential forms of overpressure may apply to furnaces and boilers: overpressure of the firebox by forced-draft fans or tube rupture; and overpressure of tubes due to loss of fluid flow or outlet blockage, with resultant overheating.

1. Firebox Overpressure - The firebox of a forced-draft furnace and boiler is designed to withstand the overpressure that can be generated by the fans with dampers in their closed position. This needs to be specially checked when both forced and induced-draft fans are provided to discharge combustion products through heat recovery facilities, since higher than normal fan pressures may be used to overcome pressure drop. In the case of high-pressure process furnaces, a tube rupture could also be the cause of firebox overpressure.

2. Boiler Steam Side Overpressure - All fired boilers are normally provided with PR valves sized to relieve the full steam rate in the event of closure of the normal outlet.

3. Process Furnace Coil Overpressure - The coil of any furnace where the process flow can be stopped by inadvertent closure of a valve in the furnace

outlet (operator error) is subject to potential overpressure and tube failure due to overheating and consequent reduction in allowable stress. Unless such mechanisms of flow interruption (arising from a single contingency) can be effectively eliminated, the furnace must be provided with a PR valve on the coil outlet. In these applications, a PR valve functions by ensuring continuity of flow through the coil as well as by limiting the maximum pressure. It is therefore essential that the feed system be capable of providing the necessary flow at PR valve relieving conditions.

Overpressure and tube failure may also result from valve closure on the inlet side of a furnace, or from feed pump failure, etc, if the coil remains pressurized by downstream equipment. In these cases, however, overpressure occurs at or below the normal operating pressure (due to overheating at no-flow conditions), and a PR valve cannot provide the necessary protection.

The design features required to prevent furnace tube overheating and subsequent overpressure are as follows:

a. Low flow alarms, fuel cut-out on loss of process flow, and furnace feed reliability should be provided. The low flow alarm and fuel cutout will provide some protection against coil failure due to overheating from loss of feed or closure of an inlet block valve.

b. Control valves in furnace inlets should fail open, or remain stationary and drift to the open position, on actuating medium or signal failure, to prevent coil overheating.

c. If a block valve is installed in the furnace inlet, a PR valve is not required and this valve need not be CSO. However, if hand-operated valves or control valves are provided in each inlet pass to provide for feed distribution in a multipass furnace, protection against the loss of flow in any one pass should be provided by:

(1) A limit stop or open bypass around each valve so that flow cannot be completely interrupted, or

(2) A flow indicator on each pass with a low flow alarm.

The limit stops or bypasses should be sized to pass at least 25% of the design flow rate to that pass. Likewise, the low flow alarm and fuel cut-out should be set to operate when the flow falls to 25% of the design flow rate.

If a PR valve is provided on the furnace feed line, the valve should be located upstream of the orifice which senses low furnace feed flow and actuates the fuel cut-out so that the fuel will be cut out in case the furnace should be blocked at the outlet.

d. The provision of only a check valve in a furnace outlet line for emergency isolation purposes does not require the provision of pressure relieving facilities.

e. A PR valve normally is required on the coil outlet when it contains a block valve which cannot be CSO. The PR valve should be installed preferentially on the coil outlet, particularly when the feed to the furnace is 100 % vapor. The PR valve should be designed for the furnace outlet conditions, including purge, if necessary, to minimize product deposition in the PR valve inlet.

f. When an RBV is installed in a furnace outlet for emergency isolation purposes, it is necessary either to provide some flow through the tubes, or to rapidly cut out the firebox fuel to prevent tube rupture in case the RBV should inadvertently be closed. While the installation of a PR valve on the coil outlet is preferred, this may not always be practical, particularly on high-pressure furnaces, due to PR valve design limitations. Therefore, either of the following two alternatives (PR valve is preferred) is permissible:

(1) PR Valve

(a) The PR valve can be installed as a bypass around the REV or may discharge to the atmosphere or to another closed system. It should be sized for at least 25% of normal furnace flow and should have a set point 10% or 170 kPa gage above the normal operating pressure, whichever is greater.

(b) The furnace feed source must have a pressure/flow characteristic such that at least 25% of the normal flow will be maintained through the furnace if the RBV should be inadvertently closed and the PR valve then opens. Allowance should be made for pressure drop in the system due to fouling.

(c) If equipment upstream of the furnace is provided with a PR valve for protection against pump or compressor shutoff overpressure, then, to maintain flow, the set point should be sufficiently high that the valve preferentially will not open in the event that the furnace outlet PR valve is functioning.

(d) The PR valve should be designed for coil outlet conditions and should include a purge, if necessary, to minimize coke deposition in the PR valve inlet.

(2) Fuel and Low Cutout - The RBV should be instrumented so that when the valve stem travel reaches 50% of the closed position, it will:

(a) Stop fuel to the firebox, and (b) Stop furnace feed pumps or compressors.

A valved CSC bypass may be provided around the RBV for onstream testing to assure RBV functioning.

Overpressure Caused by Chemical Reaction

In certain processes, decomposition reactions or temperature runaways may occur as a result of feed or quench failure, overheating of feed, contaminants, or similar causes. High-pressure hydrogen processes and methanation reactions are examples. In other cases, air introduced for chemical reaction, decoking or catalyst regeneration may cause overheating if not carefully controlled.

Overheating can result in overpressure due to reduction of allowable stress. Therefore, the design must include monitoring and control features to prevent the occurrence of decompositions and runaway reactions, since conventional pressure relieving devices cannot normally provide protection against these contingencies.

Design temperatures must be specified at a margin above normal operating temperature sufficient to permit the detection of abnormal temperatures and institution of corrective measures by manual or automatic controls.

Appropriate design features may include feed-forward temperature control, high temperature alarms, high-temperature cutouts to stop feed flow and open a vent to atmospheric or closed system, adequate temperature monitoring through catalyst beds, etc.

Overpressure Caused by Abnormal Temperature

The inter-relationship of allowable stresses (and hence design pressure of equipment) and temperature must be taken into account when one is considering operating conditions which may exist during upsets, emergencies, startup or shutdown. The effects of high temperatures in some particular contingencies were discussed earlier. Low temperatures which may result from ambient conditions, autorefrigeration, etc., must also be evaluated to ensure that vessels which may be subjected to temperatures below embrittlement transition temperatures are designed such that allowable stresses under these conditions are not exceeded.

Overpressure Caused by Thermal Expansion

Lines or equipment which can be left full of liquid under non-flow conditions and which can be heated while completely blocked-in must have some means of relieving pressure built up by thermal expansion of the contained liquid. Solar radiation, as well as other heat sources, must be considered. Lines or equipment which are hotter than ambient when blocked in and which cannot otherwise be heated above the blocked-in temperature do not need protection against liquid thermal expansion. The following are common examples of some thermal expansion mechanisms.

1. Piping or vessels blocked in while filled with liquid, and subsequently heated by heat tracing, coils, or heat transfer from the atmosphere or other equipment.

2. A heat exchanger blocked in on the cold side with flow continuing on the hot side. This situation can sometimes occur during normal operation. For example, consider an exchanger train where plant feed is preheated by exchange with hot products, with the feed flowing from a pump (with a check valve in the discharge) through the exchangers to a flash drum. A level controller on the flash drum operates a control valve between the exchanger train and the flash drum. Thus, if the control valve closes, the feed system is blocked in and subject to thermal expansion.

3. Piping or vessels blocked in while filled with liquid at or below ambient temperature, and subsequently heated by direct solar radiation. Cryogenic and refrigeration systems must particularly be examined in this respect.

Protection against thermal expansion overpressure may be provided by the following methods:

1. Installation of a PR valve.

2. Addition of a small open bypass around one of the isolating valves which could be used to block in the equipment.

3. Ensuring that blocked-in equipment is drained of liquid.

Protection against thermal expansion overpressure should be included for specific applications in accordance with the following:

1. Heat Exchanger - Shell and tube and double pipe exchangers, where the cooler side can be blocked in full of liquid while the hot side fluid flow continues, must be protected by either:

   a. A manually operated bleeder valve, plus caution sign, when both block valves are located at the exchanger. If both block valves are not at the exchanger, then method (b) or (c) below must be used.

   b. A small permanently open bypass around one of the blocking-in valves.

   c. Installation of a PR valve.

If a bypass or PR valve is provided, it must also have sufficient capacity to relieve the vapor generated from the cold fluid by heat input from the hot side under design flow conditions, if vaporization could also cause the cold side to be overpressurized.

Note that a check valve in the piping upstream of the cooler side of an exchanger is considered as a block valve.

2. Piping - Sections of piping in any liquid service, whether onsite or offsite, which can be blocked in while liquid filled and subjected to liquid thermal expansion from subsequent heating, must be protected by either.
   a. Installation of a PR valve, or
   b. A small open bypass around one of the blocking-in valves, or
   c. Means for withdrawing liquid so that the line does not remain liquid filled.

Note that short sections of onsite or offsite liquid piping which can be blocked in generally do not need thermal relief valves, since a manual bleeder and safe means of drainage is normally available. Such short sections of piping would include those between isolating valves at control valve stations, pump manifolds, etc., and are less than 30 m in length. Also, check valve leakage is usually sufficient to compensate for thermal expansion due to solar heating.

3. Vessels - All vessels and equipment which can be blocked in while liquid filled and subjected to subsequent heating and thermal expansion by any of the mechanisms described above, must be protected by any one of the methods described for protecting piping.

In cases where vessels are provided with PR valves for protection against overpressure from fire exposure or an operating failure contingency, additional thermal expansion protection is not required.

4. Control Valve - Double-seated control valves are considered to pass sufficient leakage flow that equipment blocked in by such valves need not be provided with thermal expansion protection.

Thermal expansion PR valves should be specified with a set point as high as possible above operating pressure, to avoid inadvertent releases. For vessels the ASME Code must be followed. When the PR valve is installed only for thermal relief purposes on piping systems, settings up to 1.55 (ANSI 6-31.3) short time over stress conditions times the design pressure of the equipment being protected are desirable, if local codes permit. However back pressure must be allowed for if the PR valve discharges to a closed system, such that the 1.35 times design pressure is not exceeded. It is important to note that the capacity requirement is not easy to determine, but since the volume to be released in order to relieve pressure is small, a nominal 20 mm by 25 mm PR valve, Crosby JR or equivalent, is normally specified, particularly where only solar heating is involved. Larger sizes should be considered for long uninsulated above-ground pipelines of large diameter, and for large liquid-filled vessels or heat exchangers. Also, pressures generated due to vaporization as well as liquid expansion should not be overlooked.

A thermal expansion PR valve may be installed at any convenient point on the equipment or piping which it is protecting. Availability of a suitable discharge route will determine location in many cases. For onsite locations, thermal expansion PR valve releasing liquid must discharge into a closed system if the liquid within the system is above 31°C or above its flash point. This may be a flare header or the equipment on the opposite side of one of the blocking-in valves. Onsite thermal expansion PR valves releasing liquid may discharge to atmosphere at grade level in a safe location, such as to a dirty water sewer catch basin, if the liquid within the system is below below its flash point. Material above its flash point may be discharged to a sewer if the amount is small.

In offsite locations, thermal expansion PR valves may discharge to a flare header upstream of a knockout drum, if available, or to the equipment (e.g., a tank) on the opposite side of one of the blocking-in valves, or to the atmosphere. Atmospheric discharges must be at grade level in a safe location

Thermal expansion PR valves in onsite or offsite locations which release severely toxic fluids must discharge to an appropriate closed system.

Each thermal expansion PR valve should be provided with an inlet CSO valve (and an outlet CSO valve in the case of closed discharge) to permit isolation for inspection and testing, if permitted by local codes.

Vacuum as a Cause of Equipment Failure

Equipment which can operate under a vacuum, either continuously or intermittently, must be designed to withstand vacuum conditions or otherwise be protected, in addition to any requirements for operation at positive pressure. Vacuum conditions must be specified. In some cases, this may include piping as well as vessels or other equipment.

Other equipment which does not operate under vacuum, either continuously or intermittently, may be exposed to vacuum inadvertently, by contingencies such as the following:

1. Instrumentation malfunctions.
2. Drainage of nonvolatile liquid from a vessel without atmospheric venting or gas repressuring.

3. Shutting off steam at the completion of steam purging without admitting a noncondensible vapor (e.g., air at shutdown, fuel gas at startup).
4. Maloperation of valves.
5. Low ambient temperatures resulting in subatmospheric vapor pressure of certain materials (e.g., some alcohols and aromatics) in pressure storage.
6. Loss of heat input to closed process equipment handling low vapor pressure materials (e.g., fractionation of alcohols and aromatic solvents), while cooling continues such as by a condenser or through heat loss to the atmosphere.
7. Loss of heat input to waste heat boilers with resulting condensation of steam.
8. Loss of heat input to closed process equipment where appreciable quantities of steam are generated.

In some of the above situations, vacuum protection can be provided by the installation of vacuum reliefs devices, e.g., vacuum reliefs valves on tanks and spheroids. In other cases, reliance that no vacuum will be developed is placed on proper operation by the process personnel, e.g., not draining a nonvolatile liquid from a vessel without venting or repressuring, or not blocking in a steamed vessel during startup or shutdown. Generally, however, any vacuum condition which can be created during process operation such as abnormal cooling, low ambient temperature, loss of heat or blocked suctions on certain compressors must be considered in the design.

In designing for vacuum, credit may be taken for the fact that a vacuum situation may not create a full vacuum. Thus, not all vessels or equipment need be designed for a 100 % vacuum situation. For example, if the vacuum situation is created by a blocked suction in a compressor circuit and the vacuum created is limited to 10 psia by the compressor characteristics, the system needs only to be designed for 10 psia.

As a general rule, vacuum relief devices are permitted on offsite storage vessels handling clean finished products, since there is essentially no possibility of an internal ignition source. However, vacuum relief devices which permit breaking of a vacuum with inerts or flammable vapors are not permitted on process equipment, since they are not judged to be sufficiently reliable to provide adequate protection under all circumstances. Vacuum devices which permit air to enter may be considered, however, in cases where the equipment does not or cannot contain flammables; e.g., some steam systems.

Equipment which may be exposed to vacuum by any single process contingency should be designed for vacuum. In many cases, equipment with a substantial positive design pressure is capable of withstanding full vacuum, but the design should be checked. As an alternative to designing for vacuum, means of preventing the occurrence of vacuum is permissible in certain cases.

In general, vacuum vents and inert or gas repressuring systems are not considered an acceptable alternative to vacuum design for process equipment. Repressuring systems may be provided for process reasons, but they are not considered sufficiently reliable for equipment protection. Vacuum breakers are difficult to maintain tight and may admit air into the equipment.

Vacuum vents (air) are, however, used on refrigerated storage vessels for clean products, as a backup to a repressuring and low pressure compressor shutdown system. This is acceptable, since there is no internal source of ignition and the vacuum vents would function only if the primary vacuum protection (repressuring and compressor shutdown) failed to operate. Also, when vacuum protection is required on pressure vessels such as spheres and spheroids, vacuum relief devices admitting air are acceptable since the possibility of high vacuum is slight and there are no internal ignition sources.

In the case of low-pressure fractionation equipment, vacuum design is not required if all of the following can be met:

1. There is an adequately sized atmospheric vacuum relief device to prevent vacuum.
2. There are no pyrophoric materials such as peroxides, acetylides or sulfides, or other internal ignition possibilities such as static present, either normally or through abnormal conditions.
3. Process temperatures are at least 8°C below the autoignition point of the materials handled.

Fractionators which meet the above criteria and, therefore, do not need to be designed for vacuum must, however, also be provided with both the following:

1. A reliable automatic repressuring system (inert gas or hydrocarbon gas) to minimize the possibility of vacuum conditions occurring.
2. A board-mounted low pressure alarm, set at a positive pressure, and a remote control valve operator, so that repressuring gas can be admitted from the control center, should the automatic system fail to function.

As a rule, steam systems do not require special vacuum protection, since they are normally capable of withstanding vacuum developed if steam generation should fail and residual steam condense. However, low-pressure steam systems should be examined.

Generally, equipment is not designed for vacuum due to blocking in a vessel after steaming for shutdown reasons. Reliance is placed on good operations to insure that the vessel or equipment is not bottled up. Also, vacuum design is not required for spheres, spheroids and similar vessels when vacuum conditions result only from draining of water during startup, since good operating practice would require gas displacement or venting under such infrequent circumstances.

Cone roof atmospheric storage tanks must be provided with either a pressure-vacuum valve or an open vent, depending upon the flash point of the stored product.

Possible Pressurization Paths in Pressure Relief

The following paragraphs indicate the basis for design of permissible capacity limitation in flow paths through which an item of equipment can be overpressured by a source of high-pressure fluids.

Piping - Credit may be taken for pressure drop and maximum flow through piping which forms a pressurization path to a vessel on which a PR valve is to be installed for overpressure protection. The calculation must be performed for relieving conditions. However, where credit is taken for pressure drop in determining the PR valve setting, it must also be considered that under no-flow conditions pressures will equalize throughout the system at the PR valve set pressure. Also, if vapor relief is required through a liquid system, the dynamics of liquid displacement prior to vapor release must be taken into account. For example, in the case of two liquid-filled vessels interconnected by a bottom liquid line with a PR valve on the second vessel, liquid displacement through the interconnecting line may not be sufficient to protect the first vessel if the overpressure is caused by vapor generation.

Check Valve - A check valve is normally not an acceptable means of preventing overpressure by reverse flow from a high pressure source. However, where leakage through a check valve would not result in overpressuring by more than 1.5 times design pressure, a check valve is acceptable.

Restriction Orifice - In general a restriction orifice should not be used as a means of limiting the capacity of a pressurization path. In special cases, where large incentives apply (such as reducing the size of a flare system), a restriction orifice may be used, provided that all the following conditions are satisfied:

1. Physical means of preventing inadvertent removal of the R.0. is included, e.g., by welding the R.0. to the flange.
2. A warning against unauthorized removal is provided by means of the following:
   a. A warning sign plate welded to the R.0. handle or flange.

b. A note in the relevant documentation (specification sheets, flow diagrams and operating manuals.)

3. The installation is reviewed by the appropriate safety division.

Control Valve - A control valve with a limit stop to restrict the maximum opening is not normally acceptable as a means of limiting the capacity of a pressurizing path, since the stop may later be removed or the valve changed. Credit for the limiting capacity of a control valve in the wide open position may be taken only if all of the following conditions apply:

1. There is no bypass around the control valve. (If a bypass is provided, see (4) below.)
2. The control valve is assumed to be fitted with the maximum size plug and seat for its valve body size.
3. If the control valve size is critical to the overpressure protection of the downstream equipment, and must not be increased, then this is clearly noted in all relevant documentation (specification sheets, flow diagram, operating manual, etc.,) and a warning notice plate is welded to the valve body. In such cases, an actual check of the valve installed or purchased should be made during the startup review.
4. If there is a bypass around the control valve, downstream equipment must be protected so that its pressure would not exceed the 1.5 Times Design Pressure Rule, considering that the control valve is in the wide open position, and the bypass 50% open.
5. The installation is reviewed by the appropriate safety group.

Two or more pressure vessels connected by piping may be considered as a single unit for pressure relief purposes. If these conditions are satisfied, then pressure relieving facilities for the group may be located on any one of the vessels or on the interconnecting piping. However, pressure drops through the system under relieving conditions must be such that no vessel in the group is exposed to more than its design pressure (plus permissible accumulation) during any single contingency. In evaluating the relieving rates for fire when a group of vessels is considered as a single unit for pressure relief purposes, fire exposure must be assumed on vessels in the group which are in the same fire risk area.

Piping must be of adequate capacity to handle cumulative relieving rates through the system arising from any single contingency. A special situation occurs with submerged condensers where the condenser outlet is usually below the normal liquid level in the distillate drum. If a fire occurs near the drum, heat input to the drum will make the pressure rise in the tower-drum system. When the safety valve on the tower blows, flow will be out of the drum, forcing liquid up the tower overhead line. If heat input to the drum continues, considerable liquid static head may build up in the overhead line. This depends, of course, on drum volume and the vertical length of the overhead line. If this pressure increase can be greater than 200 % of the drum design pressure, then either a safety valve must be installed on the drum or the drum design pressure must be increased.

Valve Considerations

1. Car-Sealed Open Valve - Car-sealed open block valves are permissible in pressure relieving escape paths, provided that the single contingency of closing the CSO valve does not cause any equipment to be subjected to more than 1.5 times its design pressure. (Note: Use of CSO valves is not recognized in all local codes.) The requirements for CSO valves are as follows:

a. Car sealing is a means of ensuring the correct positioning of a block valve and should be used only where the full open position is an essential part of a pressure relief system or escape path. Generally, a CSO valve should not be used for process convenience, but confined to safety applications.

b. CSO valves must be line size hand-operated ball, gate or plug valves (motor operated valves may not be considered CSO valves).

c. CSO valves must have no cross-sectional area restriction or obstruction in the open position.

d. Gate valves must be installed with a special stem orientation.

e. CSO valves should be painted a distinctive color, normally yellow.

f. Plastic car seals or wire with lead seals may be used. Each plant must establish an effective procedure for regular checking and logging of car seals, and breaking of car seals should be permitted only by authorized persons.

g. Where double block valves are required for tight shut-off, the use of double block is acceptable if both valves are CSO.

2. Car-Sealed Closed Valve - In certain cases it may be advantageous to use car sealed closed valves, such as in a bypass around a fuel gas control valve used for furnace flameout protection.The bypass is provided so that the automatic shutdown system can be periodically checked for operation. Where CSC valves are used for other purposes, they are also limited to applications where inadvertent opening of the CSC valve would not overpressure the equipment by more than 1.5 times design pressure.

3. Control Valves - Use of a control valve in a pressure relief escape path is generally not permitted. In some unique cases where large incentives apply, a control valve with a minimum opening feature may be used, subject to all the following conditions:

a. The minimum opening, which may be a hole or cut out in the valve disc or plug, must be sized to pass the design relieving rate without overpressuring any equipment. Limit stops on valve stem movement are not an acceptable means of ensuring the minimum opening.

b. The fact that the minimum opening feature is included for pressure relief, and must not be changed, should be clearly noted in relevant documentation (specification sheets, mechanical catalog, flow diagrams, operating manual, etc.) and a warning sign plate should be welded to the valve body.

c. The installation should be reviewed by the appropriate safety personnel.

A three-way control valve which splits or combines two flows in a pressure relieving path is acceptable, provided that the total open cross-sectional area is at least equal to that of the adjacent piping in all valve positions. Also, the valve should be of the type that never has a fully closed port.

It is important to note that electrically or mechanically linked control valves for splitting or combining two flows are not acceptable in a relieving path.

4. Flow Meter Orifice Plate - A flow meter orifice plate is permissible in normal process flow pressure relieving path, provided that it can pass the required emergency flow without exceeding pressure limits of the upstream equipment. However, it is not acceptable in PR valve inlets and flare headers.

5. Check Valve - A check valve is acceptable in a process pressure relieving path, provided that (1) the valve opens in the pressure relieving direction, (2) the valve is of the swing-check or wafer type with no external actuation or dampen mechanism, and (3) the pressure drop is included in the system analysis.

A check valve is not, however, permissible in PR valve inlet or outlet piping, or in any flare or safety valve header.

The potential for a flow restriction in a pressure relieving path caused by coke formation, loosened refractory lining materials, catalyst fines, plugged catalyst beds, collapsed vessel internals, etc., must be evaluated, considering the equipment design details and operating experience in similar plants. If the occurrence of such restrictions can be anticipated, the pressure relieving facilities must be added upstream of the potential restriction. If the restriction is considered a remote but possible contingency, then the 1.5 Times Design Pressure Rule may be applied. The inclusion of special instrumentation (e.g., high-pressure alarms, pressure drop recorders, etc.) may not be used to eliminate the contingency.

Discharge of pressure relieving flows through auxiliary equipment which provides an open path to the atmosphere (e.g., the regenerator flue gas system on catalytic cracking units, or the barometric condenser, tail pipe and vent on a vacuum tower) must be similarly evaluated in terms of the possibility of internal restrictions.

Flame Arrester and Demisting Screen - Neither of these devices is permissible in a pressure relief discharge system, such as a flare header or knockout drum. The design of entrainment removal devices such as demisting screens must ensure that the relieving path cannot be obstructed, either by plugging of the screen, or by the screen itself becoming dislodged and blocking the relief valve inlet.

Parallel Flow Paths - In some cases, the pressure relief paths may be through parallel items of equipment manifolded together at their inlets and outlets, with individual CSO valves for isolation of one or more of the parallel paths for onstream maintenance or lower than design throughput conditions. An example of such a case is two parallel condensers in a fractionation tower overhead system. If the tower PR valve is also used for protection of the distillate drum, the pressure relief path would be through both parallel condensers. For design purposes, both paths can be considered as available for pressure relief, since all isolation valves are CSO. However, the system should also be analyzed on the basis that only one path would be available when the condenser is isolated for onstream maintenance. In this analysis, the single available path should be of sufficient size so that the distillate drum is not overpressured to more than 1.5 times design pressure. This basis should also be applied to other items of equipment, e.g., parallel filters or reactors.

APPLICATION OF PRESSURE RELIEF DEVICES

This section describes the various pressure relief devices that are commonly used, with their characteristics and criteria for selection. Basic calculation procedures for sizing PR valves are covered in subsequent discussions.

Conventional Safety Reliefs Valve

The pressure relief device used most often in refinery and chemical plant equipment is the spring-loaded, top-guided, high-lift, nozzle-type safety relief valve, which is illustrated in Figure 2. The spring is usually external and enclosed by a bonnet for weather protection, and the bonnet chamber is vented through an internal passage to the valve outlet.

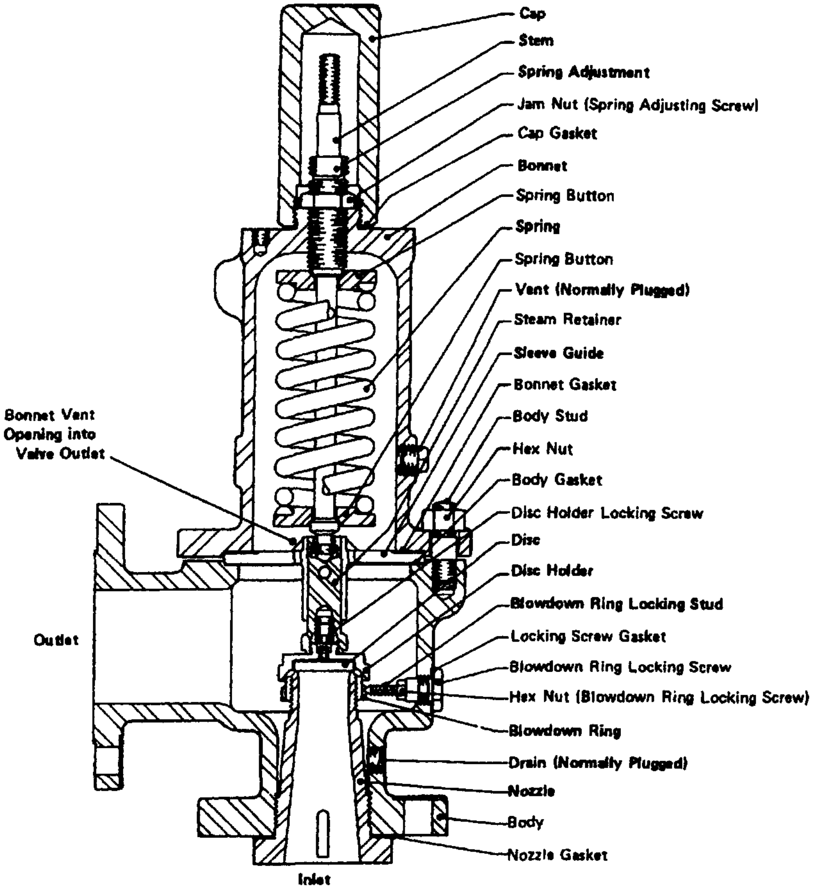


Figure 2. Typical conventional safety relief valve.

The operation and characteristics of a conventional safety relief valve are illustrated diagrammatically in Figure 3. The action of the valve as pressure rises from the initial normal operating pressure (assuming no back pressure) is described below. The effect of back pressure on PR valve operation is described later.

1. At a pressure below the set pressure (typically 95 to 98% of set pressure, depending upon valve maintenance and condition), some slight leakage ("simmer") may occur between the valve seat and disc. This is due to the progressively decreasing net closing force acting on the disc (spring pressure minus internal pressure).

2. As the operating pressure rises, the resulting force on the valve disc increases, opposing the spring force, until at the set pressure (normally adjusted to equal the vessel design pressure) the forces on the disc are balanced and the disc starts to lift.

3. As the vessel pressure continues to rise above set pressure, the spring is further compressed until the disc is at full lift. The valve is designed to pass its rated capacity at the maximum allowable accumulation (10% for contingencies other than fire, 16% if multiple valves are used; or 21% for fire exposure).

4. Following a reduction of vessel pressure, the disc returns under the action of the spring but reseats at a pressure lower than set pressure by an amount termed the blowdown (4 to 8% of set pressure). The blowdown may be adjusted within certain limits, by various means recommended by the valve vendor or manufacturer, to provide a longer or shorter blowdown.

Pressure relief valves for vapor service (i.e., safety valves and safety relief valves) are specifically designed for "pop" action. That is, they move to the full open position at only a slight overpressure, the valve remaining full open as overpressure builds up to the permissible maximum, at which condition the rated quantity is discharged. This "pop" characteristic is achieved by a secondary annular orifice formed outside the disc-to-nozzle seat. This causes additional disc area to be exposed to the operating pressure as soon as a slight lift occurs, accelerating the opening movement.

The kinetic energy of the flowing vapor, by action between the valve disc holder and the blowdown ring, adds to the opening force and causes the valve to "pop" open. This flowing kinetic energy also continues to act against the spring force as the fluid pressure returns to the pressure relief valve setting. This accounts for the fact that the PR valve reseats at a lower pressure than the set pressure; i.e., blowdown.

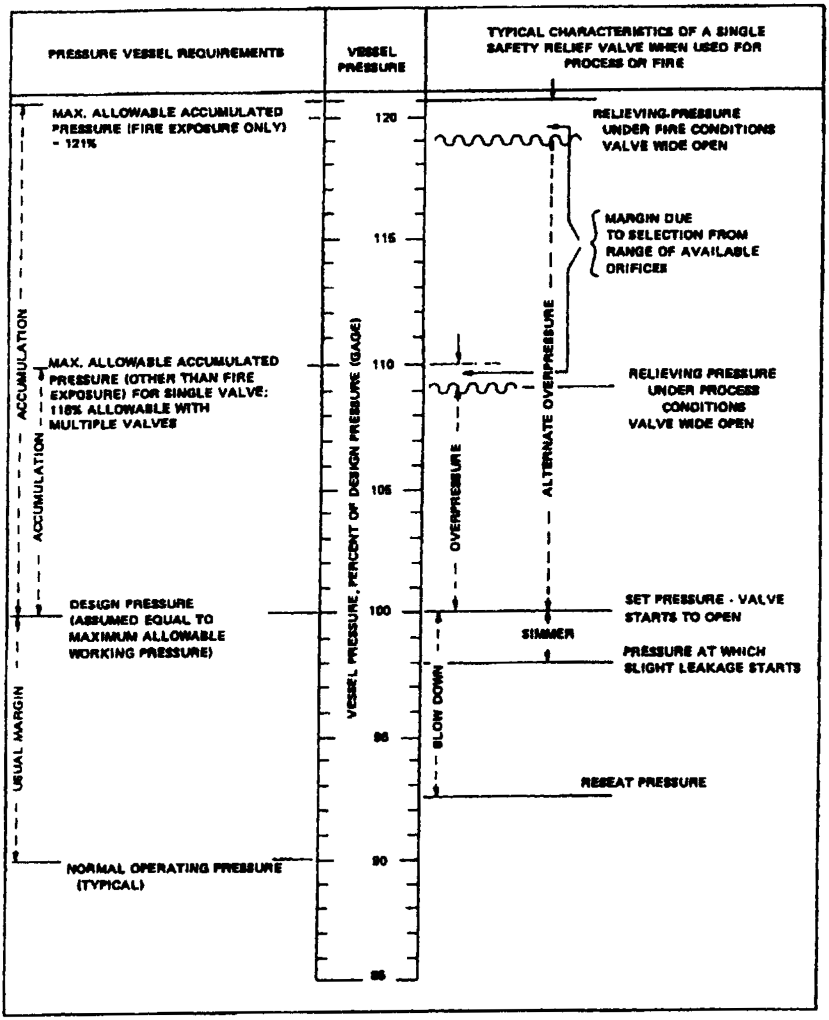


Figure 3. Characteristics of a typical safety relief valve.

As normally designed, vapor flow through a typical high-lift safety reliefs valve is characterized by limiting sonic velocity and critical flow pressure conditions at the orifice (nozzle throat), and for a given orifice size and gas composition, mass flow is directly proportional to the absolute upstream pressure.

Pressure relief valves in liquid service (i.e., relief valves and safety relief valves) have the characteristic of progressively increasing lift with rising inlet pressure until the full open position is reached at about 1% overpressure. This characteristic may vary between types and between makes.

Balanced Bellows Pressure Relief Valve

A typical bellows PR valve is illustrated in Figure 4. Bellows valves are generally specified where any of the following apply:

1. Superimposed back pressures are not constant. (Where back pressures fluctuate on a conventional valve, the valve may open at too low a pressure or permit the vessel pressure to exceed the equipment rating, depending upon back pressure fluctuation.)
2. The built up back pressure exceeds 10% of the set pressure, based on psig or it exceeds 21% of set pressure in the case of fire.
3. The service is fouling or corrosive, since the bellows shields the spring from process fluid. Note, however, that the bellows convolutions could also foul in extremely viscous service, such as asphalt, limiting the lift of the valve unless the valve is heated and insulated.

Although the bellows PR valve has the advantage of tolerating a higher back pressure than the conventional valve can, it should be recognized that the bellows is inherently a point of mechanical weakness which introduces some degree of additional risk, in case the bellows should fail and release process fluids through the vent. They should be avoided in services where the process temperature exceeds the autoignition point.

Balanced bellows PR valves may be satisfactorily used in vapor or liquid service up to a maximum back pressure (superimposed plus built-up) of 50% of set pressure, provided that the back pressure is incorporated into the sizing calculations . At higher back pressures, capacity becomes increasingly sensitive to small changes in back pressure. As an exception, back pressure up to 75% of set pressure may be used, provided that this disadvantage is recognized and the larger valve requirement can be justified. In no case should 75% be exceeded; leakage is likely if back pressure approaches 90%.

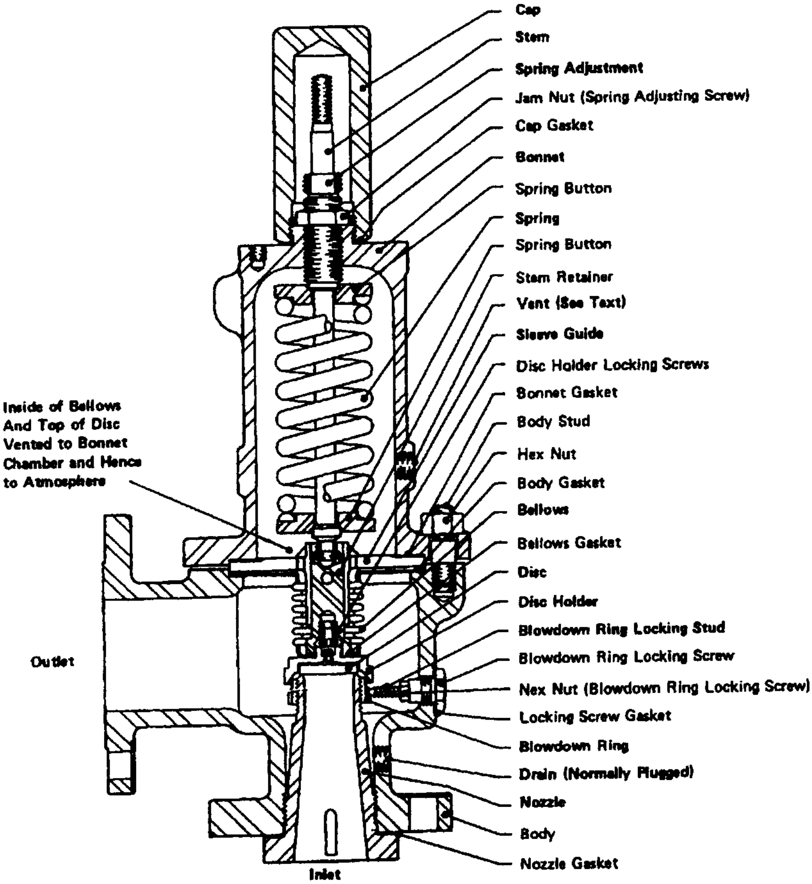


Figure 4. Typical balanced bellows safety relief valve.

In addition to the above back pressure limitations based on valve capacity, balanced bellows PR valves are also subject to back pressure limitations based on the mechanical strength of the bellows or bellows bonnet, or the valve outlet flange rating. The back pressure specified for the valve is governed by the lowest back pressure permitted by these various criteria.

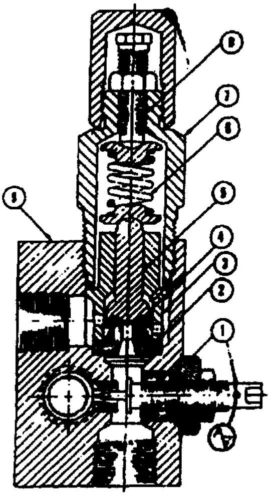
Bonnet Venting on Bellows Valves

In order to achieve the required balancing of the valve disc, the interior of the bellows must be vented through the bonnet chamber to the atmosphere, a 10 to 20 mm diameter vent hole is provided in the bonnet for this purpose. Thus, any bellows failure or leakage will permit process fluid from the discharge side of the valve to be released through the vent. Venting arrangements must therefore be carefully installed to meet the following requirements:

1. Manufacturer's shipping plugs must be removed from the bonnet vent holes before a new valve is commissioned.
2. Each PR valve must be installed so that the bonnet vent does not allow released vapors to impinge on lines or equipment, or towards personnel walkways. Where necessary, a short nipple and elbow should be added to direct flow away from such areas. In these cases, the vent piping should discharge horizontally to avoid entry of rainwater and debris, and should terminate in a position which is accessible for leak testing.
3. In cases where bellows failure would release flammable, toxic or corrosive liquids through the vent, a short nipple and elbow should be used to direct leakage to an open funnel which is piped to grade and ties into a catch basin or manhole with a sealed inlet connection.
4. Although venting to the atmosphere as described above is preferred, an alternative is to tie into a closed low pressure system, if available. This method may be used in the case of severely toxic fluids. Minimum length vent piping should be used. The effects of any back pressure must be thoroughly examined, since in such a case, superimposed back pressure is additive to the spring force.

Pilot-Operated Pressure Relief Valve

A typical pilot-operated PR valve is illustrated in Figure 5. Under normal operating conditions, the vessel pressure acts on the main valve seat at the bottom of the free-floating differential area piston, and by means of the pilot supply line is also applied to the top of the piston and under the pilot valve disc. Since the top area of the piston is larger than the nozzle area at the lower end of the piston, there is a large net load holding the piston down on the nozzle. Under static



PILOT VALVE

NO.	PART NAME
1	BLOWDOWN ADJUSTMENT
2	TWO PIECE NOZZLE
3	SPINDLE GUIDE
4	O-RING SEAT
5	SPINDLE
6	SPRING
7	BONNET
8	ADJUSTING SCREW
9	BODY

MAIN VALVE

NO.	PART NAME
1	BODY
2	NOZZLE
3	SEAT
4	SEAT RETAINER
5	LINER
6	PISTON
7	PISTON SEAL
8	SHIPPING SPRING
9	CAP
10	SUPPLY TUBE
11	PILOT VALVE
12	EXHAUST TUBE
13	LIFT ADJUSTMENT SCREW
14	DIPPER TUBE

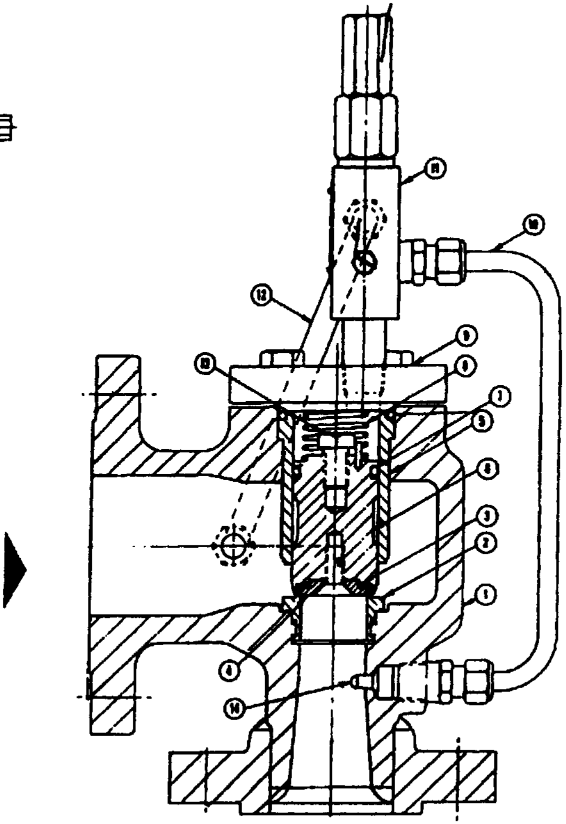


Figure 5. Typical pilot - operated PR valve.

conditions, this net downward sealing load increases as the vessel pressure increases and the valve approaches the set point. This is in contrast to the conventional spring-loaded valve, where the net force on the seat is reduced and the PR valve usually begins to simmer as the set point is approached.

When the set pressure of the pilot is reached, it opens and depressurizes the area above the piston, either to the atmosphere or into the discharge header, thus reducing the load on the top of the piston to the point where the upward force on the piston seat can overcome the downward loading. This causes an instant lifting of the piston to its full open position.

The soft-seated spring-loaded pilot valve is so constructed as to have a long built-in blowdown. For a flowing type pilot, at the point where the pilot supply line feeds the system pressure to the pilot relief valve, it passes through a variable orifice, which is also the main valve blowdown adjustment. When the pilot opens, the flow through the supply line causes an immediate pressure drop across the orifice. By adjusting the size of the orifice and thus the amount of pressure drop across it, one can obtain any desired system blowdown (5 to 7% is typical).

When the predetermined system blowdown pressure is reached, the pilot valve closes, full system pressure is restored to the dome above the piston, and the piston is quickly moved to the closed position. The pilot valve pressure sensing point may be located in the main valve inlet neck, or on the shell of the vessel being protected. The valve is less affected by inlet piping pressure drop in the latter case, as described below.

Various features may be included in a pilot-operated valve, such as backflow preventers, field testing connections, pilot line filters, manual or remote opening for depressuring purposes, etc. Details may be obtained from manufacturers' literature.

The advantages of pilot-operated PR valves are as follows:

1. A pilot-operated valve is capable of operation at close to the set point and remains closed without simmer until the inlet pressure reaches the set pressure. It is possible to take advantage of this by reducing the normal 10% margin between operating and set pressures, thus reducing vessel wall thickness requirements. This does not become a significant factor until design pressures are over 6900 kPa gage except perhaps for "debottlenecking."

2. Once the set pressure is reached, the valve opens fully and remains open, so long as the set point is exceeded. There is no need for overpressure or minimum flow to keep it open. Thus, it is not subject to chattering at low discharge rates.

3. If the pilot valve pressure tapping is taken directly from the vessel being protected (upstream of any inlet piping restrictions) a pilot-operated valve is less subject to the chattering which is normally associated with high pressure drop inlet piping. However, it is still desirable to design inlet piping for a maximum frictional pressure drop of 10% of set pressure, since some instances have been reported of resonance and chatter when higher pressure drops have been measured.

4. When the pilot exhausts to the atmosphere, a pilot-operated PR valve is fully balanced. Like the balanced bellows valve, therefore, its opening pressure is unaffected by back pressure, and high built-up back pressure does not result in chattering.

5. Pilot-operated valves may be satisfactorily used in vapor or liquid services up to a maximum back pressure (superimposed plus built-up) of 50% of set pressure, provided that the back pressure is incorporated into the sizing calculation. At higher back pressures, capacity becomes increasingly sensitive to small changes in back pressure. As an exception, back pressure up to 7% of set pressure may be used, provided that this disadvantage is recognized.

6. Ease of Setting - With a simple test connection, both the pilot pop and reseat pressures can be checked while the valve is in service.

7. Remote Depressuring - A pilot operated valve is sufficiently positive in action to be used as a depressuring device. By using a hand valve, a control valve or a solenoid valve to exhaust the piston chamber, one can open the pilot-operated valve and close it at pressures below its set point from any remote location, without affecting its operation as a pressure relief valve.

8. Pilot-operated PR valves can be specified for blowdown as low as 2%. This is an advantage for main gas pipeline and pressure storage applications, where the narrow range of pressure cycling minimizes product losses resulting from a release.

9. For applications involving unusually high superimposed back pressure, a pilot operated valve may be the only possible balanced valve that is commercially available, because of the mechanical limitations which apply to bellows.

Pilot-operated PR valves are subject to the following disadvantages:

1. They are not recommended for dirty or fouling services, because of plugging of the pilot valve and small-bore pressure-sensing lines. If the pilot valve or pilot connections become fouled, the valve will automatically open. In special cases where fouling is a function of entrained solids, this limitation may be countered by the use of a nonflowing pilot valve and a pilot line filter. With a nonflowing pilot valve, there is no normal flow in the pilot system and therefore solids entrainment is reduced.

2. They are normally limited to a maximum inlet temperature of 250 °C by the "D" ring piston seals.

3. Vapor condensation above the piston may cause the valve to malfunction unless special designs are used.

4. They are higher-cost in smaller sizes, compared with conventional and bellows valves.

The application of pilot-operated PR valves is limited to clean services, where they are an acceptable alternative to the bellows valve if a balanced characteristic is required. Their particular advantage is freedom from "simmer", which may permit a closer operating pressure/set pressure margin than normal.

Effect of Back Pressure

The effects of back pressure on PR valves and appropriate design considerations are described below. Figure 6 illustrates the forces acting on the valve discs of typical conventional and balanced bellows pressure relief valves. The existence of any superimposed back pressure on the top of a conventional valve disc exerts a closing force, in addition to the spring force, which opposes the opening force of the vessel pressure on the valve disc. The effect of superimposed back pressure would be to raise the set pressure if allowance for it were not made in the spring setting. By acting on the top of the disc of a conventional valve while in an open or partially open position, the existence of any back pressure exerts a closing force and results in reduced valve lift, and hence reduced discharge rate, assuming that the other variables remain unchanged. Excessive built-up back pressure acting on the top of the disc of a conventional PR valve may result in chattering.

Back pressure reduces the pressure drop across the orifice of any type of PR valve. This results in reduced discharge rates in the case of vapors, if the back pressure exceeds the critical flow pressure. For liquids, any back pressure reduces the pressure drop and results in a lower discharge rate.

A balanced bellows PR valve is one in which the closing force exerted by back pressure on the top of the valve disc and the back pressure exerted on the bottom of the disc are balanced by equal surface exposure, thus cancelling each other. The bellows shields the top of the disc from back pressure and the bellows area is vented to the atmosphere via the bonnet vent.

In the case of a pilot-operated valve, provided that the pilot valve exhausts to the atmosphere, the main piston is independent of back pressure and is thus also considered as a balanced valve. Balanced PR valves can be characterized by the following:

a. Opening pressure is unaffected by back pressure.

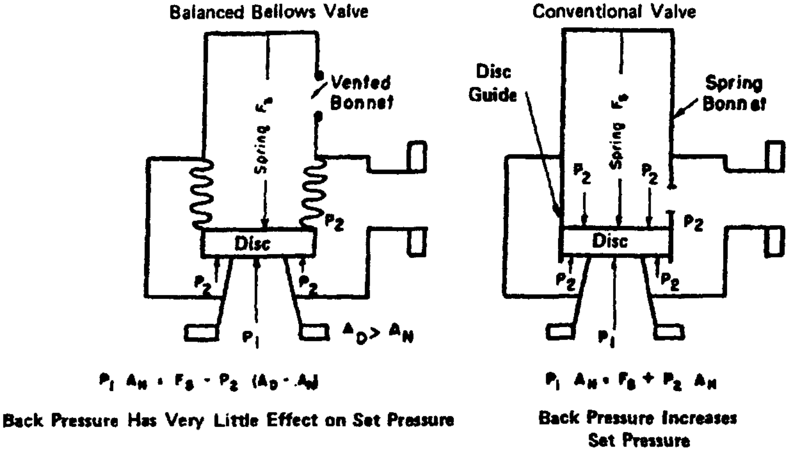


Figure 6. Forces acting on discs of conventional and balanced bellows safety relief valves.

b. They are less susceptible to chattering from built-up back pressure.
c. Valve capacity is affected by back pressure in the same way as for conventional PR valves.

Back pressure is included as a factor in PR valve selection and sizing accordingly:

1. Conventional PR valves subject to a constant superimposed back pressure are designed so that they will open at the required set pressure, by appropriate reduction in spring pressure.
2. Conventional PR valves that are exposed to variable superimposed back pressure will open at correspondingly variable pressure, since the superimposed back pressure is additive to the spring force.
3. Balanced bellows valves need no reduction in spring pressure to compensate for superimposed back pressure, and they can tolerate variable superimposed back pressure without an effect on opening pressure.
4. Conventional PR valves and discharge systems should be designed such that built-up back pressure does not exceed 10% of set pressure (both measured in psig), to avoid chattering problems. In the case where a pressure relief valve system is sized for fire conditions, with 21% overpressure, built-up back pressure up to 21% of set pressure is permissible. However, the lower rates resulting from other contingencies still must meet the 10% limitation.
5. Balanced bellows PR valves need not be restricted to the same built-up back pressure limit (10% of set pressure) as are conventional valves, since they are not subject to chattering from this cause. However, maximum back pressure is limited by capacity and in some cases by the mechanical design strength limitations of parts such as the outlet flange, bellows, or valve bonnet.

In general, the total back pressure on a balanced bellows pressure relief valve (superimposed plus built-up) should be limited to 50% of set pressure, because of the marked effect of higher back pressures on valve capacity, even when appropriate correction factors are used in sizing. In exceptional cases, such as a balanced bellows PR valve discharging into another vessel, total pressure up to 70% of set pressure may be used.

If the superimposed back pressure is less than the calculated critical flow pressure, the capacity of a conventional PR valve in vapor service is unaffected and back pressure is not a factor. However, builtup back pressure on a conventional pressure relief valve will affect its flow capacity and operating characteristics, and should not exceed 100% of its set pressure. If total back pressure (superimposed plus built-up) is greater than the calculated critical flow pressure, the capacity of a conventional PR valve in vapor service is affected, and total back pressure is incorporated into the sizing procedure. Any back pressure reduces the capacity of a conventional PR valve in liquid service, and the sizing calculation procedure is based upon the differential pressure across the valve, allowing for both superimposed and built-up back pressure.

It is important to note that back pressure affects balanced PR valve capacities in the same way as for conventional valves, and appropriate factors are included in the sizing procedures. They are subject to the same recommended limits of maximum total back pressure (superimposed plus built-up) as conventional valves. In the case of balanced bellows valves, mechanical considerations must also be evaluated, since they may limit the maximum permissible back pressure.

Valve Chattering

Chattering is the rapidly alternating opening and closing of a PR valve. This vibration may result in misalignment and leakage when the valve returns to its normal closed position, or, if it continues for a sufficient period, in mechanical failure of valve internals or associated piping fittings. Chattering may occur in both liquid and vapor service PR valves. The principal causes of PR valve chattering are an oversized valve, excessive inlet pressure drop, or excessive built-up back pressure.

In addition, a further mechanism of chattering may be introduced in some liquid PR valve installations if the response characteristics of a control valve in the same system are such that hunting between the two occurs. Generally, this can be eliminated by adjustment of instrument settings or by installation of two valves with staggered set points

Oversized Valve - "Pop" action PR valves in vapor service open at the set point by the action of static process pressure on the valve disc, and move to full open position at only a small overpressure. Typically, a flow through the valve equal to at least 2% of its capacity is necessary to keep the disc in the open position. At lower rates, the kinetic energy of the vapor flow is insufficient to keep the valve open against the action of the spring and it returns to the closed position, only to reopen immediately since the static pressure within the system still exceeds the set pressure. Chattering results from continuous cycling in this manner. It can occur when a "pop" type PR valve is too large for the quantity of fluid being discharged. In most cases, the use of multiple PR valves with staggered set points may be appropriate to eliminate this problem.

Liquid service PR valves are characterized by progressively increased lift with increasing inlet pressure, rather than the "pop" action of vapor service valves. Liquid service valves are, therefore, less liable to chattering at low relieving rates, and they will modulate down to about 10% of design flow.

Excessive Inlet Pressure Drop - A pressure relief valve starts to open at its set pressure, but under discharging conditions, the pressure acting on the valve disc is reduced by an amount equal to the pressure drop through the inlet piping and fittings. If this pressure drop is sufficiently large, the valve inlet pressure may fall below reseating pressure, causing it to close, only to reopen immediately since the static pressure is still above the set pressure. Chattering results from the rapid repetition of this cycle. To avoid this mechanism as a cause of chattering, inlet piping to PR valves should be designed for the lowest practical pressure drop (including entrance loss, and piping and isolation valve pressure drop), no more than 3% of set pressure at design relieving rate. This limitation, based on experience, is recommended by the major PR valve manufacturers. Friction pressure drop only is taken into account in this calculation. In unusual cases, such as with large PR valves in low-pressure vapor service, 5% inlet pressure loss may be used. The 3% maximum limitation is particularly important for valves in liquid service.

Excessive Built-up Back Pressure - Built-up back pressure resulting from discharge flow through the outlet system of a conventional PR valve results in a force on the valve disc tending to return it to the closed position. If this returning force is sufficiently large, it may cause the valve to close, only to reopen immediately when the effect of built-up back pressure is removed. Chattering results from the rapid repetition of this cycle.

To prevent chattering from this mechanism, conventional PR valve discharge systems should be designed for a maximum built-up back pressure of 10% of set pressure, when relieving with accumulation of 10%. In cases where pressure relief design is controlled by fire conditions, with 21% overpressure, built-up back pressures up to 21% of set pressure are permissible.

Where outlet pressure losses exceed 10%, bellows valves are often considered. However, substitution of a bellows valve for a conventional valve may not necessarily solve the chatter problem since debits associated with bellows valves reduce the rated capacity of this type valve. Hence, the valve has a tendency to become oversized depending on the amount of back pressure encountered. For this reason, revision of outlet piping to reduce the back pressure within the 10% limit is strongly preferred to the alternative of installing a bellows valve.

Multiple PR Valve Installation

In certain cases it is necessary to install two or more PR valves in parallel for a single service. These applications are described below, together with appropriate design guidelines.

The magnitudes of some large releases may be greater than the capacity of the largest single PR valve that is commercially available, necessitating the use of two or more valves. Even when a single PR valve is available, the relative cost of multiple valves should be considered. Above a certain size (typically 200 x 250 mm), structural and piping engineering considerations associated with the large piping and valves may result in a lower installed cost for two smaller PR valves. When two or more PR valves are installed for these reasons, they should be specified with staggered set points as described below to secure the additional advantage of minimizing chattering at low release rates.

In PR valve sizing it is always necessary to select the next larger commercially available orifice above the calculated size. Furthermore, a PR valve may lift as a result of various contingencies, any one of which requires a lower relieving rate than the design contingency. Both these factors affect the probability of a vapor PR valve chattering in service, since chattering (as described previously) is more likely to occur when the quantity of fluid being discharged is less than about 25% of its maximum capacity.

Where different contingencies of equal probability require substantially different capacities, it is always best to use two or more PR valves with staggered settings. For example, if one contingency required a capacity of 3 kg/s and another 12 kg/s, two PR valves would be used, with one of 3 kg/s and the other 9 kg/s minimum capacity. The lower capacity valve in this case would be at the lower staggered set pressure. When a fire contingency is the largest contingency and the next contingency is less than 1% of the fire relieving rate, multiple PR valves with staggered settings should always be used. However, when the fire contingency is the smallest load it is generally ignored. This is because fire is a remote contingency and chattering under fire conditions is not a significant concern.

When two or more PR valves are required in cases such as the above, capacities and set points should be specified in accordance with the ASME Code, as follows:

1. The code stipulates that when multiple valves are used, only one of them needs to be set at the maximum allowable working pressure (MAWP). The additional valves can be set at up to 105% of the MAWP. For design purposes the maximum allowable working pressure is assumed to be the same as the design pressure.

2. In addition, 3% tolerance on set pressure is permitted for valves nominally set at design pressure or MAWP. Thus, careful adjustment of the set point in the

field can provide some stagger, but this is not normally considered in the design. The matters of set point, stagger, tolerance, and overpressure are areas where other codes may differ from the ASME Code.

3. Where fire is the governing situation, a supplemental valve set as high as the Code permits is shown in Figure 7. This provides the maximum seat loading possible. This setting leaves only 9.1 percent overpressure available for sizing, although the pressure accumulation in the vessel is 21%.

4. The total relieving rate for some PR systems can be very high. This rate may be economically handled by one PR valve discharging liquid to a closed system and another, set at a higher pressure, discharging vapor to the atmosphere. The configuration should ensure that liquid is preferentially discharged to the valve set at the lower pressure, and that the possibility of entrainment through the vapor valve is minimized by providing a vapor space equal to at least 15 minutes of liquid holdup above the high liquid level alarm.

General Characteristics of the Spring-Loaded PR Valve

The additional features described below, available as means of improving a springloaded PR valve tightness below set pressure, may be justified for some applications. The soft seat design is illustrated in Figure 8. A synthetic "0" ring seal, or soft seat (e.g., of Viton or silicone rubber), may be incorporated into the valve disc seating area on either a conventional or a balanced bellows PR valve. With this device tight shutoff may be achieved closer to set pressure than with typical metal-to-metal seating. It is particularly applicable to difficult services, such as:

1. Operation close to set pressures, e.g., due to pressure fluctuations or pulsations. However, the normal 10% or 100 to 170 kPa between operating and set pressure should still be applied in designs.
2. Light, hard-to-hold fluids, e.g., hydrogen.
3. Presence of solids fines.
4. Vibrating equipment.
5. Corrosive fluids.
6. Nozzle icing under relieving conditions.

The Sta-Tite adapter is a device incorporated by some valve manufacturers into some conventional and balanced bellows valves to reduce leakage (or "simmer") below the set point. It functions by applying an additional auxiliary spring load onto the valve stem through a toggle linkage. The PR valve inlet pressure is applied through small piping to a piston which controls the toggle position such that increasing vessel pressure increases the seating force applied by the auxiliary spring. The linkage is designed such that when set pressure is reached, the toggle moves over-center and trips to a neutral position where no

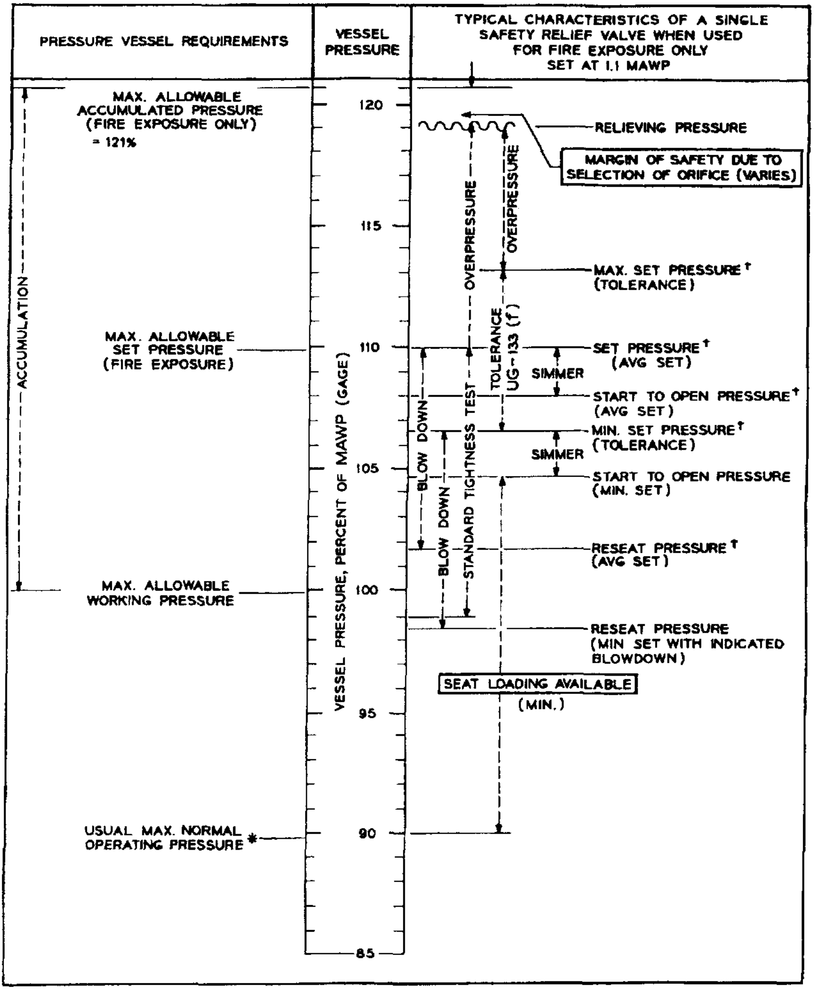


Figure 7. Pressure conditions for safety relief valve installed on a pressure vessel (vapor phase) supplemental valve used for fire exposure only.

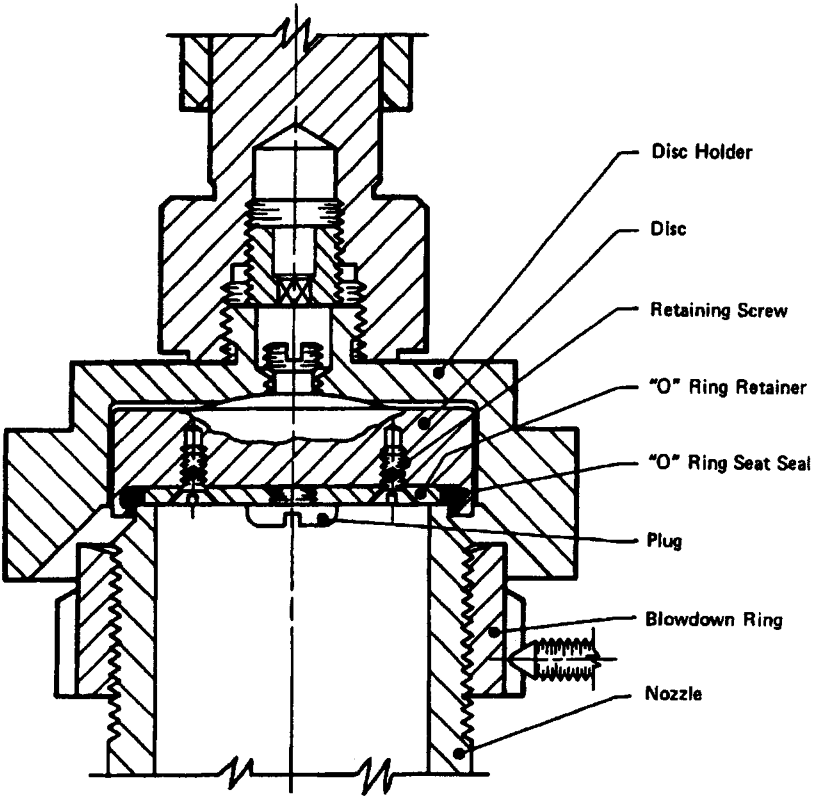


Figure 8. "O" ring seat seal safety relief valve.

auxiliary spring force is transmitted, leaving the PR valve free to operate normally. Manual reset of the toggle is necessary to reactivate the Sta-Tite adaptor after the valve has discharged. The device is fail-safe, since loss of piston pressure allows the toggle to collapse to a neutral position. It is applicable to PR valves operating close to set pressure, e.g., where pressure fluctuations or pulsations occur.

The Rupture Disc

The basic design of the rupture disc is illustrated in Figure 9. A rupture disc is a thin diaphragm mounted between flanges and designed to rupture at a designated pressure. Various types are available commercially. Normally, a prebulged disc is used, and support to preclude rupture under vacuum conditions may be included, if necessary. The most common disc materials are aluminum, monel, inconel and stainless steel, but other materials or coatings, such as carbon, gold and plastic, are available for particularly corrosive services. Rupture discs are usually available for bursting pressures up to 4800 kPa gage, and up to 41,000 kPa gage in the smaller sizes. Temperatures up to 480°C can be handled by the selection of appropriate disc materials. The advantages of a rupture disc, compared with pressure relief valves include:

1. There is no "simmering" prior to bursting.
2. It is more likely than a PR valve to be effective in relieving an explosion pressure.
3. It is less vulnerable to corrosion or plugging problems than a PR valve.
4. It is better able to handle high viscosity liquids and slurries.
5. It is suitable for applications where rapid depressuring is required, in addition to preventing overpressure.
6. Its initial cost may be less than that of a PR valve.

A rupture disc has the following disadvantages:

1. The entire contents of the protected system are lost when the disc ruptures. This necessitates a shutdown of the operation for replacement of the disc, unless a block valve is installed upstream.
2. The actual bursting pressure may deviate by + 5% of the designated bursting pressure in the "as new" condition, and the effect of fatigue in service may result in premature failure at lower pressures. It is therefore normal to allow a margin between operating and set pressure equal to 20 % of set pressure and thus a rupture disc may require a higher than normal equipment design pressure. This margin may, however, be reduced by the use of special disc construction.
3. It cannot be tested for accuracy of rupture point or operating service effects.

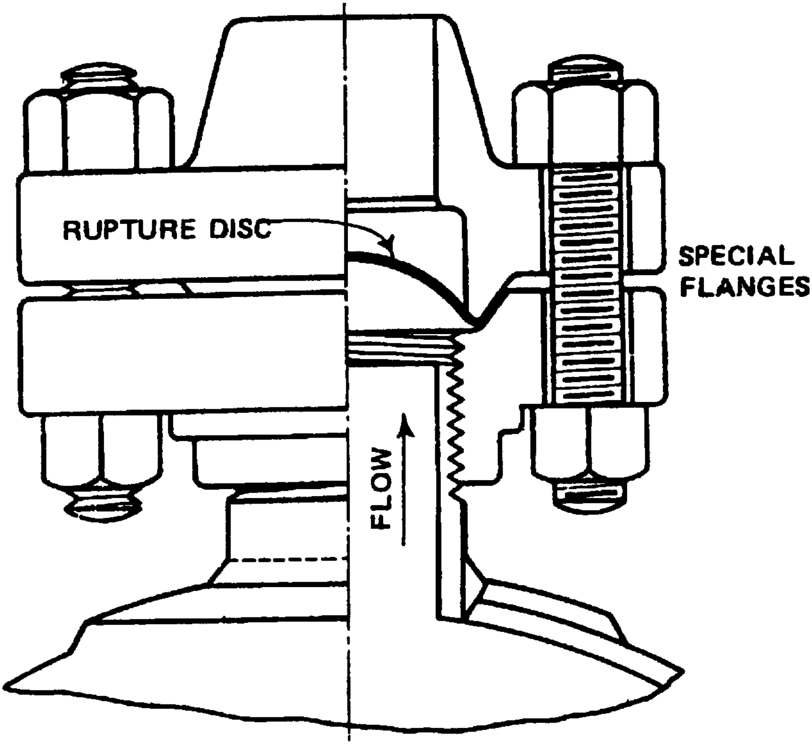


Figure 9. Typical rupture disc assembly.

A rupture disc is sometimes installed upstream of a PR valve, to avoid leakage of highly toxic or high-cost materials, or to minimize corrosion or fouling of the valve. Some means of detecting and relieving pressure buildup between disc and valve must be provided. The disc will not burst at its design pressure if back pressure builds up in this space due to leakage through the disc from corrosion or other cause. Some rupture disc holders include a threaded female connection on the downstream flange for fitting a vent. The simplest scheme is an open vent. Another acceptable scheme, limited to non-fouling services, uses a bleeder with a pressure gage and excess flow valve, which allows small amounts of leakage to be vented. If the disc bursts and the PRV opens from overpressure, the excess flow valve closes, and the gage continues to indicate line pressure after PRV reseats. A pressure gage with normally closed bleeder valve is sometimes used. This system depends on manual operation to prevent overpressure and its use is discouraged.

A rupture disc may also be used in some cases for protection against internal explosion overpressure. However, this is a matter of special design and the appropriate safety group should be consulted.

A rupture disc is sized to pass the required relieving rate determined by the same contingency procedures that are used for PR valves. When the inlet piping is short (2 diameters or less) and discharge is directly to the atmosphere, sizing can be based upon considering the rupture disc as an orifice. An orifice coefficient of 0.8 may be used for liquids and 0.6 for vapors when the expansion factor is ignored. For greater lengths of inlet piping or with outlet piping, sizing is based upon pressure drop through the whole system, allowing 50 diameters of equivalent straight pipe for the rupture disc assembly.

Explosion Hatches

A vessel which operates at essentially atmospheric pressure and is subject to internal explosion, such as an asphalt oxidizer, should be protected by an explosion hatch equivalent to at least 80% of the vessel cross-sectional area. The hatch consists of a hinged metal cover fitted over an opening on top of the vessel and sealed by its own weight. For vessels which normally operate at a slight positive pressure, a tight seal is achieved by the use of hold-down brackets with shear pins, rather than by increasing the weight of the hatch which would increase inertia and prevent quick opening. One or more hatches may be provided for a single vessel. Figure 10 illustrates a typical double hatch arrangement which can be designed to open up to 100% of the vessel cross-section area.

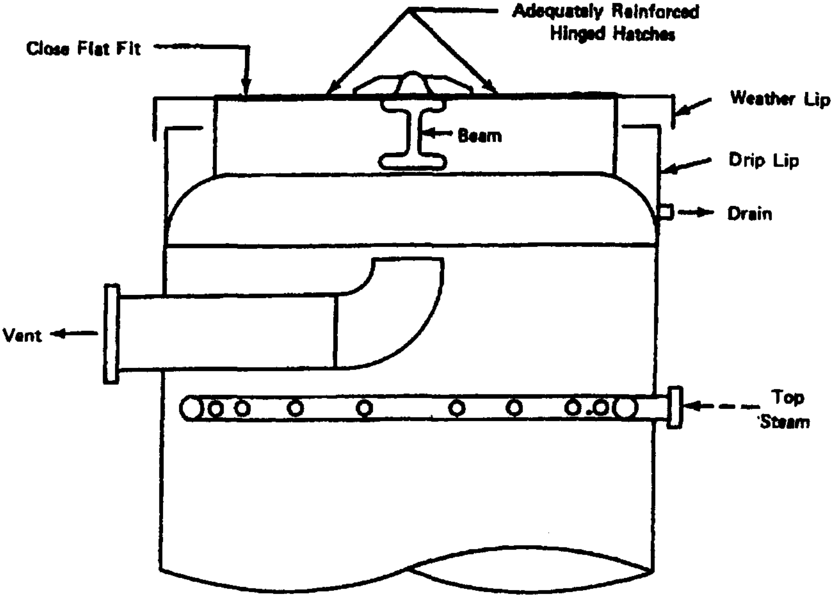


Figure 10. Explosion hatch for asphalt oxidizer.

Liquid Seals

In some cases, a hydraulic loop seal may be used for relieving overpressure on equipment operating at pressures slightly above atmospheric. Examples are certain naphtha fractionators with total condensation, where the seal would be installed on the distillate drum vapor space, discharging to the atmosphere. The seal consists of a simple U-tube containing a suitable liquid (normally water with the seal depth and diameter sized to pass the maximum relieving flow at the required design pressure. The following design features are ususally incorporated:

1. Continuous water makeup and overflow on the seal loop, to ensure that the seal is always made during normal operation, and reestablished after a blow.
2. Adequate winterizing, where necessary, to prevent freezing of the seal.
3. Safe disposal of the effluent seal water, considering possible contamination by process fluids.
4. Specific criteria which govern the acceptability of discharging process fluids to atmosphere.
5. Contingencies by which liquid hydrocarbon could be discharged through the atmospheric vent must be positively eliminated.
6. The vent line must comply with safe practices to prevent flashback and snuffing requirements.

Although liquid seals are relatively simple, reliable, and inexpensive, they are of limited application, because of the difficulty in meeting all of the criteria listed above. Also, they may not be too practical where vacuum conditions are encountered.

Pressure Relief Valve for Polymer Plugging Service

Process systems handling polymers and resins (e.g., butyl rubber or ethylene-propylene diene monomer rubbers) are often subject to plugging at dead-end locations such as PR valve inlets. In extreme cases, complete blockage of inlet piping and valve nozzle can result. This problem can be eliminated by the application of a flush-seated PR valve, in which dead-end areas are eliminated by placing the valve disc flush with the vessel wall, in the flow pattern of the contents.

VALVE SIZING AND SPECIFICATION PROCEDURES

The required relieving rate for a pressure relief valve is determined from consideration of the contingencies which can cause overpressure. Basic calculation procedures for determination of the PR valve size required to pass the design relieving rate are described below. Let us first consider the case of sizing for vapor service.

Critical and Subcritical Flow - The maximum vapor flow through a restriction, such as the nozzle or orifice of a pressure relief valve, will occur when conditions are such that the velocity through the smallest cross-sectional flow area equals the speed of sound in that vapor. This condition is referred to as "critical flow" or "choked flow".

$$V_C = 31.6\sqrt{KP_x/\rho_x} \tag{1}$$

In the above equation, V_c is the critical velocity (m/s), K is the ratio of specific heats (C_p/C_v) at inlet conditions, P_x is the pressure in the restriction at critical flow conditions (KPa, absolute - Note that this term is known as the “critical flow pressure”), and ρ_x is the density of the fluid at the critical flow temperature and pressure (kg/m^3).

The pressure drop which corresponds to critical flow is called the "critical pressure drop", i.e., P_1-P_x, where P_1 is the absolute upstream pressure.

If the pressure P_2 downstream of the restriction is less than the critical flow pressure, then the maximum obtainable flow which occurs at critical velocity is a function of P_1 and P_x, but is unaffected by P_2. If P_2 is greater than P_x, however, then the flow is termed "subcritical," and the rate is a function of P_1 and P_2. There are thus two equations for sizing PR valves in vapor service, depending on whether the flow is critical or subcritical.

In PR valve design, it is desirable to select a PR valve discharge location at a low enough pressure to permit designing for critical flow conditions, so that the relieving rate will be independent of minor back pressure fluctuations.

Determination of Critical Flow Pressure

The first step in sizing a PR valve for vapor flow is to determine the critical flow pressure P_x from the following equation:

$$P_x/P_1 = \{2/(K+1)\}^{(K/(K-1))} \tag{2}$$

For hydrocarbon vapors, values for P_x/P_1 can be obtained from Figure 11, which is sufficiently accurate under all conditions for PR valve calculations. The ratio of specific heats, K, for a particular substance, varies with pressure and

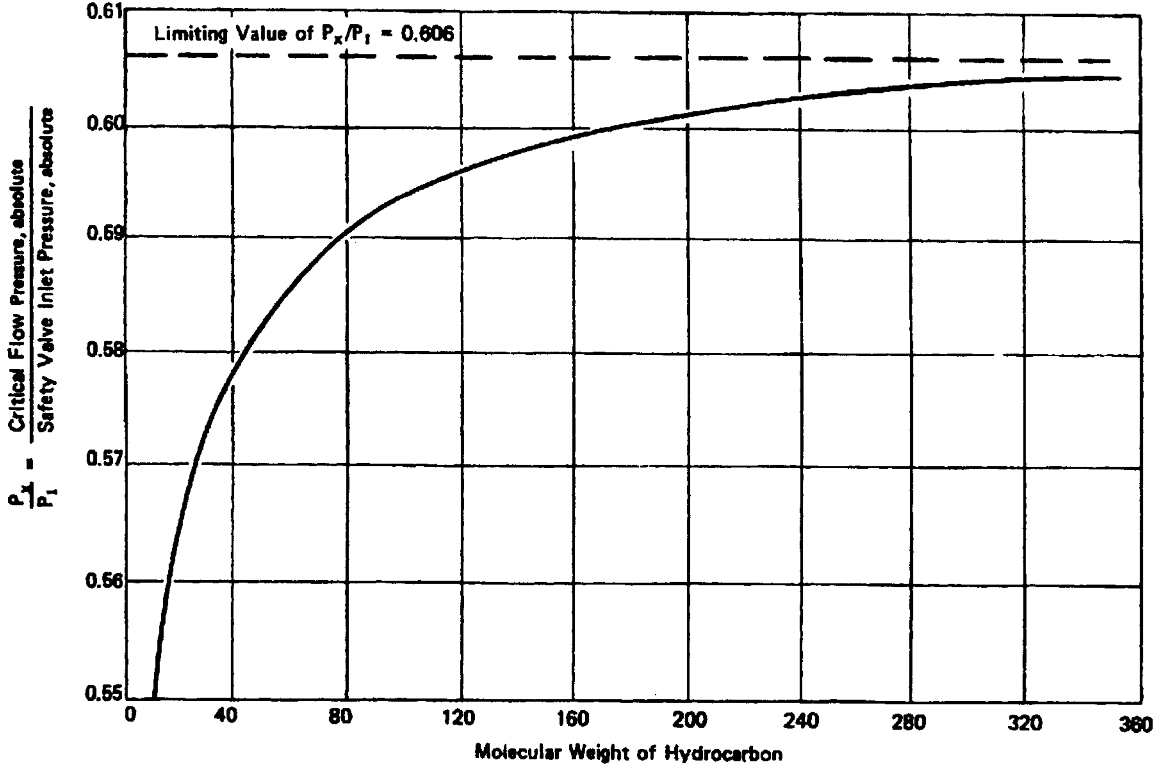


Figure 11. Critical flow pressure for hydrocarbons.

temperature, but for PR valve calculations, published values of K at 15 °C and one atmosphere may be used. K values for a range of common gases, with corresponding P_x/P_1 ratios, are listed in Table 1.

Table 1.Estimated Thermodynamic Properties of Various Substances at Standard Conditions

Gas	Specific Heat Ratio $K = C_p/C_v$	Critical Flow Pressure Ratio, P_x/P_1
Methane	1.31	0.54
Ethane	1.19	0.57
Propane	1.13	0.58
Butane	1.09	0.59
Air	1.40	0.53
Ammonia	1.31	0.53
Benzene	1.12	0.58
Carbon Dioxide	1.29	0.55
Hydrogen	1.41	0.52
Hydrogen Sulfide	1.32	0.53
Phenol	1.30	0.54
Steam	1.33	0.54
Sulfur Dioxide	1.29	0.55
Toluene	1.09	0.59

When the reduced pressure and temperature approach 1.0, P_x/P_1 approaches the limiting value of 0.606.

Sizing for Vapor - Critical Flow

For critical flow conditions (i.e., when total superimposed plus built-up back pressure is equal to or less than the critical flow pressure), the following equation can be used to calculate the required orifice area:

$$W = 2.11*10^{-8} C K_d K_b A P_1 \sqrt{M/\mu T_1} \tag{3a}$$

The parameter C is defined by the following expression:

$$C = 520\sqrt{K\left(\frac{2}{k+1}\right)^{(k+1)/(k-1)}} \tag{3b}$$

By combining the above two expressions, the following working formula is obtained:

$$W = 1.1*10^{-5} K_d K_b A P_1 \sqrt{\frac{M}{\mu T_1} K\left(\frac{2}{k+1}\right)^{(k+1)/(k-1)}} \tag{4}$$

where: W = mass flow rate. kg/s

K_d = Orifice discharge coefficient as recommended by PR valve manufacturers. Generally a coefficient of 0.975 is used for the common type PR valves. Manufacturer's literature should always be consulted in final value selection.

K_b = Correction factor for back pressure. Note: K_b for conventional safety relief valves may be taken as 1.0 for critical flow conditions. For balanced bellows valves, manufacturer's curve should be consulted for the recommended back pressure factor, K_b. K_b may be significant, i.e., less than 1.0, at back pressures less than the critical flow pressure. Figure 12 represents an average of the K_b values recommended by a number of PR valve manufacturers, and may be used when the make of balanced bellows valve is unknown. However, this chart is not reliable for set pressures less than 345 kPag and reference should be made to the PR valve manufacturer's catalog.

A = Effective orifice area, mm^2

P_1 = Flange inlet pressure at relieving conditions (including accumulation), kPa.

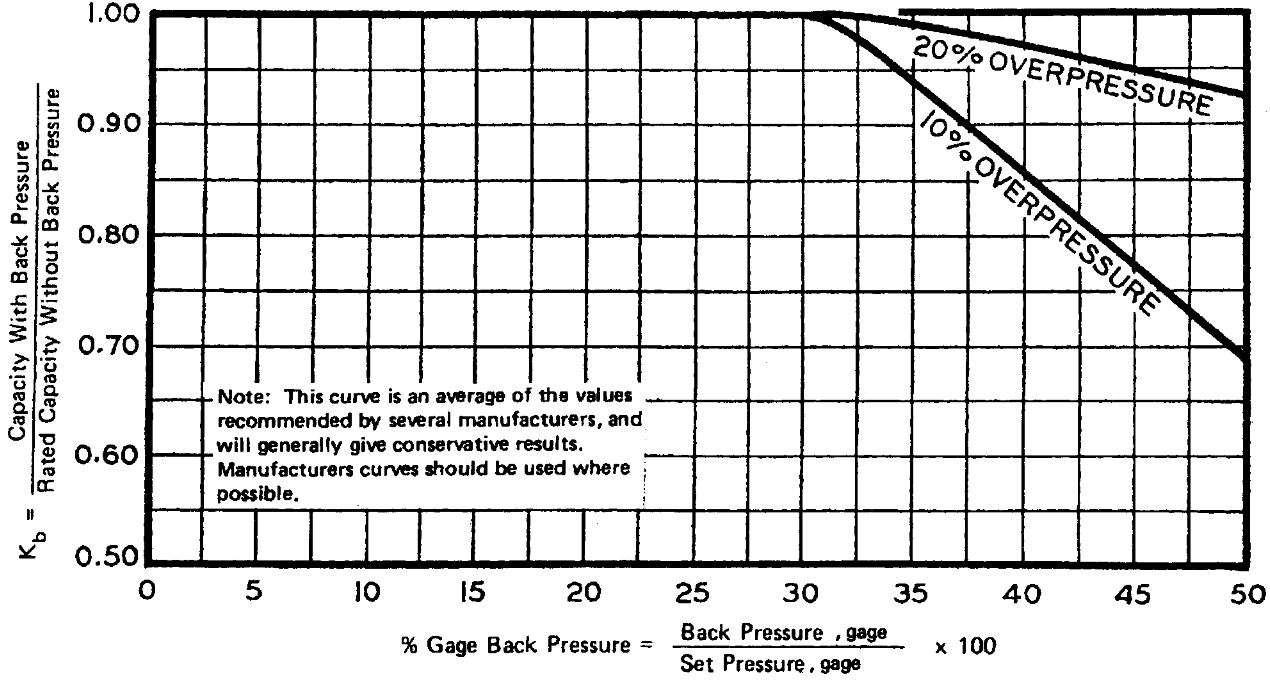
M = Molecular weight of vapor.

μ = Compressibility factor at inlet conditions

T_1 = Inlet temperature at relieving conditions, °K. Note: For the special case where a PR valve is being sized for fire, T_1 may be estimated by raising the mid-boiling point of the liquid to the inlet pressure P_1 using vapor pressure charts.

Figure III-2

VARIABLE OR CONSTANT BACK PRESSURE SIZING FACTOR, K_b, FOR BALANCED BELLOWS SAFETY RELIEF VALVES (VAPORS AND GASES)



Note: This curve is an average of the values recommended by several manufacturers, and will generally give conservative results. Manufacturers curves should be used where possible.

Figure 12. Variable or constant back pressure sizing factor K_b for balanced bellows safety relief valves (vapors and gases).

K = Specific heat ratio, C_p/C_v, at inlet conditions; given for some substances in Table 1. Note: Published values of K at 15 °C and one atmosphere may be used. If K is unknown, a conservative value of K = 1.001 may be used, in which case the factor C = 315. Note that a correction for non-ideal gases may be necessary.

C = ASME Unfired Pressure Vessel Code constant for a vapor; as described above, C is a function of the specific heat ratio, K. The equation given earlier for determining C is expressed in tabular form in Table 2.

The calculation procedure can be used to size thr PR valve:

1. For hydrocarbon vapors, equation (4) may be solved.
2. For steam, see "Sizing for Steam Flow" below.
3. For mixtures of hydrocarbon vapor/hydrogen/steam, see "Sizing for Hydrocarbon Vapor/ Hydrogen/Steam Mixtures," below.
4. For vapors other than the above, equations (3a) and (3b) should be applied.

The above procedure applies to both conventional and balanced bellows safety relief valves, provided that the appropriate back pressure factor K_b is used.

Sizing for Vapor - Subcritical Flow

For the exceptional cases of subcritical flow (e.g., where a PR valve is designed for a low set pressure and the total superimposed plus built-up back pressure exceeds the critical flow pressure), the following equation may be used applied:

$$W = 4.47 * 10^{-5} K_d A \sqrt{\left(\frac{P_1}{V_1}\right)\left(\frac{K}{K}-1\right)\left(\frac{P_2}{P_1}\right)^{2/K}\left[1-\left(\frac{P_2}{P_1}\right)^{(K-1)/K}\right]} \qquad (5)$$

where: P_2 = Total back pressure (superimposed plus built-up), kPa, abs.
V_1 = Specific volume at inlet conditions, m^3/kg. Other nomenclature is as defined after Equation (4), above.

Sizing for Steam Flow

For steam flow under critical flow conditions, the following equation is obtained by substituting the appropriate constants in Equation (4):

$$W = 1.46 \text{ x } 10^{-6} K_d K_b K_{gh} A P_1 \qquad (6)$$

where: K_{gh} = Correction factor for superheat, determined from Table 3, given below. (K_{gh} = 1.0 for saturated steam). Other nomenclature are as defined after Equation (4), above.

Table 2. Calculated Values of Parameter "C" for Flow Formula Calculations

$K= C_p/C_v$	C -Values	$K= C_p/C_v$	C -Values
1.001	315	1.36	352
1.02	318	1.38	354
1.04	320	1.40	356
1.06	322	1.42	358
1.08	324	1.44	359
1.10	327	1.46	361
1.12	329	1.48	363
1.14	331	1.50	364
1.16	333	1.52	365
1.18	335	1.54	368
1.20	337	1.56	369
1.22	339	1.58	371
1.24	341	1.60	372
1.26	343	1.62	374
1.28	345	1.64	376
1.30	347	1.66	377
1.32	349	1.68	379
1.34	351	1.70	380

For steam flow under subcritical flow conditions, the following equation

Table 3. Superheat correction factors for safety valve in steam services.

Set Pressure	Saturation Temp.	Total Temperature – Superheated Steam													
kPa, gage	°C	204°C	232°C	260°C	288°C	316°C	343°C	371°C	399°C	427°C	454°C	482°C	510°C	538°C	566°C
689	170	0.99	0.97	0.95	0.92	0.90	0.88	0.86	0.84	0.82	0.81	0.79	0.78	0.76	0.75
1034	186	0.99	0.97	0.95	0.92	0.90	0.88	0.86	0.84	0.82	0.81	0.79	0.78	0.76	0.76
1378	198	0.99	0.97	0.95	0.92	0.90	0.88	0.86	0.84	0.82	0.81	0.79	0.78	0.76	0.75
1723	208		0.97	0.95	0.92	0.90	0.88	0.86	0.84	0.82	0.81	0.79	0.78	0.76	0.75
2067	217		0.97	0.95	0.92	0.90	0.88	0.86	0.84	0.82	0.81	0.79	0.78	0.76	0.75
2412	224		0.98	0.95	0.92	0.90	0.88	0.86	0.84	0.82	0.81	0.79	0.78	0.76	0.75
2756	231			0.95	0.93	0.90	0.88	0.86	0.84	0.82	0.81	0.79	0.78	0.76	0.75
3103	238			0.96	0.93	0.91	0.88	0.86	0.84	0.82	0.81	0.79	0.78	0.76	0.75
3447	243			0.96	0.93	0.91	0.88	0.86	0.84	0.82	0.81	0.79	0.78	0.76	0.75
3792	249			0.97	0.93	0.91	0.88	0.86	0.84	0.83	0.81	0.80	0.78	0.76	0.75
4134	254			0.97	0.94	0.91	0.89	0.86	0.84	0.83	0.81	0.80	0.78	0.76	0.75
4482	258			0.98	0.94	0.92	0.89	0.87	0.84	0.83	0.81	0.80	0.78	0.76	0.75
4827	263				0.95	0.92	0.89	0.87	0.85	0.83	0.81	0.80	0.78	0.77	0.75
5171	267				0.96	0.92	0.89	0.87	0.85	0.83	0.81	0.80	0.78	0.77	0.75
5516	271				0.96	0.93	0.90	0.87	0.85	0.83	0.81	0.80	0.78	0.77	0.75
5861	275				0.97	0.93	0.90	0.87	0.85	0.83	0.82	0.80	0.78	0.77	0.75
6206	278				0.97	0.94	0.90	0.88	0.85	0.83	0.82	0.80	0.78	0.77	0.75
6550	282				0.98	0.94	0.91	0.88	0.86	0.84	0.82	0.80	0.78	0.77	0.75
6890	286					0.95	0.91	0.88	0.86	0.84	0.82	0.80	0.78	0.77	0.75
7585	292					0.96	0.92	0.89	0.86	0.84	0.82	0.80	0.79	0.77	0.76
8274	298					0.97	0.93	0.90	0.87	0.84	0.82	0.81	0.79	0.77	0.76
8964	304					0.98	0.94	0.91	0.87	0.85	0.83	0.81	0.79	0.78	0.76
9653	309						0.96	0.92	0.88	0.85	0.83	0.81	0.79	0.78	0.76
10340	314						0.96	0.93	0.89	0.86	0.84	0.82	0.80	0.78	0.76
12060	326							0.96	0.91	0.87	0.84	0.82	0.80	0.78	0.77
13780	336							0.98	0.93	0.89	0.85	0.83	0.81	0.79	0.77
15514	346										0.86	0.84	0.81	0.79	0.77
17230	354										0.88	0.85	0.82	0.80	0.78
18961	362													0.80	0.78
20670	369													0.81	0.79

from Equation (5) may be used:

$$W = 9.33*10^{-5}\ K_d\ K_{sh}\ A\ \sqrt{\left(\frac{P_1}{V_1}\right)\left(\frac{P_2}{P_1}\right)^{1.54}\left[1-\left(\frac{P_1}{V_1}\right)^{0.23}\right]} \qquad (7)$$

Note that for fired boilers, where the safety valve installation must comply with the ASME Code for Power Boilers rather than the code for Unfired Pressure Vessels, the allowable accumulation is only 6% instead of 10%. Reference also should be made to the K_d and A definitions, above.

Sizing for Hydrocarbon Vapor/Hydrogen/Steam Mixtures

The sizing of a PR valve for this type of mixture at critical flow conditions can be based on the average molecular weight and the use of Figure 11 for the determination of critical flow pressure, followed by the use of Figure 13 for determination of the required orifice area. This procedure is sufficiently accurate for PR valve sizing purposes. Unfortunately, Figure 13 cannot be used for average molecular weights of less than 10, and in such cases Equation (2) must be used to calculate P_x, using an average value of K for the mixture. An average K-value is also required if Equation (5) is applied.

When an average K for a hydrocarbon vapor/hydrogen/steam mixture is required, the following procedure for estimating this K-value is recommended:

1. Blend the specific heats of constant pressure on a weight basis.
2. Blend the specific heats at constant volume in the same manner.
3. The average K is then the ratio of the blended average specific heats.

Sizing for Liquid Service

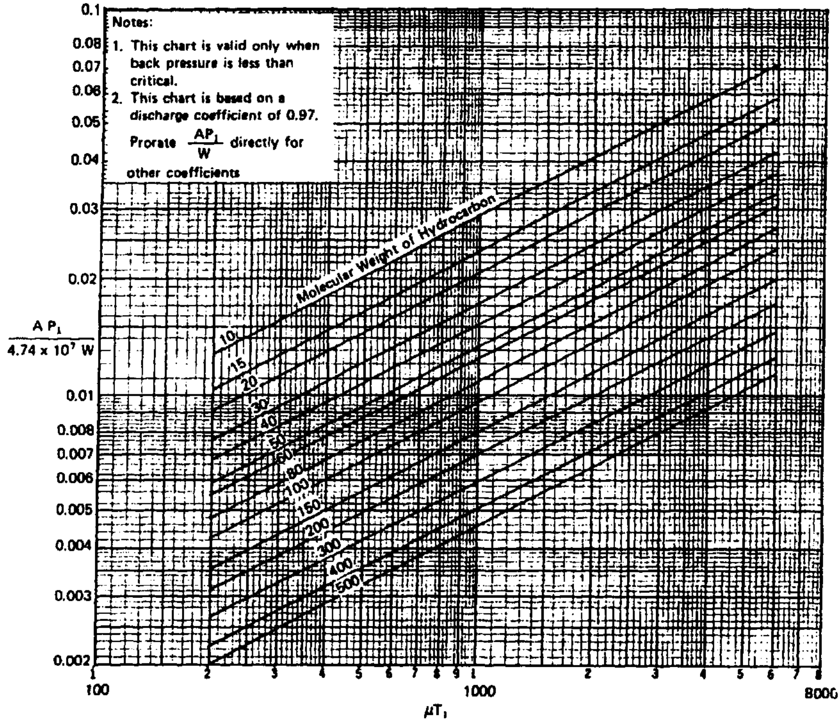
We shall first consider the case of non-flashing liquids. In this situation, there is no critical flow pressure limiting the flow of liquid through a PR valve orifice, as opposed to the case of vapor flow. The discharge rate is a function of the pressure drop across the valve and can be estimated by the following expression:

$$L = 101*10^{-3} A K_p K_u K_w \sqrt{\frac{P_d}{S}} \qquad (8)$$

where : L = Flow rate, dm^3/s

A = Effective orifice area, mm^2

K_p = Correction factor obtained from Figure 14 for overpressures less than



Notes:

1. This chart is valid only when back pressure is less than critical.
2. This chart is based on a discharge coefficient of 0.97. Prorate $\frac{AP_1}{W}$ directly for other coefficients

$$W = 1.10 \times 10^{-5}\, K_d\, K_b\, A\, P_1 \sqrt{\left(\frac{M}{\mu\, T_1}\right) K \left(\frac{2}{K+1}\right)^{\frac{K+1}{K-1}}}$$

A = Orifice area, mm^2

P_1 = Inlet press., kPaa (set press. + overpressure)

W = Rate of flow, kg/s

T_1 = Inlet temp., °K

M = Molecular weight

μ = Compressibility factor

$$\mu = \frac{PV}{RT_1}$$

K = Ratio of specific heats, $\frac{C_p}{C_v}$

K_d = Discharge coefficient = 0.97

K_b = Back pressure factor = 1.0

Figure 13. Safety valve orifice area required for hydrocarbon vapor release.

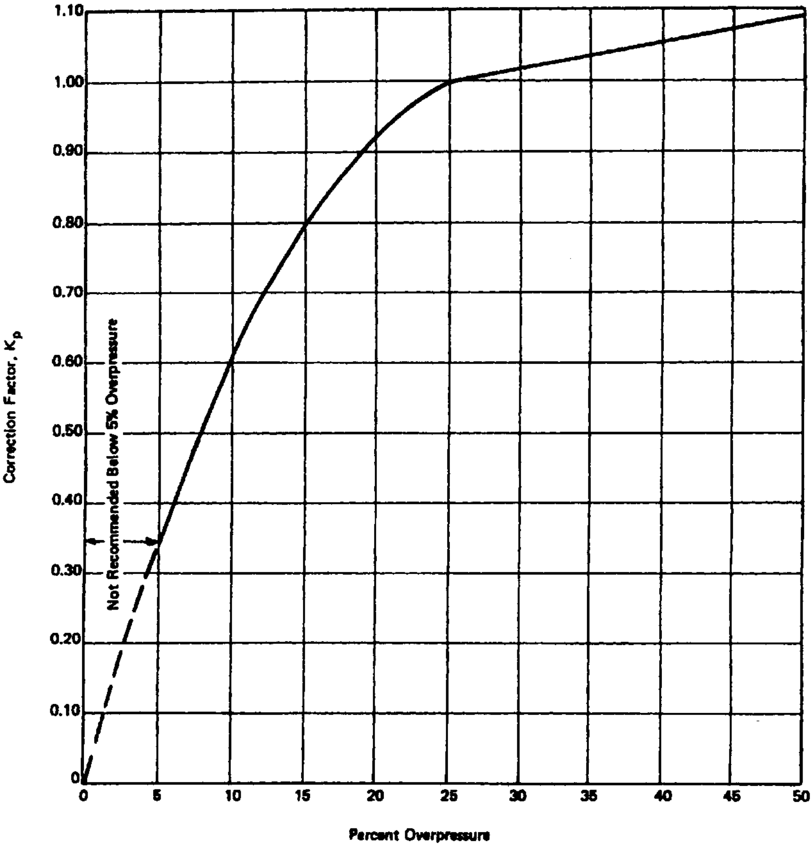


Figure 14. Capacity correction factors due to overpressure for relief and safety relief valves in liquid service. The curve shows that up to and including 25 percent overpressure, capacity if affected by the change in lift, the change in orifice discharge coefficient and the change in overpressure. Above 25 percent, the valve is at full lift and capacity is affected only by overpressure.

25% of set pressure. This is a necessary factor since liquid service PR valves generally require 25% overpressure to achieve full lift, and the sizing formula is based upon full lift. However, when the set pressure equals the design pressure, the maximum overpressure is limited under the ASME Code to 10% of set pressure (16% with multiple valves) for operating failure contingencies. K_p accounts for the reduced capacity resulting from reduced lift, reduced overpressure and changed orifice discharge coefficient.

K_u = Correction factor for viscous fluids, determined from Figures 15 and 16. A trial-and-error selection of orifice size is required in the determination of this factor.

The chart given in Figure 16 can be used in the following manner in order to size a relief valve for liquid service. First, determine the area required, A_0, without any viscosity correction (i.e., for $K_u=1$). Then select the next larger standard orifice size from manufacturer's literature. Determine the Reynolds number, based on the following definition:

$$R = 1.127x10^6 \, LS/ \mu\sqrt{A} \qquad (9)$$

where: L = Flow rate, dm^3/s
S = Specific gravity at flowing temperature versus water at 15°C
μ = Viscosity of fluid at flowing temperature, centipoises or mPa-s
A = Effective orifice area, mm^2(from the manufacturer's literature).

Obtain the viscosity correction factor, K_u , from the chart given in Figure 16. The corrected area can now be obtained from A_0/K_u. If this value is greater than A, repeat the calculation. If the required corrected area is only slightly larger a standard orifice size, one should consider using multiple smaller valves with staggered set pressures. This arrangement will minimize the tendency to chatter.

K_w = Correction factor for back pressure. Note that for conventional valves, K_w has a value of unity. A balanced bellows valve, however, has a heavier spring than a conventional valve for the same back pressure service, and the factor K_w accounts for the associated reduction in capacity. K_w is determined from manufacturer's charts, corresponding to the total back pressure (superimposed plus built-up). Significant variations exist between K_w values from different valve manufacturers. Figure 17 provides an average based on U.S. valve suppliers, which may be used as a rough design guide.

P_d = Pressure drop across the valve; i.e, the set pressure minus the total back pressure (superimposed plus built-up). It is important to note that although the actual flow is proportional to the square root of pressure drop at relieving

Figure III-5a

VISCOSITY AT OPERATING TEMPERATURE—SECONDS SAYBOLT UNIVERSAL (SSU)

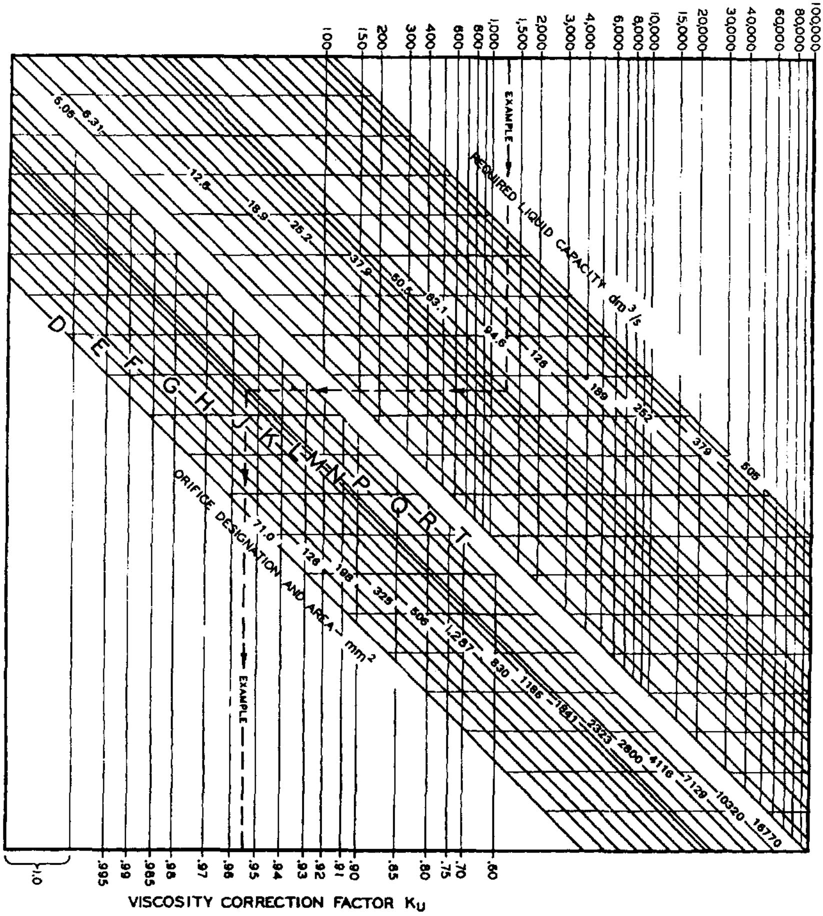


Figure 15. Viscosity at operating temperature - Second Saybolt Universal (ssu).

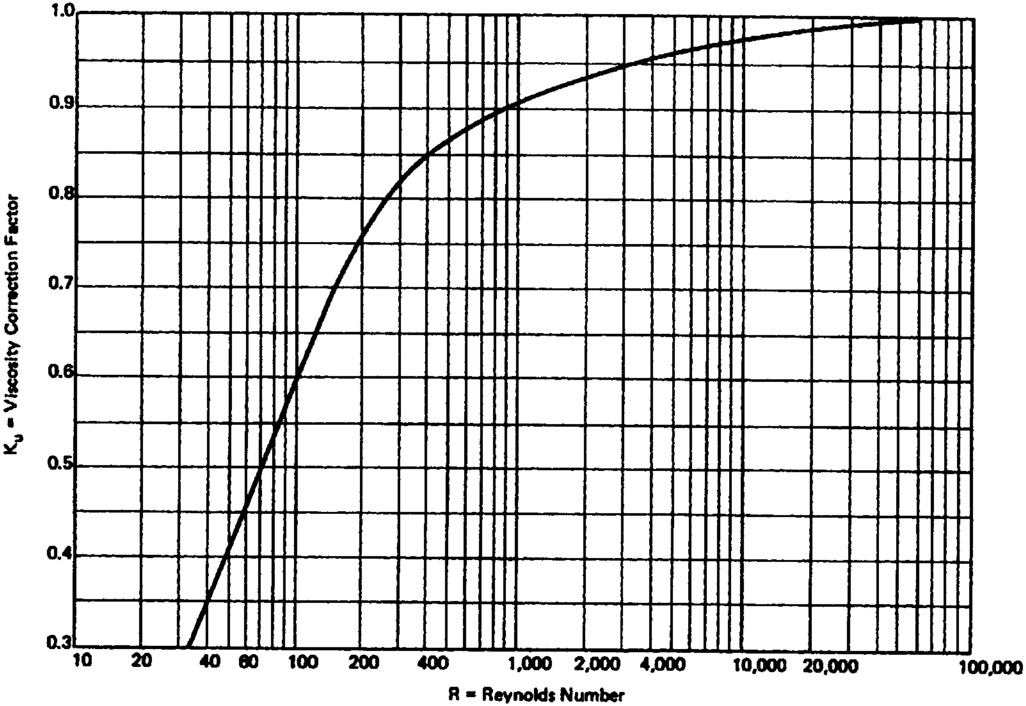


Figure 16. Viscosity correction chart.

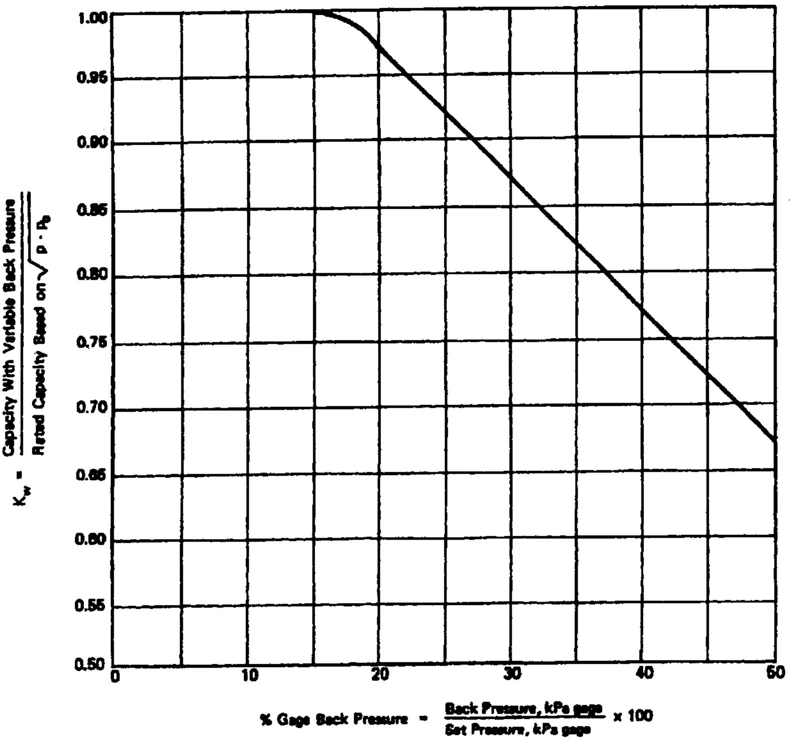


Figure 17. Variable or constant back pressure sizing factor , Kw for 25% overpressure on balanced bellows safety relief valves (liquids only). The curve represents compromise of the valves recommended by a number of relief valve manufacturers. This curve may be used when the make of the valve is not known. When the make is known, the manufacturer should be consulted for the correction factor.

conditions, this formula is based on pressure drop measured from set pressure.

S = Specific gravity at the inlet conditions, referenced to water at 15 °C.

Sizing for Flashing Liquids - PR valves handling fluids which are liquid at inlet conditions but which flash wholly or partially to vapor at discharge conditions can be sized using the following procedure:

1. By trial and error procedure, determine the amount of liquid which flashes by an isoenthalpic (constant enthalpy) expansion to the critical flow pressure (or actual pressure if greater than critical) for the flashed vapor.
2. Calculate individually the orifice area required to pass the flashed vapor component, using Equation (5a), (3b), (4), (5), or (6), as appropriate, according to service, type of valve and whether the back pressure is greater or less than the critical flow pressure.
3. Calculate individually the orifice area required to pass the unflashed liquid component, using Equation (8). The pressure drop term P_d should be made equal to the set pressure minus the total back pressure developed by the vapor portion at critical flow pressure, except when the critical flow pressure is less than the calculated total back pressure (superimposed plus built-up), considering the combined liquid and vapor flow. In the latter case, P_d should be made equal to set pressure minus the calculated total back pressure.
4. Add the areas calculated for the vapor and liquid components to obtain the total orifice area required. This may be somewhat conservative, since flashing does not take place instantaneously across the PR valve orifice.

Sizing for Mixed-Phase Vapor and Liquid Service - PR valve sizing for a mixture of liquid and vapor at inlet conditions may be calculated using the sum of the orifice areas required for each phase individually, in the same way as described above for flashing liquid service.

Sizing of Pilot-Operated Pressure Relief Valves

Sizing methods for pilot-operated pressure relief valves are in accordance with the accepted formulas described above, utilizing the appropriate discharge coefficients and effective orifice areas as recommended by the valve manufacturer. The following points should be noted:

- Typical discharge coefficients of pilot-operated valves are in the range 0.84 to 0.92. If the exact value is not known, use the lower coefficient.

- If the pilot valve exhausts to the atmosphere, a pilot-operated valve may be considered as a balanced valve.

- Pilot-operated valves have limitations in liquid service and reference should be made to the manufacturer for advice on sizing procedures.

Preparation of Design Specification for PR Valves

Table 4 provides an example of a typical Pressure Relief Valve specification sheet. The following notes indicate the basis for the times which are required in the Design Specification.

Number of Valves Required - Normally a manufacturer's standard PR valve with Driftice area equal to or larger than the calculated requirement is specified. In some cases, e.g., large relieving rates or to prevent chattering, two or more valves are necessary. Likewise, if there is an appreciable difference between the calculated orifice size and the available size, multiple PR valves are desirable to more nearly match the available area to the required orifice area. The column for "spares" indicates the requirement, if any, for spare PR valves installed on the equipment. Normally, this applies only in the case of refinery preference or local regulations, but is required in many European countries.

Temperature - Emergency operating temperature is the inlet fluid temperature under relieving conditions. This temperature is used for sizing the orifice for vapor service. Emergency temperature is not specified for liquid services, since temperature does not enter directly into the sizing calculation. The design temperature is specified equal to the design temperature of the vessel on which the valve is located, and this temperature is used for valve selection from manufacturers' temperature and pressure rating tables. The high relief temperature condition is considered as a short-time stress and is thus generally disregarded. The discharge temperature must, however, be included in the analysis of the collection system, particularly thermal expansion effects. Note that in some cases, the emergency temperature may exceed the design temperature, but the latter is still used as the basis for PR valve selection.

An example of such a case is a PR valve sized for fire on a vessel containing a high boiling hydrocarbon. The emergency temperature is the mid-boiling point of the hydrocarbon at relieving pressure and this may be considerably higher than the design temperature of the vessel and PR valve. The basis for this approach is that protection of equipment against fire exposure requires not only pressure relieving facilities to release the vapors generated from contained liquids, but also firefighting facilities to cool the equipment and prevent

Table 4. Design specification for pressure relief valves (sample design specification format).

Item Number										
Service										
Fluid										
Critical Condition										
Performance Data	Vapor	Liquid	Vapor	Liquid	Vapor	Liquid	Vapor	Liquid	Vapor	Liquid
Total Liquid Rate, m^3/s @ Cond. (1)	–		–		–		–		–	
Liquid Sp. Gr. @ Cond.	–		–		–		–		–	
Liquid Sp. Gr. @ 15°C	–		–		–		–		–	
Liquid Viscosity @ Cond., CS	–		–		–		–		–	
Total Vapor Rate, kg/hr		–		–		–		–		–
Vapor Molecular Weight		–		–		–		–		–
Compressibility Factor		–		–		–		–		–
Specific Heat Ratio Cp/Cv		–		–		–		–		–
Emergency Temperature, °C (2)										
Design Temperature, °C										
Set Pressure, kPa gage										
Allowable Overpressure, %										
Est. Superimposed Backpressure, kPa										
Est. Built-up Backpressure, kPa gage										
Est. Total Backpressure, kPa gage										
Valve Data										
Total Number Required										
Number In Service										
Number As Spares										
Differential Spring Pressure, kPa										
Req'd. Orifice Area Per Valve, mm^2										
Valve Type										
Valve Size Inlet/Orifice Letter/Outlet										

Notes: (1) For valves in flashing service, vapor and liquid rates shown are those actually used in sizing the valve.

(2) Emergency temperature is the inlet temperature when the valve is relieving.

overheating failures. Thus, the temperature under relief conditions would be no more severe than the fire exposure.

Set Pressure - The set pressure (the pressure at which the PR valve is designed to open) is specified in accordance with Code requirements, In most vessel applications, the set pressure of at least one PR valve is equal to the design pressure. However, this set pressure is adjusted (up or down) for any effect of static pressure and friction pressure drop that may apply when the valve is installed elsewhere than directly on the vessel. For example, if a PR valve is installed in a non-flowing line above a liquid-filled vessel, the PR valve set pressure would be reduced sufficiently to allow for the liquid static head between the vessel and the valve.

Back Pressure - In the case of a conventional PR valve, the "maximum back pressure" column should specify the maximum superimposed back pressure (kPa gage) under nonflowing conditions. If the discharge is routed to the atmosphere, or to a safety valve header where the pressure is essentially atmospheric under no flow, the maximum back pressure should be specified as zero.

For balanced bellows pressure relief valves, the maximum superimposed back pressure should be specified. (For balanced bellows valves discharging to a safety valve header which is at atmospheric pressure under non-flowing conditions, the maximum back pressure is zero.)

Differential Spring Pressure - For conventional valves, the differential spring pressure equals the set pressure minus the maximum superimposed back pressure under non-flowing conditions. For balanced valves, the opening pressure is not affected by back pressure, and the differential spring pressure equals the set pressure.

Allowable Overpressure and Critical Condition - Unless more restrictive local codes govern, the maximum accumulation for unfired pressure vessels should be specified in accordance with the ASME Code, i.e., 21% of design pressure maximum for fire and 10% maximum for all other failure contingencies. For PR valves with staggered set pressures, a valve set to open at 105% of design pressure may have an overpressure of only 10% for process contingencies or 16% for fire conditions when meeting ASME Code requirements for accumulation. Fired boilers, as well as steam superheating coils in process furnaces, are permitted 6% maximum accumulation by Section I of the ASME Code. (Some local codes may permit only 3% accumulation.)

The following unfired steam generators are considered as unfired pressure vessels, and maximum accumulation should be specified in accordance with theASME Code:

1. Evaporators and heat exchangers in which steam is generated.
2. Vessels, e.g., waste heat boilers, in which steam is generated incidental to the operation of a processing system containing a number of pressure vessels, such as are used in chemical and petroleum products manufacture. (Equipment which may fire a supplemental fuel should be considered as a fired pressure vessel.)

The "critical condition" in the specification sheet is entered as the contingency which determines the valve size, i.e., fire or operating failure.

PR Valve Type and Size - For convenience, valves are specified "Crosby or Farris valves equivalent accepted." The appropriate valve should be selected from the latest Crosby or Farris PR valve catalog. Sizes or ratings of PR valves that are not normally available are often obtainable by special order, and manufacturers should be consulted in such cases.

Effect of Temperature on Back Pressure Limits of PR Valves - Maximum back pressure limits are specified by the valve vendor. Usually the vendor's specification is given to a reference temperature (normally 38°C) for both conventional and bellows valves. These limits must be reduced for higher temperatures, as follows:

1. To the maximum pressure permitted for the outlet flange rating by ANSI. This applies to both conventional and bellows valves.
2. In the case of a balanced bellows pressure relief valve, to the maximum pressure permitted by considerations of bellows and bellows bonnet flange mechanical strength. This maximum pressure may be obtained by applying the following correction factor to the maximum back pressure listed for 38°C.

Temperature, °C	Multiply Maximum Back Pressure at 38°C By
200	1.00
300	0.85
400	0.70

These values can be interpolated and no correction is required below 200°C.

Remarks - The "remarks" column should include any relevant special factors such as the following:

1. The presence (and concentration, if known) of corrosive materials, other than the typical concentrations of sulfur compounds in petroleum hydrocarbon streams.

2. Requirements for any special valve features; e.g., non-standard inlet or outlet connections, valve lifting gear, etc.
3. Requirements for special materials of construction.
4. Basis for specifying any non-standard or special features.
5. Autorefrigeration if it affects materials of construction of the collection system.
6. Clarification of set pressure if different from design pressure of equipment.

PRESSURE RELIEF VALVE INSTALLATION

This section describes the requirements for the design and installation of pressure relief valve inlet and outlet piping manifolds and valving, including safety valve and flare headers.

Pressure Relief Valve Location

A pressure relief valve is normally installed at or near the top of the vessel which it is protecting. However, if local codes do not prohibit, it is permissible to mount the PR valve on the process piping connected to the vessel, provided that the relieving path from the vessel to the PR valve is free of restrictions and permissible pressure drop is not exceeded. On the same basis, it is also permissible to protect one or more vessels which are connected by piping by a single PR valve (or group of PR valves) mounted on one of the vessels or on the interconnecting piping.

Where a PR valve on a fractionator is used to protect a distillate drum with submerged inlet, the PR valve must be located on the tower itself, rather than on the overhead piping, or it must be discharged to a closed system. This is to preclude liquid release, should the PR valve discharge under fire conditions at the distillate drum.

Design of PR Valve Inlet Piping

When a PR valve is relieving at rated capacity, the total frictional pressure drop between a vessel and the inlet of the valve should be less than 3% of the set pressure (kPa). In this calculation, the effect on static pressure of fluid acceleration is ignored; rather, only friction loss is considered.

In the case of set pressures below 100 KPa gage (since pressure drop becomes proportionately larger) the maximum inlet pressure drop at design flow rate may be extended to 5% of set pressure (KPa, gage). The purpose of this requirement, which applies to both conventional and bellows type valves, is to prevent chattering.

Chattering caused by undersized inlet piping may sometimes be eliminated on pilot operated PR valves if the pilot valve pressure tapping is taken directly from the vessel being protected. However, it is recommended that the above inlet pressure drop limitations still be applied, to avoid the capacity reduction that would result from excessive inlet losses and to ensure freedom from chatter.

The entire inlet line must be at least the size of the PR valve inlet. When multiple safety valves are manifolded on the inlet side, the cross-sectional area of the manifold piping must be equal to or greater than the sum of all the inlet areas of valves open to the manifold.

Prevention of Plugging in PR Valve Inlets

1. Heat tracing of PR valve inlet piping should be provided where plugging by icing, deposition of wax or congealing of viscous liquids may occur at ambient temperatures.

2. Where inlet piping plugging may occur as the result of coke formation, solids deposition, etc., from the process stream, a continuous purge or blowback injection of clean fluid (e.g., steam) should be provided just below the PR valve. Flow is normally controlled by a restriction orifice sized to provide a flow of at least 1.5 m/s in in the inlet line. In the particular case of catalytic cracking reactor PR valves, experience has demonstrated that the inlet lines can be kept free of plugging by catalyst and coke if they are provided with an internal extension elbow within the reactor facing horizontally toward the vessel centerline. Any internal connection must be equal to or larger than the diameter of the pressure relief valve inlet. This is in addition to the steam purge.

Selection of Atmospheric or Closed Discharge for PR Valves

Discharge to a closed system is required for PR valves in the following categories:

1. PR valves handling materials which are liquid or partially liquid at the valve inlet. An exception to this is made for certain thermal expansion relief valves as described below.

2. PR valves normally in vapor service, but which under any single contingency may discharge flammable, corrosive or hazardous liquids.

3. PR valves located in the vapor space of partially liquid-filled vessels which could rapidly fill with liquid during a plant upset. This category includes towers, distillate drums, refrigeration flash drums, or any other vessels where the liquid holdup time above the high level alarm is less than 15 minutes, based on design liquid inlet rate and stoppage of liquid outflow. An exception, however, may be a closed loop refrigeration system where limited inventory precludes overfilling.

4. PR valves handling toxic or corrosive vapors which condense at ambient conditions; e.g., phenol.

5. PR valves in toxic vapor services where discharge to the atmosphere would result in the calculated concentration at the property line or at any working area (either at grade or an elevated platform) exceeding the Threshold Limit Value (TLV).

6. Releases of flammable vapors which, if discharged to the atmosphere, would in the event of inadvertent ignition result in radiant heat densities in excess of the permissible exposure level for personnel. This maximum level is defined as 19 Kw/m^2 at ground level.

An additional requirement applies to light gases, such as hydrogen, methane, ethane and ethylene, and to vapors above 315°C, since these materials have a higher probability of ignition than other typical hydrocarbons. For these streams, automatic instrumentation should be provided to effectively reduce the PR valve release so as not to exceed 9.5 Kw/m^2 heat density at ground level. This requirement would have no effect on the sizing of PR valves, but would only reduce the probability of reaching the maximum design release. However, should the instrumentation fail to operate properly, the maximum heat density should still not exceed the 19 Kw/m^2.

It should be noted that the above maximum radiant heat density criteria for application to inadvertently ignited atmospheric releases from pressure relief valves or vents are less restrictive than those used for flare design. This results from the fact that flares are continuously ignited, whereas ignition of a relieving PR valve is unlikely. In addition, the area surrounding a flare is open and offers no protection, while within a process unit access to shelter is available.

7. PR valves discharging vapors which do not fall into the above categories but which would be significant contributors to atmospheric pollution. Such releases should not normally be used to size the closed system but should be tied in up to the limit of its capacity. The order of preference for tying in is (1) malodorous vapors, (2) unsaturated hydrocarbons, (3) saturated hydrocarbons. If local requirements do not permit such atmospheric discharges, it will be

necessary to include these releases in setting the size of the closed system.

8. PR valves where atmospheric discharge is permissible, but connection to an adjacent closed header (provided that capacity is available) is less costly than an atmospheric discharge line to an acceptable location.

Criteria for Discharging PR Valves to the Atmosphere

Pressure relief valves may be discharged to atmosphere only, if the following conditions are satisfied:

1. The fluid handled must be all vapor at the valve inlet.
2. The valve must not fall into any of the categories listed under "Criteria for Discharging PR Valves to Closed Systems," above.
3. Local regulations regarding atmospheric releases are complied with.

Discharge Paths for Multiple Valves

Some equipment operating in two modes, such as reactors which are periodically regenerated, will require separate PR protection for each service. Special precautions are necessary where the PR valve in normal service discharges hydrocarbons and the valve for the regeneration cycle would discharge air. Where both valves discharge to atmosphere, a caution sign should be posted by the PR valves, and appropriate procedures clearly spelled out in the operating instructions. If the PR valve for hydrocarbon service discharges to a closed system, interlocks should be provided so that only one PR system can be in service at a time, and air from the regeneration cycle is kept out of the closed system. In addition, appropriate caution signs and proper instructions should be provided.

Application of Criteria for Routing of PR Valve Discharge

The application of the above criteria for routing PR valve discharges in a number of typical plant installations is described below:

1. Fractionating Tower - A pressure relief valve on a fractionating tower in hydrocarbon service is typically discharged to the atmosphere, provided that it meets the three criteria given immediately above. Although the PR valve location in the tower top or overhead line is normally exposed to vapor-phase material so that atmospheric discharge is acceptable, liquid may be entrained overhead under certain upset conditions, such as tower flooding or back flow

from the distillate drum. However, overpressure in conjunction with liquid entrainment is considered a double contingency which is not taken into account for design purposes, except for PR valves located in the vapor space of partially liquid-filled vessels, as described above.

Even when a tower overhead release is all vapor at the PR valve inlet the possibility of condensation of high molecular weight vapors after release may be another concern. It is assumed by some that extensive condensation will occur in the air if the lowest atmospheric temperature is below the condensing temperatures of the released hydrocarbon vapors. This does not necessarily occur, however. As vapors depressure across the PR valve, they are superheated which minimizes condensation tendencies in the rich zone near the point of release. Further downstream, the hydrocarbon vapors are cooled by mixing with air. This cooling and dilution with air reduces the hydrocarbon partial pressure and often results in the hydrocarbon dew point remaining below the actual vapor temperature, thus preventing condensation. It is believed however that few refinery PR valve releases condense, regardless of molecular weight.

The tendency is greatest, however, where pressures are close to atmospheric and "superheat" relative to atmosphere is least. Pipestill atmospheric towers and cat unit fractionators tend to fall in this category. Some operators consider that the likelihood is great that calculated condensation (dew) will coalesce to droplets which will gravitate (rain) when the partial pressure of condensibles at the dew point exceeds 1/3 atmosphere. With this factor and environmental protection in mind, some plants have diverted such releases into closed systems. Generally, however, this has not been of sufficient concern, and such releases have been treated as though they were all vapor.

2. Pumps and Furnaces - It is not always necessary for a PR valve which must discharge to a closed system to be tied into a flare header. For example, PR valves on furnaces frequently discharge to the vessel downstream of the furnace, and PR valves on pumps normally discharge to the pump suction or pump suction vessel.

When pressure variations on the pump suction could result in overpressure on the discharge side, the PR valve must discharge to some safe place other than the suction. As an illustration of a case where a PR valve cannot discharge back into the suction, assume a pump taking suction at 150 kPa gage from the bottom of a tower whose PR valve setting is 400 kPa gage. The normal pump discharge pressure is 800 kPa gage and the discharge line PR valve is set at 1000 kPa gage.

Normal procedure would be to specify the pump PR valve differential spring pressure equal to the required popping pressure minus the maximum suction pressure, or 1000 - 400 = 600 kPa. At normal operating suction pressure this valve would blow at 600 + 150 = 750 kPa. Since the normal pump discharge pressure is 800 kPa gage the PR valve would blow continuously. If the differential soring pressure were based on normal suction pressure, the differential spring pressure would be 1000 - 150 = 850 kPa. Under conditions of maximum suction pressure, this valve would not blow until 850 + 400 = 1250 kPa pump discharge pressure had been reached. This would exceed the downstream safe working pressure.

As an alternative, a balanced bellows valve could be used since it is not affected by a variable superimposed back pressure.

3. Thermal Expansion Relief Valves - Routing of thermal expansion relief valve discharges was covered earlier.

4. Waste Heat Boilers - A PR valve installed on the generating exchanger shell of a waste heat boiler (as opposed to installation on the steam drum) will in most designs discharge a mixture of boiling water and steam. This mixture cannot be safely discharged to the atmosphere, either at grade or at an elevated location, and a means of separating water from steam, such as a blowdown drum, must be provided to enable the two phases to be safely discharged.

5. PR valves installed on heat exchangers and condensers for protection against a split tube should be discharged into a closed system if liquid can be discharged.

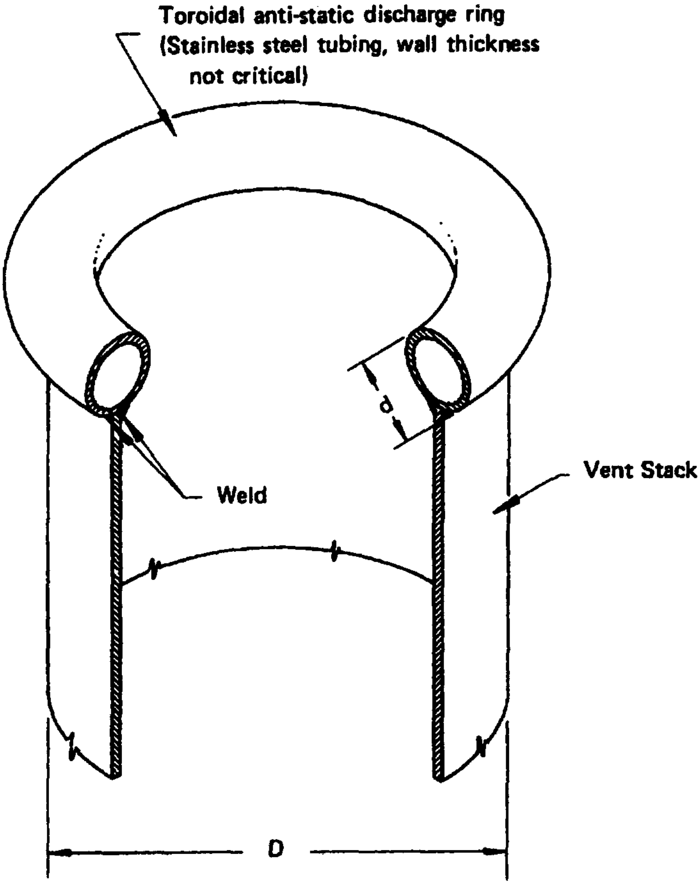
Ignition of PR Valve Releases

In applying the requirement for snuffing steam connections, 315°C should be used as an autoignition temperature criterion for typical hydrocarbon streams.

In addition to these requirements for inclusion of a snuffing steam connection, this feature should also be provided for all PR valves in hydrogen and methane service which discharge to the atmosphere. PR valve risers in hydrogen and methane service should also be provided with a toroidal ring (illustrated in Figure 18) to reduce the probability of ignition by electrostatic discharges.

Discharge Riser Elevation

When a PR valve in flammable vapor service is mounted at the top of a vessel, the riser elevation requirements should be observed and include the following:



Stack diameter D, mm	Required Toroidal Ring diameter d, mm
Less than 200	13
200 to 300	20
Over 300	25

Figure 18. Toroidal ring for hydrogen and methane vents.

1. The riser tip must be at least 3 m above the floor of the top platform.
2. The riser tip must be at least as high as any part of the vessel or its overhead piping, to prevent discharge impingement.
3. The riser should discharge vertically upward.
4. Location should also consider radiant heat levels, in case the PR valve release should become ignited.

Piping Restrictions - No restrictions such as check valves, flame arresters, orifice plates, etc., are permitted in the outlet piping.

Outlet Pipe Sizing - Outlet piping is sized according to minimum diameter and exit velocity considerations, as follows:

1. The discharge line diameter must be at least equal to the diameter of the PR valve outlet flange.
2. The PR valve riser should be sized for a maximum exit velocity, under the design contingency release, of 75% of sonic velocity. This is to limit noise problems and avoid choked flow. If it is necessary to install a section of enlarged diameter riser piping to limit the maximum exit velocity, its length should equal at least ten diameters of the enlarged piping.
3. PR valve risers in flammable service should also be sized such that exit velocities are at least 30 m/s under all foreseeable contingencies (except fire) which would cause the valve to release. On the basis of experimental work and plant experience, this minimum velocity, in conjunction with the riser elevation requirements, has been shown to ensure effective dispersion. Entrainment of air and dilution result in a limited flammable zone, with a negligible probability of this zone reaching any equipment which could constitute an ignition source.

To meet the minimum exit velocity criteria, consideration must be given to overpressure contingencies which result in relieving rates less than those of the design contingency. In considering exit velocity it is important to assure that built-up back pressure limits are not exceeded. In some cases this factor, in conjunction with considerations of PR valve chattering at low release rates, will justify the installation of two PR valves with staggered set points.

Design Temperature of Outlet Piping - The design temperature of outlet piping from PR valves discharging to the atmosphere is normally ambient. However, autorefrigeration and need for brittle-fracture-resistant materials or thermal expansion forces should be examined if the release pipe is unusually long.

Multiple PR Valve Discharges to a Single Atmosphere Riser - Where two or more PR valves are manifolded into a single riser discharging to the atmosphere, the following additional requirements apply:

1. Isolation Valves - Isolation valves in the individual PR valve outlet lines should be provided to permit safe removal of one PR valve for maintenance during plant operation.

2. Back Pressure - The combined atmosphere discharge system must be designed to comply with the superimposed back pressure limitations.

3. Fatigue Failure - The combined discharge header system should be designed in accordance with the same considerations of potential piping fatigue failure as described below for closed discharge systems.

4. Exit Velocity - Exit velocity from the combined riser should comply with the same maximum and minimum velocity requirements defined for single PR valve installations. The maximum velocity should be based on the largest single contingency release from the group of PR valves which are manifolded together. The minimum velocity should be based upon the *smallest single contingency* release from one PR valve in the group. This may require more than one riser, if multiple PR valves are installed.

Design of Closed Systems for Pressure Relief Valves

Closed systems for PR valve releases are designed in accordance with the following:

Routing of Closed Systems

1. Conventional Flare System - The majority of pressure relief valve discharges which must be routed to a closed system are manifolded into a conventional blowdown drum and flare system. The blowdown drum serves to separate liquid and vapor so that the vapor portion can be safely flared, and the separated liquid is pumped to appropriate disposal facilities. The blowdown drum may be of the condensible or noncondensible type, according to the characteristics of the streams entering the system. Selection criteria, as well as the design basis for each type of blowdown drum, are detailed later in this volume. The design of flares, including seal drums and other means of flashback protection, is described later.

In addition to handling PR valve releases, the flare header is also used to route certain other emergency releases to the blowdown drum. These include drainage from fuel gas, compressor and absorber knockout drums.

2. To Atmosphere Via Condensible Blowdown Drum - Releases which can be totally condensed may be routed to a condensible blowdown drum, which may be vented to the atmosphere.

3. Segregated H_2S Flaring System - Continuous releases of hydrogen sulfide are normally routed to a segregated H_2S flare system to limit the extent of fouling and plugging problems. PR valves in concentrated H_2S service are not considered "continuous releases," and should not be tied into a segregated system. This is to avoid the possibility of a plugged or fouled release system for the PR valves.

H_2S flare headers should be constructed for each of isolation, washing out and dismantling for cleaning. The need for periodic cleaning of the H_2S flare header must be recognized, and alternative routing for H_2S releases must be provided for such occasions if a shutdown of the H_2S sources cannot be tolerated.

4. Other Segregated Closed Systems for Special Services - Special closed systems are also provided for PR valve releases in certain services where operating problems or hazards would result from discharge into the regular flare header. Such services include severely toxic, corrosive, polluting or high-cost materials and the following are examples of the special facilities required:

a. Segregated header and condensible blowdown tank for PR valves discharging phenol, methyl ethyl ketone (MEK) or dimethyl formamide (DMF) in processes which use these materials as solvents.

b. Segregated header and blowdown drum system (which vents into a regular flare header) for PR valves discharging corrosive material such as acid or acid extract in processes such as alkylation or isobutylene extraction.

c. Segregated sections of flare header to prevent the mixing of vapors which would react chemically causing the deposition of solids. For example, ammonia must be segregated from H_2S or CO_2 to avoid header plugging by ammonium sulfide or ammonium carbonate. In some situations, heating of the flare line may prevent reaction and formation of solid deposits.

Sizing of Flare System and Other Closed Release Systems

The factors affecting the sizing of a closed release system to the flare are described below. The same principles apply to the sizing of other closed release systems; e.g., PR valve releases which are manifolded into an atmospheric vent.

1. Consideration of All Releases into the System - All releases tied into the closed system must be considered. In addition to PR valve discharges, these may include fuel gas compressor and absorber knockout drum drainage, vapors vented from water disengaging drums, feed diversion streams, closed drainage from equipment, vapor blowdowns and liquid pulldowns.

2. Maximum Flow to be Handled - The flare system must be sized to handle

the largest total flow from any of the sources connected into it which results from any one single contingency. Since the flare system consists of a number of collecting headers, laterals, blowdown and disengaging drums, etc., this principle must apply to all parts of the system, i.e., each part of the system must be capable of handling the maximum single contingency flow which can occur in that section of the header. (Any continuous load, such as excess gas flaring, is additive to the largest contingency.)

3. Pressure Available and Component Pressure Drops through the Flare System - Pressure drop through the flare system is made up of component pressure drops through the following:

a. PR valve discharge laterals
b. Safety valve headers
c. Blowdown drum
d. Flare header
e. Flare seal (determined by dipleg submergence)
f. Flare stack
g. Flare tip (typically 7 kPa for proprietary smokeless tips).

With the flare tip and flare seal pressure drop and flare elevation fixed, the flare stack, headers and laterials are sized for the largest release, while not exceeding the maximum allowable operating pressure on the associated blowdown drums and water disengaging drums. These maximum allowable operating pressures are in turn determined by:

h. Maximum allowable back pressure on PR valves, which is a function of set pressure, type of valve and largest contingency.

i. Maximum allowable back pressure on vapor diversion streams which will permit the design rate to be released from equipment.

j. Maximum allowable back pressure on water streams from equipment which will permit the design flow to be released to the water disengaging drum.

Design of Closed Release System

1. Routing of Flare Header through Process Areas - Flare headers in process areas should be routed to avoid locations of particularly high fire risk, such as over pumps, near furnaces, etc. The headers and subheaders should also be laid out and provided with isolating CSO valves and spectacle blinds, unless prohibited by local codes, such that it is not necessary for flare lines to remain in service in units which are shut down separately. Blowdown and water disengaging drums should be spaced from process areas.

2. PR Valve Outlet Piping - The outlet piping from a PR valve must be at least equal in size to the valve outlet flange.

3. Liquid Drainage from Closed Relief System - Accumulation of liquid in closed relief systems can impose appreciable back pressure and reduce relieving capacity. The following design features must be included to avoid these problems:

a. PR valves should be mounted above the header, so that the outlet piping drains into the header. In exceptional cases, location below the header is permitted, with special drainage for the outlet piping.

b. Closed headers should slope continuously down to the blowdown drum from all points where laterals from PR valves or other releases enter it. Flare headers should slope down continuously from the flare seal drum back to the blowdown drum. The required slope is 0.2%.

c. There must be no low points or other liquid traps where liquid can accumulate in closed release systems.

d. Heat tracing of PR valve outlet piping and headers should be provided where plugging by deposition of wax, ice or congealing of viscous liquids may occur.

4. Thermal Expansion in Flare Header - Sliding-type expansion joints may be used in flare headers as an alternative to piping expansion loops, if required to achieve a reduction in pressure drop or where expansion bends may result in liquid surging, subject to the following conditions:

a. They are permissible only in low fire risk areas, such as offsite areas, at least 4.5 m from roadways and 15 m from continuous ignition sources such as furnaces.

b. They must not be used in segregated H_2S flare headers or similar services, because of the toxicity of the fluid, in case leakage should occur.

c. The liquid or vapor is non-corrosive to the sliding surfaces.

d. The system in which they are installed operates below 545 kPa gage (specify 345 kPa gage minimum design pressure).

e. Maximum temperature limit is 400°C and a minimum of 10°C for intermittent service, since ice formation could cause the joint to bind.

If a sliding joint is specified, the design specification should also include the following notes:

f. Provide guides and anchors to eliminate piping moments and absorb end thrust on joints.

g. Sliding cylindrical parts of the joint shall be 18-8 stainless steel.

h. Joint shall be internally guided to maintain axial alignment.

i. Joint shall have a minimum of 6 packing rings with lantern ring and provisions for non-combustible lubricant injection. The packing material shall be braided asbestos suitable to 400°C and shall be at least 13 mm square. Teflon impregnated packing is not acceptable.

j. Joint shall be protected with tie rod limit stops to prevent overextension or compression.

k. Joint marking shall include design pressure and temperature, temperature range, ambient setting, maximum movement range for compression and extension.

5. Design Temperature of Closed Release System - The design temperature of all piping, headers, blowdown drums, etc., in a closed release system must consider the most extreme actual release conditions associated with any single contingency. The application of this design basis includes the following interpretations:

a. *Consideration must* be *given to* low *temperatures* which may result from autorefrigeration or expansion as process fluids are discharged through PR valves.

b. Credit may be taken for expansion temperature drop in the case of hot releases, and also for atmospheric cooling of the header if this can be realistically calculated.

c. The extremes of discharge temperature used for design of pressure relief systems are based upon operating failure contingencies. Thus, the design maximum temperature for the purpose of materials selection for PR valves and inlet and outlet piping is taken as the design maximum temperature of the protected vessel, with credit for temperature drop as described above. However, in external fire exposure contingencies, these temperatures may be exceeded by the process fluid temperature (e.g., in the case of high-boiling liquids) or by equipment surface temperature (as a result of flame exposure); but, with certain exceptions, these fire contingency temperatures are not used for pressure relief system design. These exceptions are:

(1) The fluid boiling point at relieving pressure is used for sizing a PR valve orifice for the contingency of fire.

(2) The coldest and hottest relieving temperature, whether resulting from ambient conditions, operational failure or fire contingency, is used for the purpose of designing for thermal expansion in the complete system of piping and headers. Credit may be taken for temperature rise or drop in the system, as described above.

d. Design of closed relief system for low temperatures should take the following factors into consideration:

(1) Materials of construction must be adequate to withstand the lowest

possible temperature which may occur in service.

(2) Thermal contraction of piping must be allowed for, based on the lowest possible service temperature.

(3) When applying low temperature requirements, one should consider safety valve and flare headers to be subject to "shock chilling" if they can be exposed to cold liquids released into the system. This includes flare headers from blowdown drums into which cold liquids are discharged. Where laterals of different piping material are combined, the material of the lower-temperature header is continued for the rest of the combined line, and is also extended back into the other lines for 6 m.

(4) An in-line heater on the vapor outlet of a blowdown drum is a permissible means of protecting the downstream header and flare against low temperatures that could result from cold vapor releases, or from the weathering of cold liquids in the drum. Normally, the in-line heater consists of a section of steam jacketing on the header, with a continuous steam supply and an independent low temperature alarm. The design must also provide positive means for condensate removal, so as to avoid loss of heat transfer and possible ice formation. Acceptable designs include a barometric seal leg (where steam pressure is low enough) or a steam drum with 15 minutes holdup between a normal high level and an emergency high level with independent dump system. It is important that no condensate appears in the in-line heater at any time. However, flare line heaters are not permissible in lines subject to liquid shock chilling where the possibility of heater failure could result in brittle fracture.

(5) Flare system designs must also include means of preventing freezing of seal water in the flare seal drum, if entering vapors may be below 0°C.

(6) Isolation Valves for Pressure Relief Systems - Block valves for maintenance isolation purposes are permissible in pressure relieving systems, provided that they are car sealed open and comply with the requirements of CSO valves. The particular locations where such CSO valves are permitted are:

a. PR valve inlets, where isolation of the PR valve for onstream maintenance is required, subject to compliance with local codes.

b. PR valve outlets which are manifolded to a closed system or combined atmospheric vent, where isolation of the PR valve for onstream maintenance is required, subject to compliance with local codes.

c. A flare header at the battery limit of a unit that shuts down independently of other units tied into the same header.

d. Flare inlet lines, when a flare system is tied into more than one flare and it is required to isolate a flare for maintenance.

(7) Acoustically Induced Vibration Problems in Flare Header Systems - The current trend toward larger units and equipment increases the potential for complex vibration problems. In particular, gas piping systems with high capacity pressure reducing valves have experienced problems of fatigue failure where excessive turbulence and high acoustic energy existed. The turbulent forces excite

certain complex modes of vibration in downstream piping components. These vibrations can in turn result in stresses exceeding the endurance limit for the materials and thus, fatigue failure. Pressure relief valves may have the capability of generating sufficient acoustic energy to cause fatigue failures in downstream discharge laterals and/or flare header piping.

Potential vibration problems of this type should be considered early in the design stage of the flare header system. The following screening criteria have been developed to assist the designer in recognizing services with potential vibration problems requiring further detail evaluation:

(1) Downstream line size 400 mm and greater: mass flow rate greater than 91,000 Kg/hr or pressure ratio greater than 3.

(2) Downstream line size 200-350 mum downstream line velocity greater than 50% sonic and pressure ratio greater than 5.

(3) Downstream line size less than 200 mm swaged up or "Teed" to 200 mm or larger line: downstream line velocity greater than 50% sonic and pressure ratio greater than 5.

The above criteria are a guide for detecting potential problems with gas letdown systems and apply for the first 90 m of piping downstream of the pressure reducer under concern. Systems with only liquid flow are not considered potential problems and need not be investigated. For systems with two phase flow, use the conservative assumption of the total mass flow rate as gas. Any system exceeding these criteria should be further evaluated.

A distinction must be made regarding the length of service of the pressure reducing systems. Fatigue failure of any mechanical system depends on time, i.e., the number of cycles to failure. Therefore, the treatment required for a continuous service may not be justified for a short term service. A System in short term service is defined as one which operates a total of 12 hours or less during the life of the plant. Pressure relief valves typically meet this limit. Systems in short term service exceeding the screening criteria indicated above should be evaluated.

GENERAL DESIGN NOTES AND CONSIDERATIONS

The following design features should be applied to the discharge piping and header for approximately 90 m downstream from the PR valve in question.

1. Use pipe with a minimum wall thickness of 13 mm to increase flexural stiffness.

2. Use completely welded full wrap-around reinforcement pads at branch connections with pad thickness equal to header wall thickness.

3. Use wrap-around reinforcement at welded support shoes and anchors. Alternatively, all welding of these fittings to the pipe wall may be eliminated by the use of bolted shoes and anchors.

4. Eliminate all vents, drains and small diameter connections.

Continuous Service - Pressure reducing valves which will operate more than 12 hours during the life of the plant should be considered to be in continuous service. Such systems which exceed the screening criteria given earlier should be further evaluated. Systems in continuous service believed to be fatigue prone require more positive action to reduce the acoustically induced vibrations because of the greater potential for fatigue failures in these systems. Treatment alternatives for these services typically require measures to reduce the acoustic energy generated at the source.

Maximum Line Velocity

Sonic conditions at piping discontinuities such as at branch connections, reducers, etc, can also result in unacceptable acoustically induced vibrations. Since these piping components are not designed to be controlled pressure reducing stations, they have proven to be more susceptible to fatigue failure. Maximum vapor or mixed phase flow velocities in piping downstream of PR valves in continuous service should not exceed 50 percent of sonic. However, for short term services such as a relief device that discharges into a flare header system, maximum line velocities can be as high as 75 percent of sonic.

Determining Vapor Loads from Liquid-Containing Vessels Exposed to Fire

The following is a procedure for calculating the required relieving rates for a vessel that is exposed to an external fire.

Step I - Amount of Heat Absorbed

The amount of heat absorbed by a vessel exposed to an open fire is markedly affected by the size and character of the installation and by the environment. These conditions are evaluated by the following equation, in which the effect of size on the heat input is shown by the exponent of A, the vessel wetted area, and the effect of other conditions is included in a factor F:

$$Q = 43.1\ F(A)^{0.82} \tag{10}$$

where:

Q = Total heat absorption (input) to the wetted surface, kW

A = Total wetted surface, m^2

F = Environmental factor

The chart given in Figure 19 can be used to solve for Q.

The value of the environmental factor, F, depends on the thermal conductance of the insulation on the vessel; other environmental effects are ignored. The chart in Figure 19 shows approximate values for F- for a bare vessel ($F=1.0$) and with insulation having a conductance of 6 to 23 W/m^2 - °C.

Thermal conductance is the thermal conductivity divided by insulation thickness. The thermal conductivity to use here is at the mean temperature between 900°C and the process temperature expected at relieving conditions. A conservative mean temperature of 540°C is suggested. For insulation types, a thermal conductivity of 0.144 W/m °C may be used. For gunite and concrete, a value of 1.15 W/m °C may be used. For jacketed vessels, a conservative F value of 1.075 may be used. Generally, no credit should be taken for water spray or deluge systems in sizing PR systems. Supplying enough water to absorb most of the radiant heat becomes impractical for most installations. Emphasis should be placed on providing enough water to the outside of a vessel exposed to fire, to keep metal temperature below a point where failure might occur. Freezing weather, high winds, clogged systems, undependable water supply, and vessel surface conditions are among the factors which may prevent adequate water coverage, so no reduction in environmental factor is recommended.

Note that economics will in some cases favor the provision of additional insulation (beyond that required for heat conservation), in order to reduce the required capacity of the pressure relieving system, when it is sized by fire considerations.

Note that the wetted surface A used to calculate heat absorption for a practical fire situation is normally taken to be the total wetted surface within 7.5 m of grade. "Grade" usually refers to ground level, but any other level at which a major fire could be sustained, such as a platform, should also be considered. In the case of vessels containing a variable level of liquid, the average level is considered. Specific interpretations of A to be used for various vessels are as

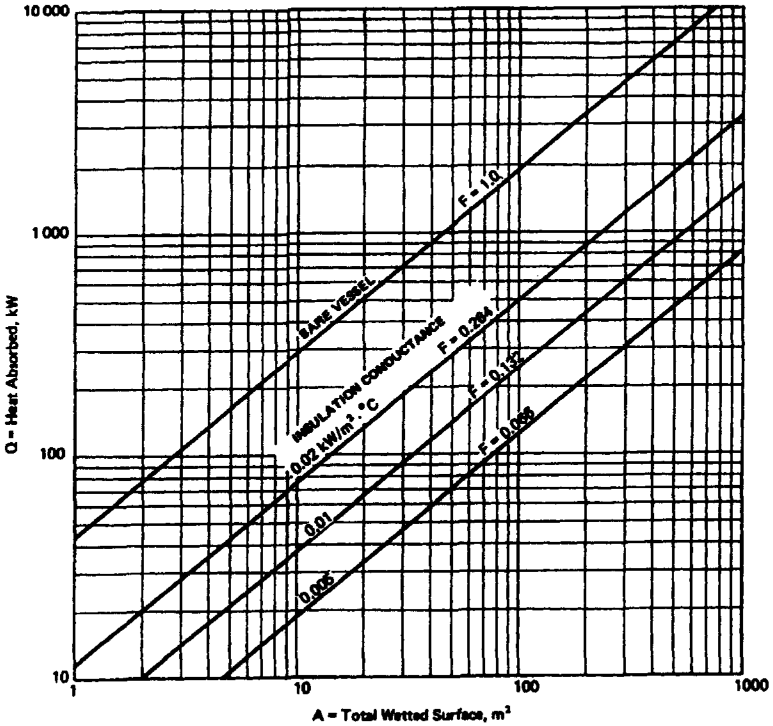


Figure 19. Chart for determining the amount of heat absorbed by a vessel exposed to an open fire.

follows:

1. Horizontal Drums

a. Less than Half Full (to normal HLL) - The wetted vessel surface up to normal high liquid level is used, regardless of distance above grade.

b. More than Half Full (to normal HLL) - The total wetted vessel surface within 7.5 m of grade, or up to the vessel equator, whichever is greater, is used.

2. Vertical Drums - The wetted vessel surface within 7.5 m of grade, based on normal high liquid level, is used. If the entire vessel is 7.5 m or more above grade, then only the surface of the bottom head need be included.

3. Fractionators and Other Towers - An equivalent "tower dumped" level is calculated by adding the liquid holdup on the trays to the liquid at normal tower bottom (high liquid level). The surface that is wetted by this equivalent level and which is within 7.5 m of grade is used.

4. Storage Spheres and Spheroids - The total exposed surface within 7.5 m of grade, or up to the elevation of the maximum horizontal diameter, whichever is greater, is used.

5. Heat Exchangers, Air Fins and Piping - These items of equipment are noramlly not considered for establishing the wetted surface of exposed vessels in a fire risk area. However, there may be special situations, such as congestion and substandard spacing, or an unusually large piece of equipment, such as a reboiler with liquid holdup of about 4 m3 or more, which warrant their inclusion.

Step 2 - Vapor Release Rate

All heat absorption from fire exposure is considered as latent heat and no credit is taken for the sensible heat capacity of the fluid in the vessel. The vapor release rate W is calculated from:

$$W = Q/\lambda \tag{11}$$

where: W = Vapor rate, kg/s

Q = Heat absorption calculated previously, Watts

λ = Latent heat of vaporization of the liquid in the vessel, Btu/lb, evaluated at the pressure existing at the PR valve inlet under full discharge conditions.

The vapor to be relieved is the vapor that is in equilibrium with the liquid under conditions existing when the valve is relieving at maximum capacity. It should not be overlooked that the vapor and liquid composition may change as vapors are released from the system. Therefore, temperatures and the latent heat value could change. The latent heat and molecular weight values to be used in calculating the rate of vaporization should be those pertaining to the conditions capable of generating the maximum vapor rate. Duration of a fire does not normally enter into this analysis.

Estimation of latent heat becomes unreliable at temperatures approaching or exceeding the critical temperature. The following values can be used as rough approximations:

- For liquid vaporization conditions approaching the critical point, the minimum latent heat that should be used is 47 kJ/kg.

- For liquid vaporization conditions at or above the critical point, the rate of vapor discharge depends on the rate at which the fluid will expand. For such situations a latent heat of 116 kJ/kg can be used.

9

DESIGN OF DISPOSAL SYSTEMS

INTRODUCTION

This chapter covers the design of facilities to handle equipment drainage and contaminated aqueous effluents that are sent for appropriate disposal; blowdown drum systems to receive closed safety valve discharges, emergency vapor blowdowns, etc.; and facilities for process stream diversion and slop storage. Also covered are criteria for selecting the appropriate method of disposal. Design of flares is covered in a subsequent chapter.

The purpose of the facilities described in this chapter is to provide for safe handling of various drainage materials and emergency streams, so that they may be safely routed to the sewer, tankage, flare, or other appropriate destination. Drainage systems specified herein ensure that flammable or toxic materials may be disposed of without hazard of fire or injury when equipment is taken out of service. Also described are systems to handle process water drawoffs, cooling water, and other aqueous effluent streams which may be contaminated with hydrocarbons, and which could otherwise create hazardous conditions if they were discharged directly to the sewer.

Safety valve releases are routed to blowdown drums when the presence of liquid, toxic properties or other factors would make discharge to the atmosphere hazardous. Product and intermediate process streams may need to be diverted to alternative disposal if they are off-specification (e.g., during startup) or in the event of emergency shutdown of downstream equipment.

EQUIPMENT VENTING AND DRAINAGE

This section covers low point drains and high point vents, as well as connections specifically provided for equipment drainage and venting at shutdown or when taken out of service. Also covered are vents and drains for instruments, gage glasses, sample points, etc. Regularly used drain connections and sample points in light ends service must be double valved. The requirements for discharge of dangerous materials to closed drain systems are supplemented by the following paragraphs.

Disposal of Drainage of Process Equipment Contents - When items of onsite process equipment are taken out of service, either individually during plant operation or for general turnaround, means of draining and safe disposal of the residual liquid hydrocarbon contents must be provided. The following are important considerations and terminology in the design :

1. The term *vessels* in refinery operations is generally meant to include towers, drums, and miscellaneous onsite equipment such as filters, strainers, separators, etc. Heat exchangers are normally considered separately.

2. The term *inventory* refers to liquid hydrocarbon contents at the top of the working level range. Tray holdup is included, but piping contents are disregarded.

3. Open connections to an oily water sewer catch basin should be at least 15 m from any continuous ignition source. Normal practice is to flush with water at the point of discharge. However, this disposal route should not be used for drainage of equipment containing high pour point materials that would solidify in the sewer, unless light flushing oil connections are provided for displacing such materials from the equipment.

4. Controlled release to the atmosphere should be at least 15 m from any continuous ignition source.

5. Connections to the closed drain header may be justified for large volume light ends pumps.

6. Refinery preference may exceed these requirements. In such cases, additional 25 mm connections from equipment to the closed drain header may be installed (20 mm connections are generally considered adequate for pumps).

7. Heat exchangers which are valved for on-stream maintenance may be considered as vessels, according to inventory and contents. Shell sides and tube sides should be treated separately. However, if the liquid contents of an exchanger can be gravitated into connected equipment through the process piping before closing all the isolating valves, then the requirement for drainage to the

closed drain header may be deleted.

Heat exchangers which are not valved for onstream maintenance require only a means of drainage for a unit shutdown situation. If the liquid contents of a heat exchanger cannot be gravitated or displaced (as part of the shutdown procedure) into a connected vessel which is provided with an appropriate means of drainage, then the exchanger should be considered as a vessel and provided with drainage facilities according to inventory and contents.

8. The methods of equipment drainage described above are considered to provide safe disposal of liquid hydrocarbon contents for application to most process unit designs. Draining to atmosphere or sewer, when permitted by the above guidelines, is subject to good operating judgement in consideration of prevailing conditions (wind direction, adjacent ignition sources, need for protective clothing, etc). Pollution control considerations may require the more extensive use of closed drain header connections.

Table 1 provides a summary of the safe practices applied to the disposal of liquid hydrocarbon contents from various sources.

Table 1. Guidelines for Disposal of Liquid Hydrocarbons

Source	**Light Ends**	**Heavier than light ends, at temperatures above flash point**	**Heavier than light ends, at temperatures below flash point**
Vessels with liquid inventories > 0.1 m^3	Closed drain header	Closed drain header	Sewer
Vessels with liquid inventories ≤ 0.1 m^3	Atmosphere	Sewer	Sewer
Pumps	Atmosphere	Sewer	Sewer
Compressor casing, cylinder and knockout bottle drains	Closed drain header	--	--

Closed Drain Header Systems for Flammable Liquids

Closed liquid drain headers are provided, according to the criteria described in the preceding paragraphs, for the safe drainage of light ends and light stocks which would otherwise cause hazardous releases of hydrocarbons to the atmosphere or to the sewer. The connections are relatively small, intended for preparation of equipment for maintenance. The design of closed drain header systems are usually based on the following guidelines:

1. Connections to equipment are typically 50 mm and 80 mm for process vessels and exchangers, according to the size of the equipment. Each connection includes an accessible block valve. Double block valves are provided if required. A check valve should be included if overpressure or other hazard could result from reverse flow during simultaneous drainage from more than one vessel. Individual connections from the equipment are made into the top of the drain header.

2. The header is normally a 80 mm diameter pipe (50 mm may be adequate for small units) and is routed via an overhead pipe rack (which is generally sloped) to a non-condensible blowdown drum.

3. The header is rated the same as the highest pressure rated equipment connected to it; or it is fitted with a safety valve if designed for a lower pressure rating. Sections of the header, separated by check valves, may be designed for different pressure ratings, but safety valve protection is still required for the lower-rated sections, unless the header cannot be overpressured to more than 1.5 times the design pressure.

4. The header must be designed for the extremes of high and low temperatures and corrosive conditions which can arise from the discharge of process streams into it. The flashing and autorefrigeration of light ends liquids may require special materials (e.g., killed carbon steel for C_3's, and alloy steels for C_2's and lighter). It is usually economical to minimize the use of special materials by segregating such streams into separate sub-headers of the closed drain system. These sub-headers may be routed separately to the blowdown drum with a high-level cut-off valve in each, or may be combined into a single line with high level cut-off valve. Where sub-headers of different piping materials are combined, the material of the lower-temperature header is continued for the rest of the combined line, and is also extended back into the other header for 6 m, preceded by a low temperature check valve.

5. The header and laterals should be heat traced and insulated where ambient temperatures or stream temperatures could result in solidification of heavy

process streams, or in freezing of water or moisture that may be present.

6. Several 25 mm valved stubs should be provided at appropriate points in the header, to which temporary drain connections can be made from equipment where permanent connections would not be justified because of infrequent usage or small inventory. The stubs should be located at grade, within 30 m of such equipment.

Closed Drain Headers for Special Materials

Closed drain headers are normally provided for safe drainage of equipment containing severely toxic, corrosive, pollutant or high cost chemicals (e.g., phenol, sulfuric acid, monoethanolamine, sulfur dioxide, catacarb) where there is an appreciable inventory in a number of processing vessels in a plant. The header should be at least 50 mm in diameter, and should be tied into the major vessels and equipment with 25 mm minimum size connections (20 mm is considered adequate for pumps). The header may be routed to a gravity drain drum (with recovery to the process by pump or gas pressurization), or to a pump-out pump returning to the process, or in the case of sulfuric acid, to an acid blowdown drum.

If the drain drum requires a vent, the vent should be piped to a closed system for safe location, depending upon the toxicity and vapor pressure of the materials drained.

Disposal of Hydrocarbon-Contaminated Aqueous Plant Effluents

Aqueous Drawoffs from Hydrocarbon Vessels - Water or aqueous materials that are withdrawn continuously or intermittently from vessels where they directly contact hydrocarbons (e.g., process water from distillate drums, and spent wash water or spent caustic solution from settlers) must be disposed of in such a way that entrainment or inadvertent withdrawal of hydrocarbon will not create a hazard. Disposal is therefore a function of hydrocarbon category, as follows:

1. *Vessels Containing Light Ends* - Discharges should be sent to either a water disengaging drum, a sour water disengaging drum or a spent caustic disengaging drum, and accompanied by appropriate treatment methods.

2. *Vessels Containing Hydrocarbons Heavier than Light Ends, at Temperatures Above their Flash Points*- The following notes apply to materials that do not fall into category 4 below:

a. Continuous automatic level controlled drawoff discharge as for (1), above, or into a vented section of the oily water sewer through a closed connection. If the drawoff is sour water or spent caustic, see comment (c) below.

b. Intermittent manually controlled drawoff discharge through open connection to oily water sewer catch basin. If the drawoff is sour water or spent caustic, see comment (c) below.

c. Sour water or spent caustic must be discharged as in (1) above, or to an atmospheric collection tank for subsequent disposal, provided that it has adequate venting capacity for the contingency of receiving hydrocarbon and has means of skimming the liquid hydrocarbon.

3. *Vessels Containing Hydrocarbons Heavier than Light Ends at Temperatures Below their Flash Points* - The following comments refer to normal practices for this case:

a. Discharge to an oily water sewer catch basin through an open connection, except when the drawoff is sour water or spent caustic.

b. Sour water or spent caustic must be discharged as in (1) above, or to an atmospheric collection tank for subsequent disposal, provided that it has means of skimming the liquid hydrocarbons.

4. *Hydrocarbon Liquids Heavier than Light Ends* - The following comments apply to hydrocarbon liquids that are heavier than light ends but which are at elevated temperatures such that their true vapor pressure is 103 kPa, absolute or higher. Under this condition, these products must be considered as light ends. Aqueous drawoffs from vessels containing such materials (e.g., crude desalters) must therefore be discharged in accordance as described above.

5. *Water from Tankage* - Special disengaging facilities are generally not provided for water withdrawn from offsite tankage and pressure storage.

Aqueous Effluents from Heat Exchangers

Tube failure in a water-cooled or steam-heated exchanger used in hydrocarbon service can result in the contamination of the effluent cooling water or the condensate by the process stream, especially if the latter is at a higher pressure. Such effluents must be disposed of in such a manner that the hydrocarbon contaminations can be safely contained. The following are some safe design practices :

1. Special disengaging facilities are normally required for the following cases:

a. Coolers and condensers in light ends service with the hydrocarbon inlet pressure greater than the cooling water outlet pressure under normal operating conditions, and

b. Steam heaters and reboilers with the hydrocarbon inlet pressure greater than the condensate outlet pressure under normal operating conditions, and where the hydrocarbon at steam condensate temperature has a true vapor pressure of 103 kPa absolute, or greater.

2. The special disengaging facilities normally consist of one of the following:
   a. A water disengaging drum.
   b. A cooling tower.
   c. A condensate disengaging drum, in the case of steam condensate where recovery and reuse is required.

3. For coolers, condensers, steam heaters and reboilers. The effluent cooling water and condensate are typically discharged as follows :
   a. To clean water, oily water or by-pass sewer,
   b. To an atmospheric collection tank, in the case of steam condensate where recovery and re-use is possible. If the hydrocarbon pressure at the exchanger inlet is greater than the condensate outlet pressure, then the tank must have a means of skimming the liquid hydrocarbon and the tank vent should be large enough to safely relieve the pressure generated. Also, the vent must be positioned in a safe location.

BLOWDOWN DRUMS

The purpose of a blowdown drum is to disengage closed safety valve releases and various drainage, blowdown and diverted materials into liquid and vapor streams which can be safely disposed of to appropriate storage and flaring facilities, respectively. Entrainment of liquid hydrocarbons into a flare stack is not acceptable, since the potential exists for burning liquid falling onto the ground or adjacent facilities. For this reason, a blowdown drum is required.

It is important to note that even if the blowdown is effective in disengaging liquid and vapor, further condensation could occur downstream especially if the vented vapor exits the drum at a temperature above ambient conditions. A proportion of such condensible materials in the blowdown drum vapor release may condense as a result of cooling in the flare header and contact with seal water, and then disengage in the flare seal drum; while condensible vapors which are not condensed out at this stage may condense in the flare stack or its inlet line, thus creating the potential for hazardous fallout of burning liquid from the flare. Condensed hydrocarbon in the seal drum can be entrained out with the effluent seal water, which is normally routed to the sewer, and may result in pollution, toxicity or separator overload problems. If the extent of hydrocarbon condensation downstream of the blowdown drum is such that the magnitude of these resulting problems exceeds local acceptable limits, then one or more of the following features may be considered as methods of reducing or eliminating the problem:

1. Select a condensible blowdown drum for condensible releases, rather than the non-condensible type. If a condensible blowdown drum is not suitable for handling the total blowdown service (e.g., if cold liquids are involved), then a combination of a condensible and a non-condensible drum may be used.

2. Locate the blowdown drum (when the non-condensible type is used) at a minimum permissible spacing from the flare, to minimize condensation in the flare header.

3. Install a knock-out drum immediately upstream of the flare seal drum, to remove material condensed in the flare header.

4. Provide settling facilities to separate hydrocarbon liquid from the flare seal water effluent, and appropriate means of disposal, e.g., to slop storage.

5. Where a group of connected vessels is considered as one unit for pressure relief purposes, consider the possibility of an alternative location for the PR valve such that the discharge stream would contain a smaller quantity of condensible materials.

The following are the principle advantages of condensible blowdown drums:

1. They are effective as a means of removing heavy hydrocarbon vapors from emergency release streams, thus minimizing condensation problems in downstream equipment.

2. They are effective as a means of reducing flare capacity requirements.

3. They are able to disengage oil mist better than non-condensible types .

The disadvantages of condensible blowdown drums can be summarized as follows:

1. Even though some oil may be removed through skimming connections (if provided), condensed hydrocarbon is discharged with the effluent water, often in the form of an emulsion, which may result in pollution, toxicity or separator

overload problems. However, these may be eliminated, if justified, by emulsion breaking and/or settling facilities and appropriate means of disposal for the separated oil.

2. They are unable to handle significant quantities of liquid light ends or materials cooler than 0 °C.

3. Large condensing loads, if handled on a steady state basis, result in appreciable cooling water and blowdown drum capacity requirements. These loads may be reduced, however, by the use of unsteady state condensing, e.g., by a worm cooler.

Non-Condensible Blowdown Drums (Normal Service)

A typical non-condensible blowdown drum and its associated equipment and headers are illustrated in Figure 1. A single blowdown drum may be used for more than one process unit, if economically attractive. However, when this is done, all units served by it must be shut down in order to take the drum out of service, unless cross connections are made to another system of adequate capacity. Normally all closed safety valve discharges are combined into one header entering the drum, although separate headers and inlet nozzles are acceptable if economically advantageous. The following releases are also normally routed into the safety valve header:

a. Fuel gas knockout drum condensate and absorber overhead gas knockout drum liquid.

b. Compressor suction and interstage knockout drum liquid.

c. Emergency vapor blowdowns, if provided.

d. Vapor streams diverted from process units, if this facility is provided. Dry gas streams, where there is no possibility of liquid entrainment may however be diverted directly to the flare header.

Usually, the closed liquid drain header is run as a separate line to the drum and provided with a high level cut-off valve with local manual reset. In some cases the closed drain system is segregated into a number of subheaders, as described earlier. Hydrocarbon liquids may be bypassed around the drum through a connection from the closed drain header directly to the pumpout pump suction, provided that the liquid can be routed to a safe disposal location, considering its vapor pressure and temperature. Emergency liquid pulldown connections, if provided, are routed to the blowdown drum via the closed drain header.

1. The diversion of liquid streams in the light ends range, when provided on

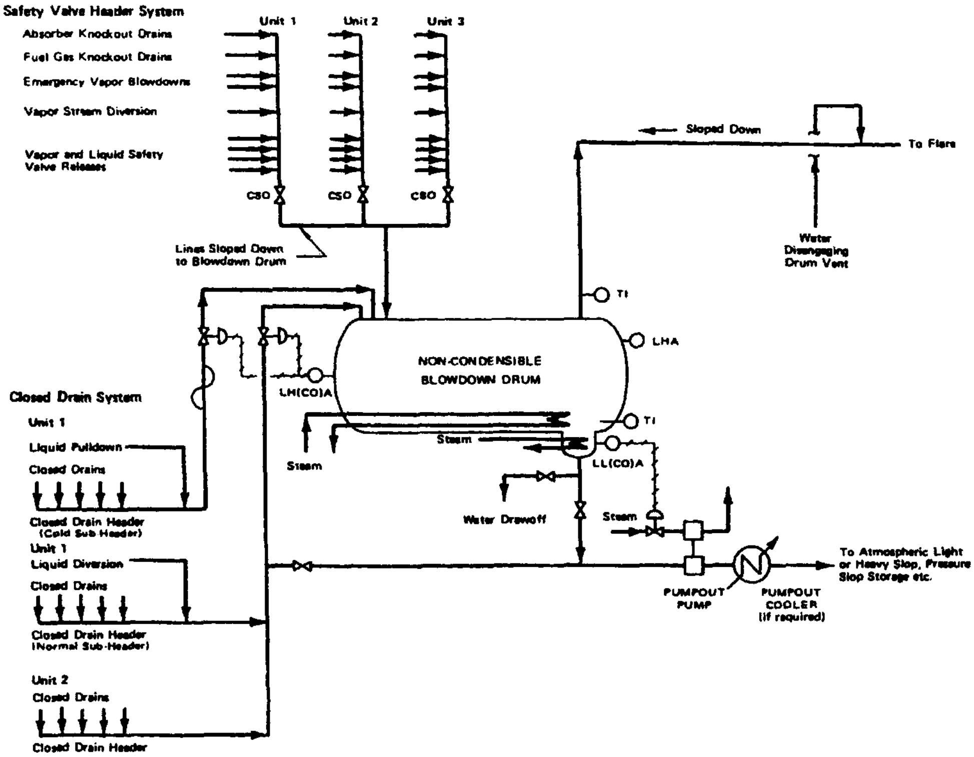


Figure 1. Typical non-condensible blowdown drum arrangement.

process units, may in some cases be routed to a non-condensible drum for disposal. In these cases, the diversion stream is normally tied into the closed drain header upstream of the high level cut-off valve, increasing the header size if necessary.

2. Sizing of the blowdown drum and location of the level instruments are based on the following guidelines. The reader should refer to Figure 2 while reviewing the following items.

a. Liquid holdup below the LH(CO)A (A-B in Figure 2) is the light ends stream diversion requirement or the closed drainage requirement for residual process liquids at a normal shutdown, whichever is greater. This closed drain requirement is taken as 10% of the total liquid hydrocarbon inventory of all vessels in one process unit which are provided with closed drain header connections. This assumes that 90% of the liquid inventory of these vessels can be removed by pressuring or pumping out through normal process disposal routes. The process unit to be used for sizing purposes is the one which has the largest closed drain requirement and which can shutdown independently for turnaround. Inventory of vessels is calculated at the top of the working level range, excluding tray holdup and the contents of piping. In the case of vessels containing large liquid inventories, e.g., surge drums, the individual closed drain header requirement may be reduced below 10%, where appropriate, by taking credit for alternative means of disposal of the liquid contents, e.g., water displacement to storage. When an oil layer is required for steam coil protection, appropriate drum volume .

b. If the facility of light ends stream diversion to the blowdown drum is required to be continuously available for safety reasons (as opposed to economic or operability reasons), then the sizing basis described in above can be modified as follows:

(1) An LH(CO)A is provided at a low level, actuating a cut-off valve in the closed drain header. Holdup below this LH(CO)A is the closed drainage requirement.

(2) A second LH(CO)A is provided at a higher level, actuating a cut-off valve in the liquid diversion stream, which is routed to the blowdown drum by a separate line.

(3) The holdup between the two LH(CO)As must equal the liquid diversion requirement.

c. The space in the drum above the LH(CO)A (above the upper LH(CO)A when two are installed) is made up of a holdup capacity (B-D in Figure 2) for 50 minutes accumulation of liquid safety valve releases, plus a vapor space (D-F) for the associated vapor release. The drum sizing is

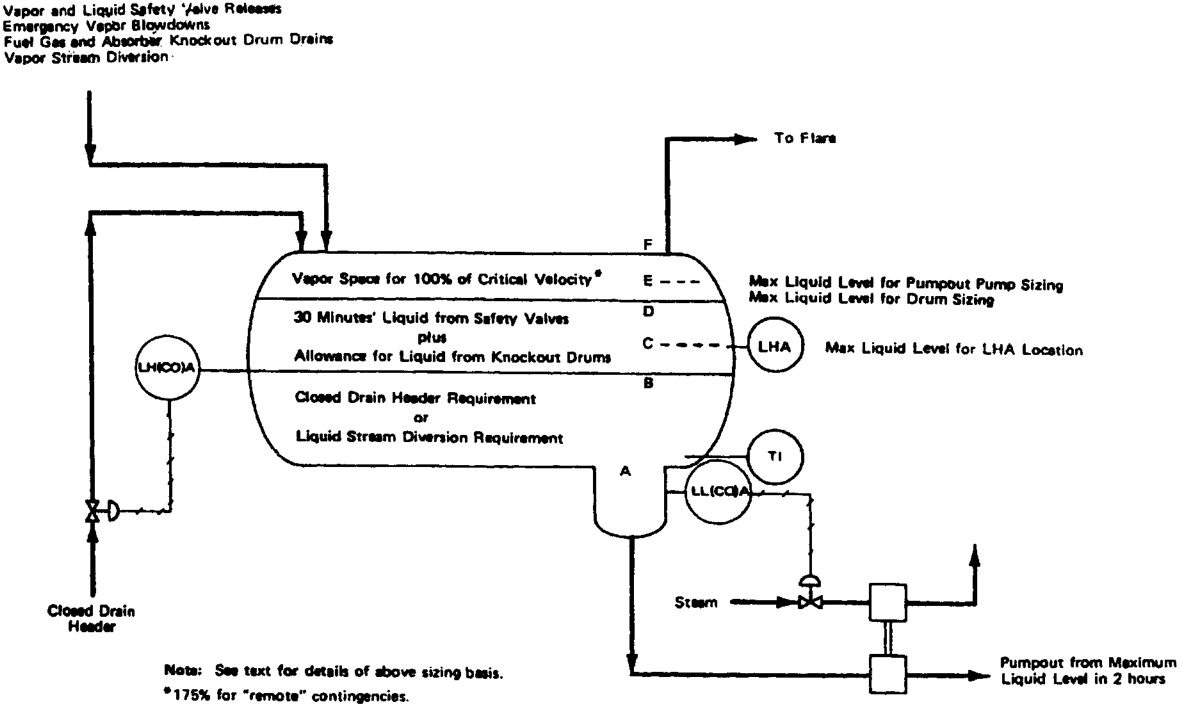


Figure 2. Non-condensible blowdown drum sizing.

determined by the single contingency which requires the maximum combined space B-D plus D-F.

d. In addition to the contingency which requires the maximum combined space B-D plus D-F and which determines the drum sizing, other contingencies are considered as follows:

(1) The single contingency which results in the largest accumulation (B-E) of liquid safety valve releases during 30 minutes, regardless of any associated vapor rate. The level at point E is used for pumpout pump sizing.

(2) The single contingency which results in the largest vapor load regardless of any associated liquid load, is used to determine the maximum required vapor space C-F, and a high level alarm is placed at point C.

e. In considering the contingencies described in paragraphs c and d above, vapor and liquid loads are evaluated on the following basis:

(1) Vapor load considerations must include all safety valve, emergency vapor blowdown and vapor stream diversion sources which release as a result of a single contingency.

(2) Liquid loads are considered from all safety valves that discharge as a result of a single contingency, plus in each case an allowance for knockout drum liquids (fuel gas knockout drums, absorber overhead knockout drums, and compressor suction and interstage knockout drums) equal to the inventory of all drums which discharge to the blowdown drum, at their LHA point.

(3) Vapor space velocities normally should not exceed 100% of critical; experience demonstrates that this keeps liquid entrainment into the flare line within acceptable limits. However, a velocity of 175% of critical is permitted when one is applyingthe 1.5 times Design Pressure Rule to remote contingencies,

$$V_C = 0.048 \sqrt{(\rho_L - \rho_V)/\rho_V} \qquad (1)$$

where : V_c = Critical vapor velocity, m/s
ρ_L = Liquid density, kg/m^3 at operating conditions
ρ_v = Vapor density, kg/m^3 at operating conditions

(4) The depth of the vapor space should under no contingency be less than 20% of the drum diameter, or less than 300 mm.

f. A LL(CO)A with local manual reset is provided to trip the pumpout pump when the liquid has been pulled down to a low level.

3. The blowdown drum design pressure is 345 kPa gage.

4. The maximum allowable operating pressure in the blowdown drum is determined by the lowest of the following:

a. The maximum allowable back pressure on safety valves which discharge to the blowdown drum, according to set pressure and type of safety valve, or

b. The pressure at which vapor diversion from any gas compressor suction to the blowdown drum is required to be released, or

c. The maximum allowable operating pressure on any other condensible blowdown drum, water disengaging drum, etc., which vents into the same flare header.

Overall sizing and pressure drop in flare systems are covered in a later chapter.

5. The blowdown drum design temperature is set by the extremes of emergency operating temperature which can result from any of the streams tied into it. If materials are handled at temperatures below 15 °C, or if they can autorefrigerate to below 15°C, a minimum design temperature must also be specified.

6. Blowdown drum materials must be adequate for any corrosive substance that may be released into it, and for the temperature limits defined.

7. A steam coil is provided in the blowdown drum for deicing, winterizing and weathering purposes. Sizing of the coil is based upon weathering off the light ends from the flashed liquids at the maximum level accumulated as a result of any of the design contingencies described above. This material must be weathered in two hours to temperature and vapor pressure conditions which will permit safe pumpout to associated slop or other receiving facilities. In some cases weathering must be followed by cooling of the pumpout stream. For most applications a steam coil consisting of a nominal 60 m of 50 mm pipe is adequate. The coil should be sloped to insure condensate drainage.

8. For services where the heating coil may be exposed to cold or autorefrigerated liquids, the design should be such as to prevent blockage by freezing of steam condensate. The following methods are available to achieve this:

a. A 50 mm steam trap bypass direct to sewer. This is required in all cases where temperatures below 0°F may occur in the blowdown drum.

b. Provision of a line for injection of low-pour gas oil or similar material into the blowdown drum. A level of gas oil submerging the coil acts as a heat sink, but this volume must be allowed for when the drum is sized. This is the normal protection used when temperatures below -45 °C may occur in the blowdown drum.

c. Connections for methanol injection into the steam coil inlet and outlet piping for deicing.

d. Use of vertical double-pipe (bayonet type) steam heaters.

e. Use of a hot oil heating medium to the coil, or a cascade heating system (e.g., steam methanol).

9. The drum is provided with a drawoff boot of nominal 600 mm diameter by 900 mm in height, with a separate steam coil fabricated from 25 mm pipe. Normally it is not necessary to withdraw hydrocarbon and water separately, and the pumpout pump takes suction from the bottom of the boot. The LL(CO)A is located as close as possible to the top of the boot to ensure that the pump is shutdown before losing suction.

10. If the drum may occasionally receive water, caustic or similar aqueous streams, which would create problems in receiving facilities if pumped out with the hydrocarbon, then means of separate drainage should be be included. This may consist of a connection to the sewer from the bottom of the boot; or in the case of sour water, a connection off the pumpout pump discharge routed to sour water facilities or other suitable disposal.

11. A steam-driven manually controlled pump is preferred for the pumpout service. A reciprocating pump is preferred due to its greater ability to hold suction with volatile liquids; however, if a centrifugal pump is used, both suction and discharge lines must be vented back to the drum, the discharge vent being sized for 15% of the pump capacity. Pump sizing is based on pumping out in 2 hours the total drum contents from the maximum accumulated liquid level. Due to the wide range of fluids handled, the pump should be specified for 2 to 2.5 m NPSH requirement at the suction flange. Drum elevation should be such as to meet the pump NPSH requirement. The pump design temperature should be the same as that of the blowdown drum, and design pressure is set according to the disposal routing downstream.

12. Disposal of pumpout material from the blowdown drum is normally to pressure slop storage, light atmospheric slop storage, or other atmospheric tankage. Design features must be incorporated to avoid the hazards of excessive vapor evolution or boilover which can result from routing light or hot materials to atmospheric tankage. In addition, a cooler should be provided in the discharge line from the pumpout pump if either of the following applies:

a. The blowdown drum can receive hot liquids (above 93 °C), or

b. The blowdown drum liquid (after weathering if necessary), if routed to an atmospheric tank for disposal, could result in the true vapor pressure of material in the tank exceeding 90 kPa, absolute.

The cooler should be sized to cool the maximum pumpout flow to 50°C.

Non-Condensible Blowdown Drums (Special Service)

In some cases, because of severe corrosion problems or for special process reasons, a unit must have its own separate blowdown system. A sulfuric acid alkylation process is an example. Here the discharge from safety valves which can contain acid emulsion presents two particular problems: corrosion and slow disengaging of hydrocarbon from acid.

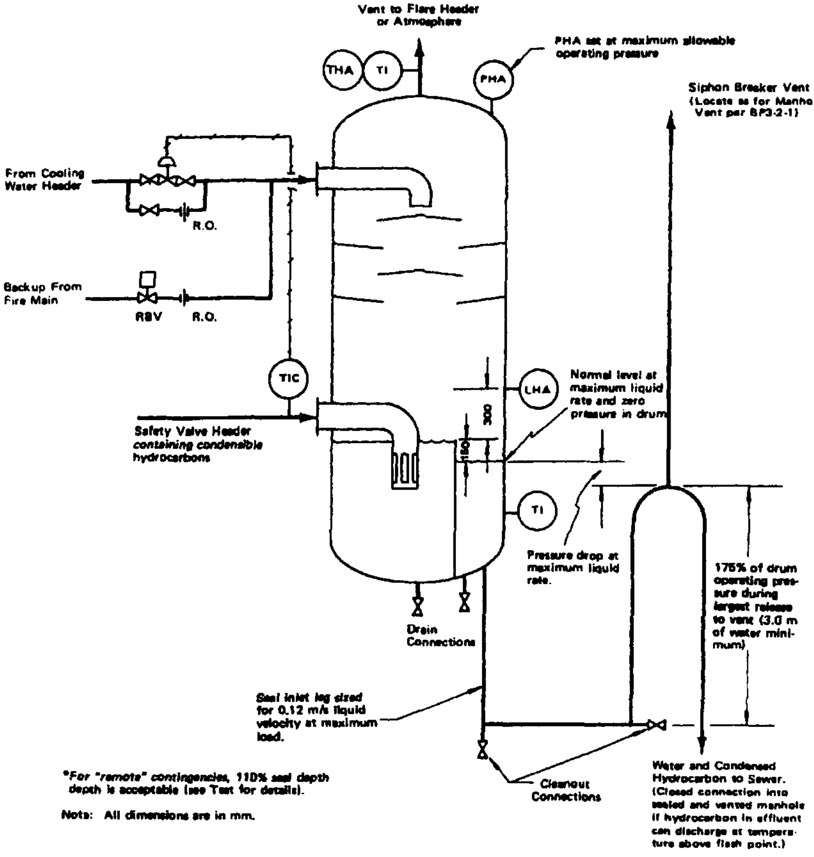
The first vessel in the blowdown system is therefore an acid-hydrocarbon separator. This drum is provided with a pump to transfer disengaged acid to the spent acid tank. Disengaged liquid hydrocarbon is preferably pumped back to the process, or to slop storage or a regular non-condensible lowdown drum. The vented vapor stream from the acid-hydrocarbon separator is bubbled through a layer of caustic soda solution in a neutralizing drum and is then routed to the flare header. To avoid corrosion in the special acid blowdown system, no releases which may contain water or alkaline solutions are routed into it.

Condensible Blowdown Drums

Condensible blowdown drums are provided as a means of preventing liquid hydrocarbon condensation in flare systems, to reduce flare capacity requirements, or to prevent discharge of condensible hydrocarbons to the atmosphere. A typical system is illustrated in Figure 3. In some cases they serve the additional purpose of reducing the maximum temperature of flared gases and hence minimizing thermal expansion problems in the mechanical design of flare stacks. A condensible blowdown drum functions by a direct contact water spray arrangement, which condenses entering hydrocarbon vapors heavier than light ends. Condensed hydrocarbons and effluent water are discharged through a seal to the sewer, and uncondensed light hydrocarbon vapors are vented to the flare or to the atmopshere.

The design basis for condensible blowdown drums is as follows:

1. The maximum vapor load on the drum is based on the largest release from safety valves discharging as a result of a single contingency. Vapor velocities in the drum are based on 100% of critical velocity (refer back to Equation 1). However, a velocity of 175% of critical is permitted when one is applying the 1.5



*For "remote" contingencies, 110% seal depth depth is acceptable (see Text for details).

Note: All dimensions are in mm.

Figure 3. Condensible blowdown drum.

times Design Pressure Rule to remote contingencies.

2. The vapor outlet should preferably be connected to the flare system. However, when the safety valve releases and other streams tied into the drum contain only a small quantity of noncondensible hydrocarbons or inerts, and where no pollution problems are anticipated, then an atmospheric vent is acceptable, subject to the following conditions:

a. The vent must be located at least 15 m above grade, and at least 5 m higher than any equipment with a horizontal distance of 15 m.

b. The vent must be located such that if inadvertent ignition of the maximum hydrocarbon vapor release should occur, the resulting radiant heat densities at grade do not exceed the prescribed limits for personnel exposures.

c. The dispersion of flammable or toxic materials must be adequate in relation to adjacent equipment and working areas.

d. The vent must be provided with steam or inert gas injection for flashback protection and snuffing.

e. The drum design pressure should be 1030 kPa gage.

3. The drum design pressure should be 345 kPa gage, unless the drum is connected directly to the flare (without a seal drum), in which case the design pressure of the blowdown drum should be 1030 kPa gage.

4. Water requirements are normally based on reducing gas and liquid outlet temperatures to about 65 °C. Selection of the optimum temperature is based on considerations of temperature and composition of entering streams, and the extent to which subsequent condensation of effluent vapors downstream of the drum can be tolerated. The water supply should be taken from a reliable water system. If a recirculating cooling water system is used, then the circulating pumps and cooling tower basin must have adequate capacity to supply the maximum condensible blowdown drum requirement for 30 minutes. An on-off type temperature controller in the inlet line actuates the control valve in the water supply line, and a restriction orifice in the control valve bypass admits 0.6 to 1.2 dm^3/s continously to maintain the outlet seal. An emergency backup water connection from the firemain is provided, with an RBV actuated from the control house, and a restriction orifice sized for the maximum water requirement. A high temperature alarm is required in the vapor outlet from the drum.

5. The water holdup in the base of the drum is sized.

6. The seal height in the liquid effluent line (assuming 100% water) normally is sized for 175% or the maximum drum operating pressure, or 3 m, whichever is greater.

7. The maximum allowable operating pressure for the purposes or sizing flare headers and calculating safety valve back pressures is taken as 7 to 14 kPa gage at the drum, unless lower pressures are required by special process considerations. Consideration must be given to steam that is generated by evaporation or cooling water at high loadings.

8. Because of the continuous water flow through a condensible blowdown drum, it can safely handle cold or autorerrigerating releases only to the extent that effluent liquid and vapor temperatures remain above 0°C.

Unsteady State Condensible Blowdown Systems

In some cases where condensing loads are high, or where it is required to recover condensed liquid blowdown material for pollution, toxicity or economic reasons, an unsteady state condensing system may be appropriate. Examples or such applications are as rollows:

By use of a worm cooler in the condensible blowdown drum inlet, the high water requirement for direct contact condensing is avoided. The worm cooler must be elevated and the coil continuously sloped to avoid any liquid traps in the safety valve header. The holdup or static water in the cooler must be adequate for the design condensing duty during 30 minutes, allowing for heating-up of the water in that period. This type or design should not be applied where solidification or heavy materials within the cooler coil may occur.

Another example of an unsteady state condensible blowdown system is the design for a phenol condensible blowdown tank. A blowdown tank is used in phenol treating plants to handle streams containing phenol and heavy hydrocarbons (lubricating oil stocks). The blowdown tank is illustrated in Figure 4. The design basis is as rollows:

a. The maximum vapor load to the tank is based on the largest release from safety valves discharging as a result of a single contingency.

b. Equipment design temperature is 175°C. Design pressure is 1.5 kPa gage in the vapor space with the tank full of liquid, and the tank construction incorporates a weak roorshell weld seam.

c. The atmospheric vent stack is concentric with the tank and terminates at least 15 m above grade and at least 3 m above the highest equipment within a horizontal distance of 15 m. Additional elevation must be provided if necessary to ensure that phenol concentrations at grade and at working platforms do not exceed the Threshold Limit Value (TLV). The stack is slotted at the top or the tank, with the slot area sized for a maximum tank vapor space pressure or 1 kPa gage at maximum vapor load.

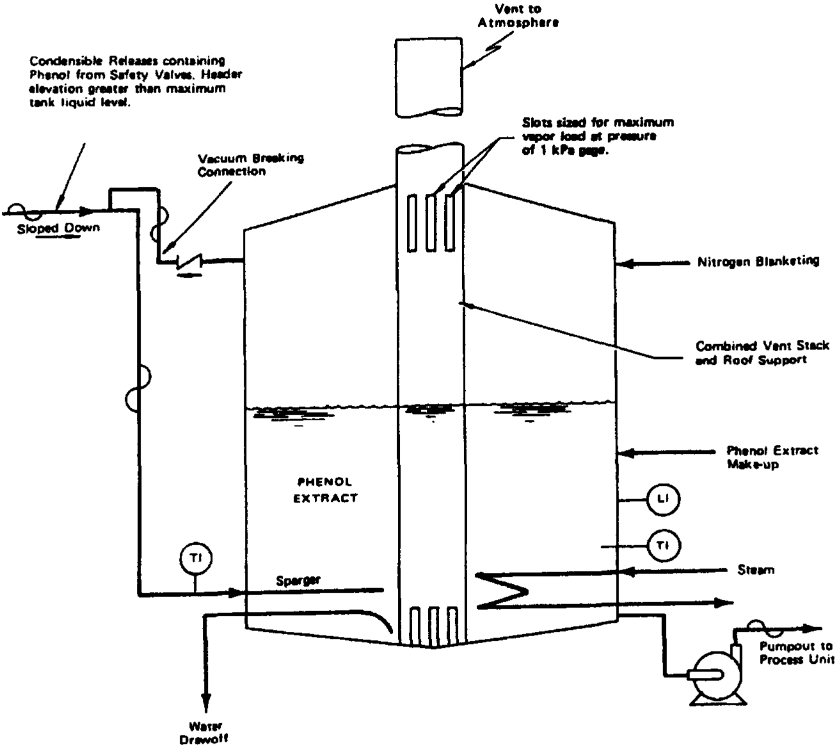


Figure 4. Condensible blowdown tank - phenol service.

d. The tank is provided with continuous nitrogen blanketing for flashback protection.

e. A level of phenol extract is held in the tank to cool and absorb the phenol in the entering vapors. Sufficient extract, at 58°C, is required to absorb the largest quantity of phenol discharged in 50 minutes by safety valves as a result of a single contingency, without exceeding 93°C extract temperature. Tank size must be adequate to hold the mixed phase inventory existing during maximum load conditions, and tank dimensions are selected such that the maximum liquid level is below the elevation or the header which collects the phenol-bearing releases.

f. The tank bottom must be designed for complete withdrawal of water. Safe disposal for the phenolic water is required; normally this is returned to the process.

g. Pumpout facilities are provided to return saturated liquid to the process after a safety valve discharge into the tank.

h. Inlet vapor is distributed through the tank by a sparger.

Condensible Blowdown Tanks for Other Service

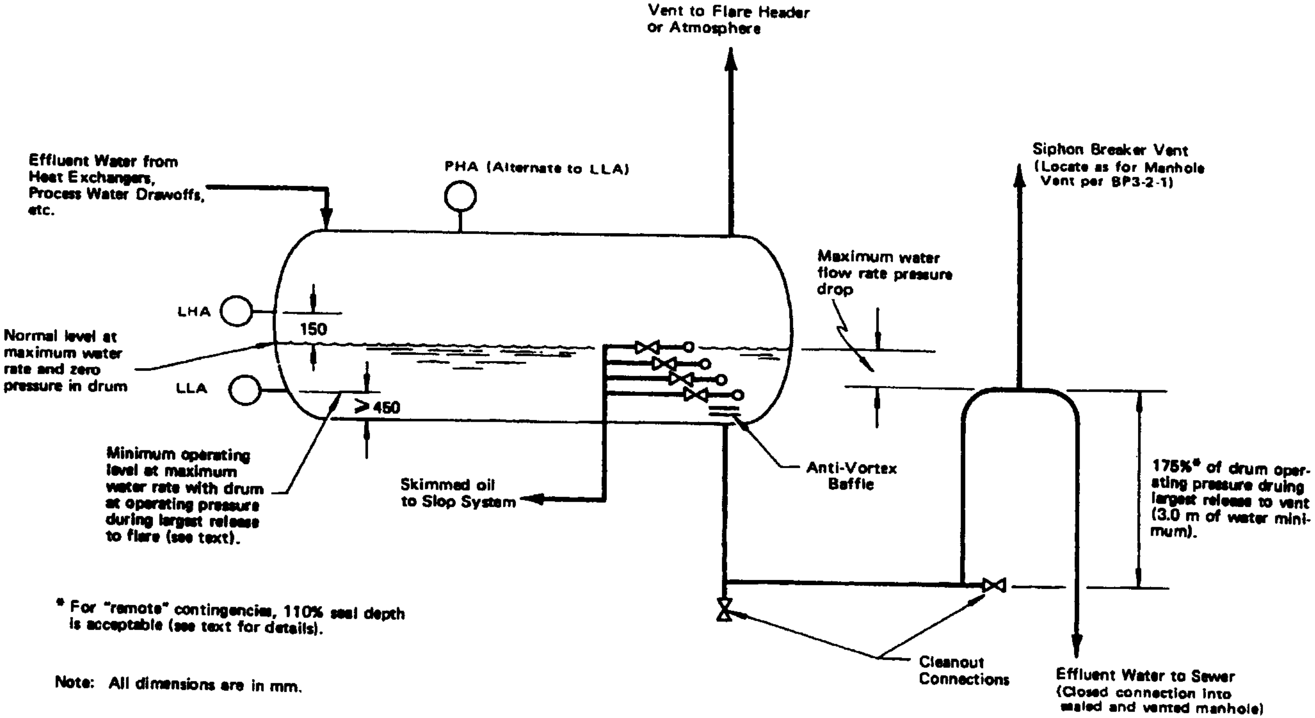
A condensible blowdown tank, designed on a similar basis to that described above for phenol, may be provided in other services where a conventional condensible blowdown drum would not be acceptable (e.g., due to effluent water pollution considerations). Examples of such cases are methyl ethyl ketone (MEK) and dimethyl formamide (DMF). A suitable absorbing material is specified (e.g., a lube oil stock for MEK; water for DMF), and the design must include consideration of maximum permissible operating temperatures to prevent excessive vapor evolution or the boiling of water.

It is important to note that in some installations where local pollution regulations would not permit venting a condensible blowdown tank in toxic service to the atmosphere, a pressure drum or sphere, vented to a flare, may be necessary.

Effluent Disengaging Systems

Disengaging drums are provided to remove hydrocarbon liquid and vapor contaminants from aqueous plant effluent streams, to permit them to be safely discharged to the sewer. Criteria for the routing of effluent streams to disengaging drums were described earlier.

A typical water disengaging drum system is illustrated in Figure 5. The



* For "remote" contingencies, 110% seal depth is acceptable (see text for details).

Note: All dimensions are in mm.

Figure 5. Water disengaging drum.

liquid inlet header is sized for the maximum water rate to the drum. The available pressure drop for flow is based on the vessel from which the water has the lowest exit pressure and the maximum allowable back pressure on the drum.

The vapor load on the drum results from the entering hydrocarbon vapor or liquid flashing to equilibrium conditions at atmospheric pressure. The design vapor load is the largest quantity of vapor resulting from a single contingency, such as a split exchanger tube or failure of a water drawoff valve in the wide open position (multiple control valve failures are not considered, provided that the control valves are specified to close on air failure). Water drawoffs are examined to determine the maximum hydrocarbon load resulting from the water outlet control valve failing wide open, with flow assumed to be all hydrocarbon.

The vapor outlet may be discharged either to the atmosphere or to a flare. Atmospheric discharge may be considered, provided that a safe location can be achieved, as defined by the following :

a. The vent must be elevated to at least 15 m above grade and at least 3 m above the highest equipment within a horizontal distance of 15 m.

b. It must be located such that if inadvertent ignition of the maximum hydrocarbon vapor release should occur, the resulting radiant heat densities at grade do not exceed the prescribed limits for personnel exposures.

c. The TLV of any toxic vapor that may be discharged from the vent is not exceeded at grade or at any working platform.

d. An open-ended vent pipe directed vertically upwards is required, with steam or inert gas injection for flashback protection and snuffing.

The maximum allowable operating pressure in the disengaging drum for purposes of sizing flare headers, water headers, and laterals from exchangers is determined by the lower of the two following values: (operating pressure should be designed as low as practical, to maximize disengaging effectiveness.)

a. The maximum allowable operating pressure on any condensible or noncondensible blowdown drum which vents into the same flare header, or

b. The maximum pressure at which normal water flow can still enter the drum from the lowest pressure source.

A design pressure of 545 kPa, gage is normally specified for water disengaging drums. The water outlet system is designed to seal the drum and prevent entrainment of hydrocarbon or air into the sewer. Figure 5 indicates the normal layout incorporating a single loop seal.

Sizing of the drum and seal leg is determined by the following:

a. When the pressure in the drum vapor space is atmospheric, the level must be such that the vapor space is adequate to disengage hydrocarbon droplets from hydrocarbon vapor at 15% of V_c at design vapor load, and the seal leg must have a large enough diameter to pass the maximum water rate. If operating variations could result in less than 60 dm^3/s of water to the water disengaging drum, an external system should continuously provide 60 to 120 dm^3/s to maintain the seal.

b. When the pressure in the vapor space equals the maximum allowable operating pressure, the water surface should be depressed to a minimum operating level, at which capacity for disengaging vapor from water is adequate for the design load. Also, the back pressure must not depress the water level below the effective depth of the vortex breaker. This applies, regardless of whether the back pressure results from vapors being released through the drum, or from the back pressure surge imposed from a closed release system. The minimum operating level should be not less than 450 mm above the bottom of the drum. A low level alarm (LLA) is provided at the minimum operating level, or alternatively a high pressure alarm may be installed, set at the corresponding vapor space pressure.

c. The height of the seal in the liquid outlet normally should be equivalent to 175% of the maximum allowable operating pressure, or 3 m, whichever is greater. However, a seal height of only 110% of the maximum drum operating pressure is permitted when one is applying the 1.5 times Design Pressure Rule to remote contingencies.

The drum should be provided with a high level alarm (to give warning of overload or seal blockage) located 150 mm above the level which corresponds to zero gage pressure in the vapor space and maximum water flow. Also a vortex breaker should be installed at the water outlet to prevent hydrocarbon entrainment to the sewer. Effluent water from the seal is discharged through a closed connection to a vented sewer manhole, so that any air drawn in through the siphon breaker vent may be disengaged, and to prevent hydrocarbon release at grade level.

Four skimming connections with trycocks are normally provided at the outlet end of the drum, at the normal liquid level, and at 150 mm, 450 mm, and 500 mm below the normal level. Liquid hydrocarbon skimmed from these connections can be pumped to a suitable slop system. A connection to the suction of a blowdown drum pumpout pump, if available, is adequate for this purpose.

Disengaging Drums for Other Aqueous Streams

Aqueous plant effluent and drawoff streams such as steam condensate, sour water, or spent caustic soda solution may require disposal to a disengaging drum, but the regular water disengaging drum may not be suitable. Special disengaging drums may therefore be required for example, in the following cases:

1. Condensate is to be recovered and returned to treating and boiler feed water facilities.

2. Sour water is to be routed to sour water stripping facilities.

3. Spent caustic is to be recycled to fresh caustic make-up facilities, or routed to deodorizing or other disposal facilities.

The design of these drums generally follows the same basis as that for water disengaging drums, except that a pump (with spare) is required to transfer the aqueous liquid under level control to the appropriate receiving facilities.

Combination Disengaging and Blowdown Drums

In some cases, it is possible to combine the functions of blowdown and disengaging drums in one vessel. However, PR devices discharging liquid hydrocarbons lighter than pentane should not be connected into the drum if there is a possibility that such liquids could accumulate and be released to the sewer through the seal leg. Also, the drum vent should be sized to prevent pressure buildup due to vaporization. In these applications, the design criteria for both services must be met and special attention should be paid to potential hazards and problems which may be introduced, such as :

1. Liquid traps in safety valve release lines.

2. Combinations of water and hot hydrocarbon releases which could result in steam generation and pressure surges.

3. Combinations of water and cold or autorefrigerating hydrocarbons which could result in freezing problems.

4. Reliability of water supply, if condensible blowdown and water disengaging services are combined.

5. Contingencies which may require the drum to be used for both services simultaneously.

Process Stream Diversion and Slop Storage

Plant designs must include means of safe disposal for various slop materials, such as the following:

1. Liquid hydrocarbons accumulated in non-condensible blowdown drums, originating from safety valves, closed drain headers, knockout drum drainage, etc. Facilities are normally provided at the drum for weathering volatile liquids and cooling hot liquids before disposal.

2. Oil-water mixtures and emulsions, e.g., from separators, tank bottoms, ballast water, etc. Heating of such materials is often necessary to separate oil and water.

3. Off-specification products during startup, shutdown or plant upsets. Means of disposal of all off-specification product streams must be available. In many cases, blending off in product tankage or downgrading of another product is possible.

4. Streams which must be diverted because of emergency shutdown of downstream equipment (e.g., compressor failure). Diversion routes should be provided where such a contingency would otherwise require the immediate shutdown of the affected process unit, resulting in appreciable economic and operational debits.

The following means of disposal may be considered for disposal of slop materials such as the above:

1. *Flare* - Vapor streams, such as compressor suction diversion on catalytic cracking and steam cracking units, are normally routed to a flare.

2. *Gas burning main* - Light hydrocarbon vapors may be routed into gas burning mains for disposal. If a propane vaporizer is available, this may be used as a means of routing liquid light ends to the burning main.

3. *Storage facilities* - Recycling or blending of liquid streams into feed or product storage, etc., can be used in many cases. However, the design of such disposal systems must include consideration of the potential for excessive vapor evolution and boilover that can arise from routing light materials or hot streams to tankage.

4. *Slop storage* - Slop storage facilities are of three basic types according to the materials handled:

a. Pressure slop storage, for light ends materials.

b. Light atmospheric slop storage for materials which do not require heating for emulsion breaking.

c. Heavy atmospheric slop storage, for materials requiring heating for emulsion breaking.

Materials accumulated in slop storage are normally routed to rerunning facilities or blended into appropriate tankage for disposal.

When selecting the means of stream disposal one should use routes which, to the maximum extent possible, utilize normal plant facilities and tankage, etc. Streams which cannot be handled in this way require slop storage facilities. Sizing that cannot be handled in this way require slop storage facilities. Sizing of slop storage facilities is usually based upon the normal flow rates of all streams which must be diverted to slop under a single contingency, for the period of time necessary to eliminate the contingency or to carry out a controlled shutdown.

If pressure slop storage is required to handle slop materials in the light ends range, it must comply with the following:

a. The type of pressure slop storage vessel is selected on the basis of cost. Generally a sphere or spheroid is cheaper than a drum for capacities over 160 m^3.

b. The vessel is vented to a low pressure gas line (if available) or to the flare header, through a pressure control valve. The design vapor load is based on the single contingency (e.g., feed diversion from a particular unit) which results in the largest quantity of vapor flashed from entering liquids.

c. Overpressure and vacuum protection must be provided.

d. A high level alarm is provided at 85% of the vessel volume capacity, and also a high level cut-off which shuts off inlet flow when the level reaches 92% full.

e. A manually controlled pumpout pump is provided to transfer slop to a suitable process unit for rerunning. The size of the pump is determined by rerunning requirements. When the blowdown drum and slop storage vessel are close together, the pumpout pumps may be manifolded so that both are interchangeable in either service.

As an alternative to special pressure slop storage, the necessary holdup may be provided in a non-condensible blowdown drum.

Light and heavy atmospheric slop storage are normally both necessary in the average refinery. Sizing depends on the refinery complexity, volume and number of finished and intermediate products, and available means of disposal or rerunning.

10

FLARE GAS DESIGN PRACTICES

INTRODUCTION

This chapter discusses some of the criteria for selecting, designing and spacing elevated, burning-pit, and multijet flares. The design of safety valve and flare headers was covered in an earlier chapter, as well as discussions concerning associated blowdown drums, water disengaging drums, etc.

The flare is a key component of the closed emergency release system in a refinery or chemical plant. Emergency releases originating from safety valves, vapor blowdowns, process stream diversion, etc., and equipment drainage, which cannot be discharged directly to the atmosphere for reasons of safety or pollution control, are routed through closed systems to a blowdown drum where liquids and vapors are separated. Other emergency vapor streams are vented from disengaging drums handling contaminated effluent cooling water, process water drawoffs, etc.

The flare provides a means of safe disposal of the vapor streams from these facilities, by burning them under controlled conditions such that adjacent equipment or personnel are not exposed to hazards, and at the same time meeting pollution control and public relations requirements.

FLARE TYPES AND APPLICATION

Three types of flare systems are commonly used : the elevated flare, the ground flare, and the burning-pit flare. Although the three basic designs differ considerably in required capital and operating costs, selection is based primarily on pollution/public relations considerations such as smoke, luminosity, air pollution, noise and spacing factors. Table 1 summarizes the advantages and

Table 1. Comparison of flare types.

	Elevated Flare	Multijet Flare	Burning Pit Flare
Pollution Characteristics			
Smoke	Can be made smokeless except at high loads.	Relatively smokeless	Poor
Noise	Noisy, due to steam used for smoke reduction (compromise necessary).	Relatively quiet	Relatively quiet
Luminosity	High, but can be reduced with steam.	Some	Relatively quiet
Air Pollution (odor)	Best obtainable, if elevation is adequate.	Poor dispersion, because of low elevation; severe problems if poor combustion or flameout.	Poor
Major Problems	● High cost if high elevation. ● Visual and noise pollution. ● Radiation requires wide spacing.	● High cost. ● High maintenance requirement. ● Low level odor pollution. ● Hazardous if flameout occurs.	● Low-cost and simple; but pollution is not acceptable in most cases. ● Wide spacing required.
Application	● General choice for total flare load, or as overcapacity flare in conjunction with multijet flare. ● Generally the only acceptable flare where products of combustion or partial combustion are toxic or malodorous.	● Use for base load or partial flaring rates if noise and visual pollution are critical. ● Suitable only for "clean burning" gases, i.e. where products of combustion are not toxic or malodorous. ● Not suitable upwind of residential areas.	● Remote locations where no pollution requirements apply and space is available.

disadvantages of the various types of flares. A brief description of each type of flare system is provided below.

Elevated Flares

The elevated flare, by the use of steam injection and effective tip design, can be made smokeless and of reasonably low luminosity up to about 20% of maximum flaring load. Steam injection introduces a source of noise, and a compromise between smoke elimination and noise is usually necessary. If adequately elevated, this type of flare has the best dispersion characteristics for malodorous and toxic combustion products, but visual and noise pollution can present public relations problems. Capital and operating costs are relatively high, and an appreciable plant area may be rendered unavailable for plant equipment because of radiant heat considerations.

Despite its disadvantages, the elevated flare is the general choice either for total flare loads, or for handling overcapacity releases in conjunction with a multijet ground flare. For most applications, the elevated type is the only acceptable means of flaring "dirty gases," i.e., gases high in unsaturates or hydrogen sulfide, or which have highly toxic combustion products.

Stacks for Elevated Flares

Three types of stack for elevated flares are used:

Guyed Stack - This type is usually the least expensive to build but in some cases the guy wires result in restrictions on the use of adjacent land, in addition to normal spacing restrictions.

Derrick Type Stack - This type of unit is well-suited for tall structures subject to strong winds. However, derrick type stacks are the most expensive to erect and maintain.

Self-Supporting Stack - This type of unit is designed so that the flare riser pipe has no lateral structural support. For short flares, this type is the least expensive to erect and maintain.

Ground Flares

Various designs of ground flare are available. The type which has been used almost exculsively is the multijet flare.

Smokeless operation can generally be achieved, with essentially no noise or luminosity problems, provided that the design gas rate to the flare is not exceeded. However, since the flame is near ground level, dispersion of stack releases is poor and this may result in severe air pollution or hazard if the combustion products are toxic or in the event of flame-out. Capital and operating cost and maintenance requirements are high.

The multijet flare is suitable for "clean burning" gases (i.e., where toxic or malodorous concentrations are unlikely to be released through incomplete combustion or as combustion products) when noise and visual pollution factors are critical. It should not be used in locations upwind of adjacent residential areas. Generally, it is not practical to install multijet flares large enough to burn the maximum release load, and the usual arrangement is a combination with an elevated overcapacity flare. The latter is normally not provided with steam injection, and smoke formation is accepted during the small number of major releases.

Other designs of ground flares suitable for refinery application are available. In some of these cases, noise is appreciable, in comparison with the multijet type, but their compact size, low space requirement, simplicity, and hence low cost, may give an overall advantage.

Burning-Pit Flares

The burning pit is of simple construction, with low capital and operating costs, and it can handle liquid as well as vapor hydrocarbons. Its use is usually limited by spacing requirements and smoke formation, and it is applied only in remote locations where there are essentially no pollution restrictions.

BASIC DESIGN CONSIDERATIONS

In this section we shall discuss flare spacing, location and height. Spacing, location and height of flares are determined by consideration of the following factors:

Radiant Heat - Acceptable levels of radiant heat density in areas where personnel may be present.

Burning Liquid Fall-Out - The possibility of burning liquid fall-out from an elevated flare, if liquid hydrocarbons should be entrained into it.

Pollution Limitations - (i.e., smoke formation, malodorous or toxic combustion products, noise) which may be based on statutory and/or public relations requirements.

Flare Capacity and Sizing

Flare systems are designed to handle the largest vapor release from safety valves, vapor blowdowns and other emergency systems, which results from the design contingency. Normally the largest single contingency basis is used.

Flare sizing is covered in some detail under a later section in this chapter, as well as sizing of the overall flare system.

Flashback Seals

Flare systems are subject to potential flashback and internal explosion since flammable vapor/air mixtures may be formed in the stack or inlet piping by the entry of air, and the pilot constitutes a continuous ignition source. Flares are therefore always provided with flashback protection, which prevents a flame front from travelling back to the upstream piping and equipment. Design details are described later.

Design of Elevated Flares

We shall first consider the design of elevated flare systems. Sizing of flare systems is a function of maximum allowable back pressure on safety valves and other sources of release into the emergency systems.

Flare stack sizing and pressure drop is included with considerations of pressure drop through the safety valve headers, blowdown drums, flare headers, seal drum, etc. Elevated flare tips incorporating various steam injection nozzle configurations are normally sized for a velocity of 120 m/s at maximum flow, as limited by excessive noise and the ability of manufacturers to design tips which will insure flame stability. This velocity is based on the inclusion of steam flow if injected internally, but the steam is not included if added through jets external to the main tip.

Pressure drops for tips are obtainable from manufacturers' charts such as

Figure 1. Available pressure drop may in some cases dictate acceptance of a lower maximum velocity, but at least 75 m/s is recommended to insure good dispersion. Flare tips consisting of a simple open-ended pipe with a single pilot are subject to flame lift-off and noise problems at lower velocities, and should therefore be designed for a maximum velocity of 50 m/s.

Elevated Flare Location, Spacing and Height - Location, spacing and height of elevated flares are a function of permissible radiant heat densities, possible burning liquid fall-out, and pollution considerations. Design requirements are as follows :

1. Flares should be at least as high as any platform within 150 m horizontally, and in no case less than 15 m high.

2. Any source of ignitable hydrocarbons, such as separators or floating roof tanks, should be at least 60 m from the base of the flare stack.

3. Flare location and height must be such as to meet all applicable standards of noise level and atmospheric pollution by combustion products.

4. Flare elevation and spacing must be such that permissible radiant heat densities for personnel at grade are not exceeded under conditions of maximum heat release. The appropriate calculation procedures and personnel exposure criteria are described later. In some special cases, flare elevation and spacing may be governed by radiant heat exposure of certain vulnerable items of equipment, rather than personnel.

5. Flares should be located to limit the maximum ground level, heat density to 1.6 kW/m^2 at any property line. The minimum distance from the base of the flare stack to the property line should be 60 m.

Elevated Flare Tips and Steam Injection

Types of Flare Tip - A simple open tip with no steam injection is used when there are no limitations on smoke emission and luminosity. Flare tips incorporating steam injection are used where smoke emission and luminosity must be minimized. Generally, tip selection and steam injection capacity are designed to give smokeless operation at flare gas rates up to about 20% of design, since this will cover a large proportion of releases in a typical plant. The economics of providing continuously available reserve steam capacity for higher loads are prohibitive unless special pollution regulations apply. The required steam to hydrocarbon ratio for smokeless operation is mainly a function of unsaturates

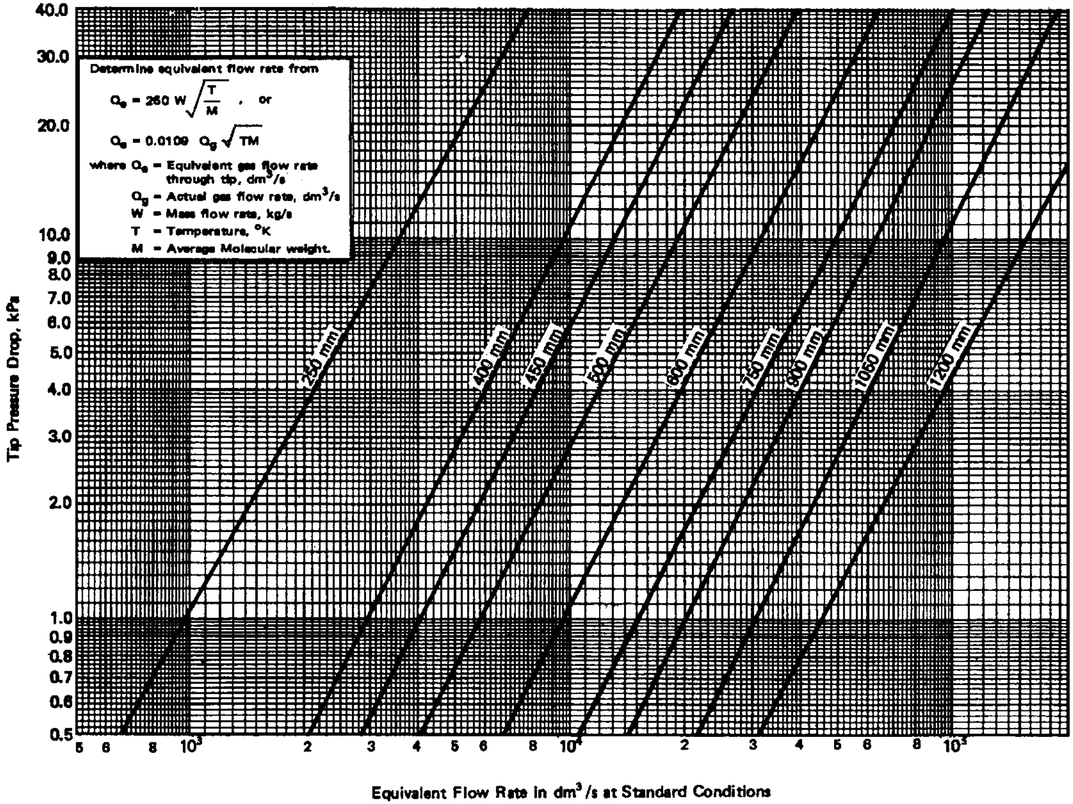


Figure 1. Manufacturer's chart showing tip pressure drop as a function of diameter and equivalent flow rate.

content of the flare gas, although molecular weight, discharge velocity and type of flare tip also have an effect.

Recommended nominal steam rates at 60 m/s exit velocity for a typical flare tip are shown in Figure 2. At lower velocities, higher steam ratios are required. Typical steam control consists of a flow ratio controller with adjustable ratio set point, related to flare gas flow. The ratio adjustment, located in the control house, provides for the higher steam ratios necessary at low flaring rates.

The topic of flare gas measurement is treated below. If necessary, continuously vented surplus low pressure steam can be used for smoke control at low flaring rates, with high pressure steam cutting in through a flare ratio controller designed to handle large releases.

Steam injection introduces an additional source of noise. An effective flare tip is one which achieves a good balance of smoke and luminosity reduction without exceeding acceptable noise levels. Low-frequency noise is encountered at relatively high steam to hydrocarbon ratios.

A flare performance chart for the hydrocarbon being flared, should be consulted for additional guidelines on flare tip design. Figure 3 provides a provisional performance chart for propane. The chart defines the design envelop of exit velocities and steam ratios necessary to avoide smoke formation, excessive noise, flame boilover and flame lift-off.

There are various designs of flare tips that incorporate such features as central steam injection, an annular ring of steam nozzles, internal air-inspirating steam nozzles, windshields, etc. Table 2 provides some details of suitable types from which selection may be made.

Some final details to note on flare tips include:

a. Incoloy 800 and stainless 310 are the recommended materials for flare tips,

b. The length of the flare tip is normally 3 m,

c. A maintenance platform should be provided at the base of the flare tip (at the flange), with access by means of a caged ladder with intermediate platforms. Access onto the ladder should be permitted only when the flare is out of service.

In the consideration of elevated flare pilots and igniters, proprietary flare tips are normally provided with the manufacturer's recommended igniter and pilot system. Usually, one to four pilots are used depending on the flare tip type and diameter. The forced air supply type of igniter system (described below) is normally preferred. Controls should be located at a distance from the base of the

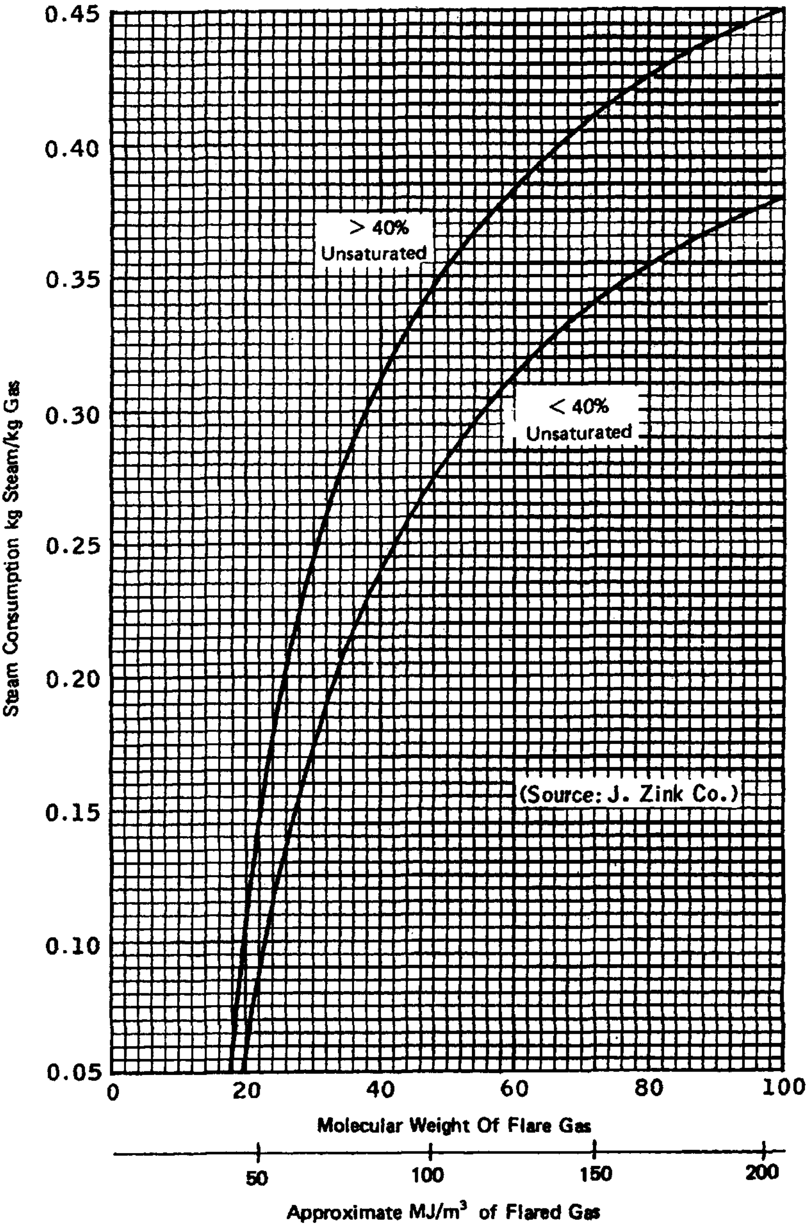


Figure 2. Nominal steam requirement for a typical flare tip.

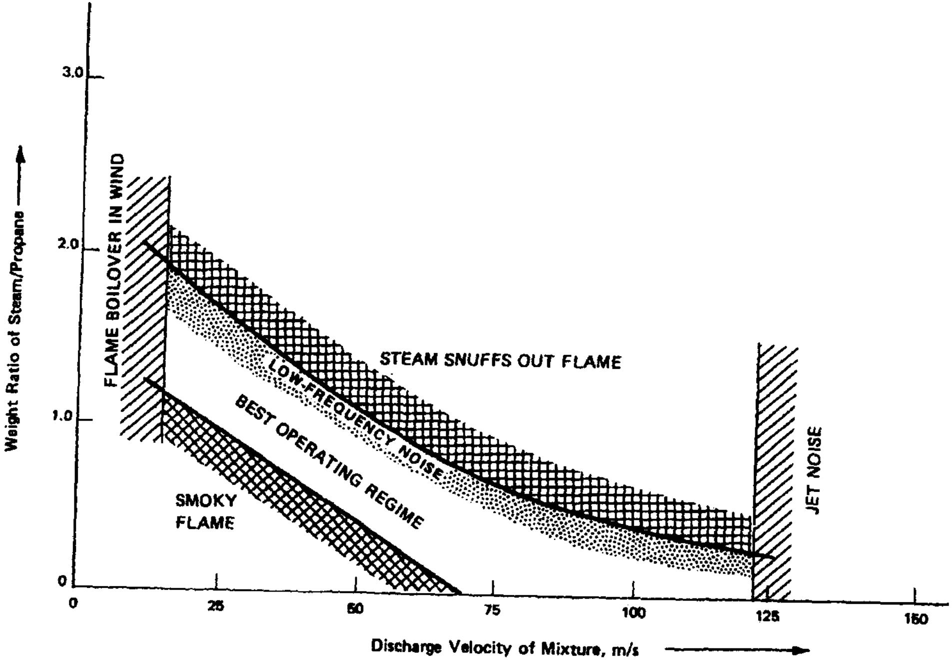


Figure 3. Flare performance chart for propane.

Table 2. Elevated flare tips.

Type	Generic Name	Features
Smokeless	Steam Ring	Basic smokeless flare. Relatively simple to build and inexpensive. Design of steam nozzles varies among vendor. Center steam nozzle is sized for smokeless burning of purge gas (optional). Refractory generally disappears after several months of service. Steam injection system can be sized for steam pressures as low as 140 kPa at tip. External steam injection produces steam noil. Generally design for 7 kPa AP of flare gas at the tip, but can vary in both directions. Steam lines are usually sized for smokeless operation at up to 20% of peak flaring rate, but can vary from this if more steam is avails. Low frequency flare noise has been observed with this type of tip at low flaring rates. Wind loading is relatively low, since tip diam is the same as the stack diameter.
Smokeless	Steam Ring plus Steam Inspirating Air	Combines steam ring with internal injection of steam and inspirated air. Designed to use internal steam for smoke suppression at low flaring rates, with top ring cutting in at higher rates.
Smokeless	Steam Inspirating Air	Coanda effect nozzles are used to inspirate air into the mixing chamber. Claim high efficiency of steam use. Claim low steam noise. Complex nozzles are a high-cost component. Brochure shows much R&D back-up of basic design, in marked contrast to those of other vendors. Coanda effect is well understood and widely used, therefore working principle is well established.
Smokeless	Steam Inspirating Air	Flare gas flows from the stack into the center body and then into spokes. Air flows between spokes. Basis for this design is jet engine afterburner technology. Air flow can be induced by steam injected at throat of venturi shroud or can have forced draft in an air stack concentric with the flare stack. Flow control plates in spokes are potentially subject to fouling. High aerodynamic loading. Typical shroud diameter is 2235 mm and height is 2440 mm for 600 mm diam. flare stack. Expensive sheet metal work. Quite Different designs for refinery & chem plant flaring, purging of tankers, pipeline flaring, etc. No refractory is used. Measured metal temperatures ~ 230° C with steam flow. Claim 7 kg to 9 kg air inducted per kg steam.
Smokeless	Center Steam	Cheapest steam-injection flare tip. Steam jet emerges at high velocity and penetrates to the exit plane of the flare without mixing completely with flare gas. Results are intense steam noise (much greater than with steam ring for the same steam rate) and higher steam consumption than the steam ring.
Smokeless	Coanda	Gas releases at base of elevated tulip create low pressure region. Gas film follows Coanda profile mixes with air and is ignited by pilot.
Non-Smokeless	Utility or Field Flare	Cheapest flare tip. Produces trailing smoke even with natural gas.

flare such that the pilot ignition can be readily observed, subject to a minimum of 7.5 m. Strainers are generally needed in order to prevent plugging (pilot nozzles are small and hence subject to fouling).

Multijet Flares

Generally, the economics of providing a multijet flare sized to handle the entire release from the largest design contingency are prohibitive, because of the low frequency of such major releases. Instead, it is designed to handle a proportion of the maximum flow (typically 20 %), so that releases up to this level will be relatively smokeless and non-luminous. To achieve further reductions in smoke, steam injection at rates of 0.3 kg steam/kg gas may be provided. This will cover a large proportion of releases in a typical plant, but variations on this sizing basis may be dictated by considerations of the number and type of upstream process units, the type and probability of major release contingencies, and the cost of the flare facilities.

Normally an overcapacity line to an elevated flare is provided to handle the excess flow when the flaring rate exceeds the capacity of the multijet flare. The overcapacity flare is usually not equipped with steam injection, and smoke formation is accepted during infrequent operations. The overcapacity line and flare is designed to handle the entire maximum flow so that it can spare the multijet flare when the latter is shut down for maintenance.

The recommended minimum spacings for heat releases between 90 MW and 300 MW are given below. For heat releases outside this range, a special study must be made.

1. Clearance from property lines: 60 m.

2. Clearance from structures higher than the flare stack: 60 m. In addition, no structure where personnel access may be required while the multijet flare is in operation shall exceed in height of a projected diagonal line from the base of the flare stack to the top of the stack wall diametrically opposite.

3. Clearance from structures lower than the flare stack: 45 m.

4. Clearance from refinery roads and pipe bands: 22.5 m.

5. Clearance from the elevated overcapacity flare must comply with radiant heat spacing requirements for elevated flares, considering personnel exposure when maintenance work is being performed on the multijet flare and the

overcapacity flare is taking full design load. This clearance should not, however, be less than 50 m.

Stack Design and Dimensions for Multijet Flares

Dimensions - The inside diameter of the stack is based on the rate of heat release at design capacity. The following empirical equation can be used to calculate this dimension :

$$D = 0.47\sqrt{Q} \tag{1}$$

where D is the stack inside diameter in meters (m) and Q is the heat release at the design flaring rate, MW (Based on lower heating value).

For stack diameters up to 7.5 m the stack height is normally 15 m above grade. The bottom of the stack is elevated to allow air for combustion to enter. The minimum clearance between the bottom of the stack and grade is either 1.8 m or 0.3 D, whichever is greater.

Insulation - The steel shell of the stack is lined for its entire length with monolithic type castable refractory.

Windbreaker - A windbreaker is necessary to prevent the wind from extinguishing the flames. It serves also to hide the flames. Since a solid wall produces undesirable eddies, a louvered type is used. The windbreaker is octagonal and is placed 2.4 m from the stack. The height should be at least 0.6 m more than the stack clearance. The slats should be at least 225 mm wide and overlap by at least 50 mm. They should slope 45° to direct the flow of air downward on the inside of the enclosure. Four access doors are provided in the windbreaker, equally spaced around the periphery.

The area below the stack and windbreaker is paved with concrete, surrounded by a 200 mm curb, and graded to a central drain point from which a drain line is routed to a manhole in a vented section of the oily water sewer. The water inlet should be sealed and the manhole should be located at least 15 m from the windbreaker.

Burner Design and Back Pressure for Multijet Flares

The "burner" consists of a large number of small burner jets, arranged in a grid pattern inside the stack, near the bottom. To increase the rangeability of the burners, a two-stage piping system is provided, such that one set of burners handles low flows and the second set cuts in as the gas flow rate increases.

Staging Design - A single-stage multijet burner has a turndown ratio (maximum smokeless flaring rate/minimum flaring rate) of about 10:1. A two-stage burner therefore has a maximum turndown ratio of 100:1. In an actual two-stage design, the first stage is designed to handle about 20 percent of the design load of the multijet flare. Although this reduces the overall turndown ratio to about 50:1 it permits a large overlap between the maximum burning rate of the first stage and the minimum burning rate of the second. The overlap is desirable for smooth operation. The first stage burner grid should extend full width with the tube ends near the pilot to more readily promote ignition at very low flaring rates.

A simple sequential water seal system with two seal drums is used to control the distribution of flare gas to the two stages and to provide flashback protection, as illustrated in Figure 4. The seal in the overcapacity line is designed to start releasing when the pressure in the second stage burner header reaches a value corresponding to maximum design flow to the multijet flare. Design of the seal drums, loop seals, disposal of effluent seal water, etc., follows the procedure described below for elevated flares, with the following exceptions:

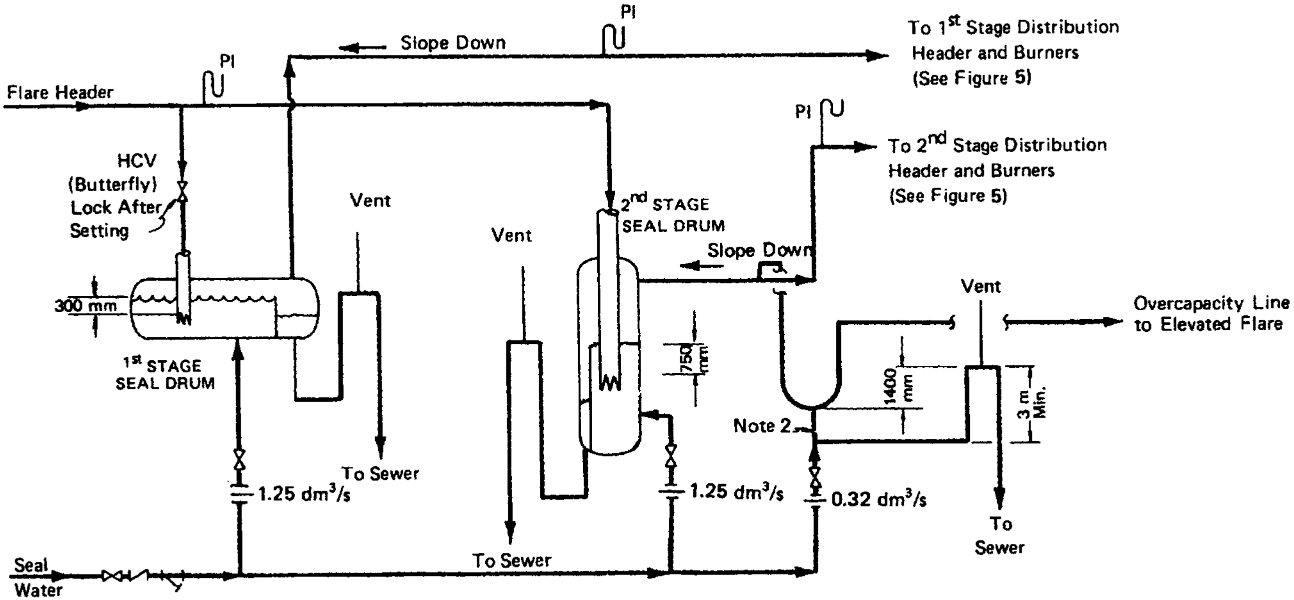
a. Vapor inlet dipleg submergence is selected to control progressive operation of the stages. Typical values are indicated in Figure 4.

b. The second stage seal drums should be vertical, to minimize its size, provided that the required slope up to the burners can be achieved.

A butterfly valve in the line to the first stage seal drum limits the maximum flow to the first stage burner. The valve is set by observing the burners while flaring at design capacity. Once adjusted, the valve should be locked in position.

Pressure indicators are specified to aid in making adjustments and in evaluating burner performance. Water manometers may be used in this service, but must be shut off when not in use.

Piping to Burners - First and second stage piping and headers, as well as the burner lines themselves, are sized to minimize pressure drop and velocity effects. Thus, maldistribution of flow to the burners will be minimized. The burner lines are fabricated from standard 100 mm pipe, and are arranged in a split grid layout with distribution headers and split feed lines on opposite sides, for both first and second stage burners. First and second stage headers must be sloped so that any condensate will drain back to the seal drums. However, the burner lines must be accurately installed in a horizontal plane.



Notes:

1. Locate drums as close as possible to windbreaker.
2. Design for 0.12 m/s to allow vapor disengaging.
3. Also see Figures 8 and 11 for ancilliary features.

Figure 4. Typical flare seal drum arrangement. For use on multijet or staged elevated flares.

Size, Spacing and Number of Burner Jets - The jet nozzles are 380 mm lengths of 25 mm stainless steel pipe. They discharge vertically from the horizontal burner lines, which run across the bottom of the stack. The jet nozzles are not insulated.

The number of jets is based on gas velocity. For 25 mm standard pipe, the recommended maximum velocity permits a flow rate of 72.2 m^3/h (actual or standard) of gas per jet. The following empirical equation can be used :

$$N = 0.013\ V \tag{2}$$

where: N = Number of jets (N should be rounded off upward to a whole number).

V = Flare design capacity, m^3/h (for a flare system, actual and standard cubic meters are virtually equivalent).

The jets are laid out on an approximately square pitch, with spacings in both directions varying between roughly 450 mm and 600 mm. A first approximation of the required spacing is obtained from Eq. (3)

$$P_1 = P_2 = 833\ \sqrt{D^2/N} \tag{3}$$

where: P_1 = Center-to-center spacing of adjacent burner lines, mm

P_2 = Center-to-center spacing of jets along a burner line, mm

D = Inside diameter of stack (from Equation 1), m.

This equation is based on a negligible wall effect. The design values of P_1 and P_2 must be determined from a scale drawing, which is made to allow the required number of jets to be installed in the available area. This area is restricted by the limitation that no jet should be placed closer than 300 mm from the inside wall of the stack. The spacing is also affected by air flow considerations , which may require the layout to be modified.

Air Flow - The capacity of a multijet flare to induce air flow must be calculated, to make sure that it is adequate to meet the maximum air flow requirement for smokeless combustion. (W_a of Equation 4 below must be $\geq W_r$ of Equation 5). The term *air flow capacity* refers to the primary air flow rate which will be induced around each jet, and may be estimated from the following equation :

$$W_a = 2.02 \times 10^{-6} A_j [H_l (1-289/T_S)]^{0.50} / K [1+10.9(A_o/A_B)^2+0.138]^{0.50} \quad (4)$$

where: W_a = Available primary air flow per jet, kg/s

A_j = Net free flow area per jet, mm^2 = $P_2 (P_l - d)$

d = Outside diameter of burner line (including insulation), mm

H_l = Net stack height, m

T_S = Average stack gas temperature, °K. This temperature depends on the heating value of the flare gas and the percent excess air. It may be assumed that T_S = 1538°C = 1811°K, which is a reasonable approximation and is further justified by the fact that Equation 4 is relatively insensitive to changes in T_S.

A_o = Total free flow area for primary air, m^2

A_B = Peripheral area between bottom of stack and grade, m^2

$K = (1- A_o/A_S)^2$

A_S = Total inside cross-sectional area of stack, m^2

The primary air flow rate per jet necessary for smokeless combustion depends on the molecular weight and degree of unsaturation of the flare gas. Experience indicates that it varies linearly with percent unsaturates, from a minimum of 20 % excess air for a flare gas containing 0 % unsaturates to 35 % excess air for a gas containing 67 mol % unsaturates. Based on this relationship and a gas flow rate of 72.2 m^3/h per jet, the required primary air flow rate can be computed directly from the gas composition, or approximated conservatively from the following equation :

$$W_r = 1.26 \times 10^{-4} [120 M + 56 y_u + 22 M y_u + 300] \quad (5)$$

where: W_r = Required primary air flow rate, kg/s per jet, for a gas flow rate of 72.2 m^3/h per jet.

M = Molecular weight of flared gas

y_u = Mole fraction unsaturates in flared gas.

If $W_r > W_a$, it will be necessary to modify the burner layout in order to provide more air flow area per jet. If necessary, the stack diameter must be increased. (Another possibility is reducing the piping diameter, if pressure drop is not limiting.)

The value of W_r, calculated from the gas composition or from Equation 5 should be considered to be a minimum requirement. A multijet flare should be designed with a calculated air capacity W_a as high as possible, as limited by practical limitations of economics and geometry.

Flameholder - Flameholders are necessary to prevent the flame from "riding" up to the top of the stack. They provide a surface at which burning can take place and also promote better mixing of air and gas by the additional turbulence which they cause above the jets. Construction is simply a solid, 25 mm diameter rod of refractory material (silicon carbide) supported horizontally above each burner line. The bottom of the rod should be 13 mm above the tips of the jets.

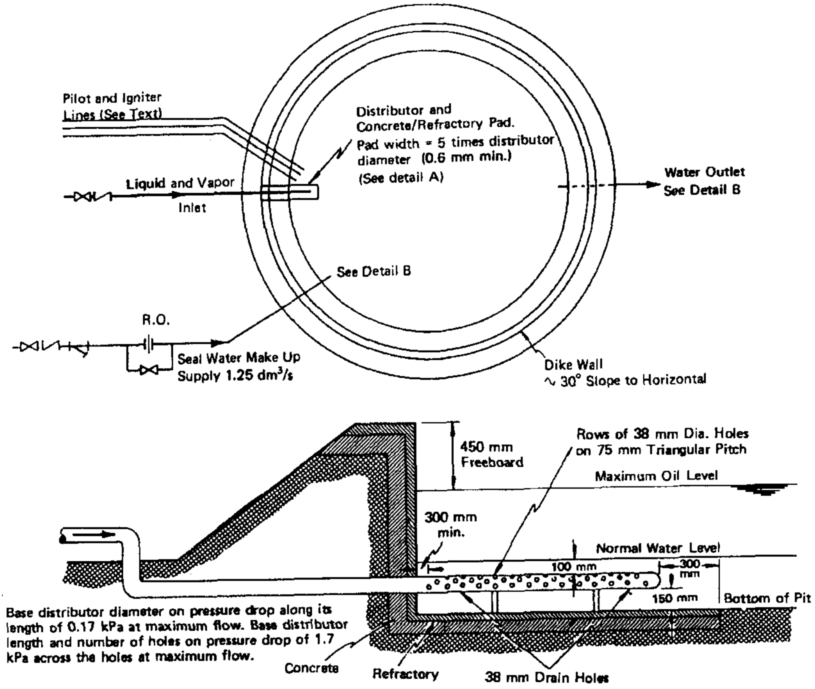
Critical Burner Dimensions - The position of the flameholder's and burner lines relative to the bottom or the stack is critical for efficient operation. For example, the multijet flare has a turndown ratio of 10:1 when the flameholder centerline is 125 mm below the bottom of the stack but only 2:1 when it is 150 mm above the bottom of the stack.

Burner Insulation - The burner lines should be insulated with a 25 mm layer of castable insulation.

Pilots and Igniters - Duplicate continuous gas pilots are required at each side of the flare, corresponding to the split burner grid layout. Selection of pilot and igniter systems follows the guidelines described below, and the controls should be located 15 m to 30 m from the windbreaker.

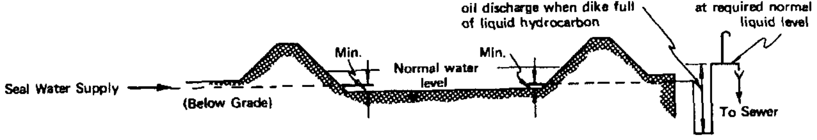
Because of the potential hazard of release of unignited hydrocarbons at ground level, a flame scanner with alarm in the control house is included for each pilot. The flame scanner must be located so that interference of ultra violet rays from the main flame or other sources do not cause false readings. Ultraviolet detectors should be mounted such that they are looking straight down through the pilots toward the ground. The installation should also provide strainers in each gas or oil line to pilots.

Steam Injection - While the multijet flare will achieve a significant reduction in the smoke produced, it does not provide true smokeless combustion over its full operating range. This is particularly true with the heavier (C_4+) and unsaturated gases. Steam injection at a rate of about 0.5 kg steam per kg of gas will provide an additional reduction in smoke for most gases. Steam should be



Base distributor diameter on pressure drop along its length of 0.17 kPa at maximum flow. Base distributor length and number of holes on pressure drop of 1.7 kPa across the holes at maximum flow.

DETAIL A
LIQUID AND VAPOR DISTRIBUTOR



DETAIL B
SEAL WATER INLET AND OVERFLOW SEAL

Figure 5. Typical burning pit flare design.

injected between the seal drum and the flare burner grid.

Burning-Pit Flares

Burning-pit flares can handle flammable liquids or gases or mixtures of the two. A typical design is shown in Figure 5. A circular pit is illustrated, but any convenient shape may be used. The burning pit is simply a shallow earth or concrete surfaced pool area enclosed by a dike, with a liquid/vapor inlet pipe through the wall, and provided with pilots and igniters. While the design basis outlined below is adequate for handling emergency releases, a more conservative approach is recommended for continuous flaring services, incorporating up to twice the calculated pit area.

Burning Pit Flare Sizing - The burning-pit area is sized to provide sufficient surface to vaporize and burn liquid at a rate equal to the maximum incoming liquid rate. The calculation procedure is as follows :

Step 1. Determine the linear regression rate of the liquid surface (i.e., the rate at which the liquid level would fall as a result of vaporization by radiant heat from the burning vapor above it, assuming no addition of incoming liquid):

$$R = 1.27 \times 10^{-9} Q/H_v M \tag{6}$$

Step 2. Determine the pit area necessary to vaporize and burn liquid at a rate equal to the liquid input rate :

$$A = 103 M/ R\rho_L \tag{7}$$

where: Q = Total heat released by combustion of incoming vapor and vaporized liquid, W

A = Pit area required to vaporize and burn liquid, m^2

M = Rate of vaporization and burning of liquid, kg/hr (selected as equal to the rate of flashed liquid entering the pit)

R = Linear regression rate of liquid surface, mm/s

ρ_L = Liquid density, kg/m^3

H_v = Heat required to vaporize liquid, MJ/kg

Step 3. The dike wall height above the water level is selected to provide holdup capacity for the largest liquid release resulting from a single contingency during 30 minutes, plus 450 mm freeboard. The liquid rate is based on the actual flashed liquid entering the pit, assuming no burning or further vaporization in the pit. The height of the dike wall above the water level should not, however, be less than 1.2 m.

Spacing of Burning Pit Flares - Spacing is based upon radiant heat considerations at maximum heat release. The following simplified calculation procedure can be used :

$$D = 0.0316\sqrt{FQ/4\pi K} \qquad (8)$$

where: D = Distance from flame to point under consideration, m

Q = Total heat released by liquid and vapor, W

K = Radiant heat density, kW/m^2

F = Fraction of heat radiated. Refer to the last section of this chapter for some appropriate values of F.

The center of the flame is assumed to be 1-1/2 pool diameters from the center of the pool at a 45° elevation, in the direction of the point where radiant heat density is being considered. This assumption is used to allow for flame deflection by wind. Permissible radiant heat densities for personnel are described in the last section of this chapter. In some special cases, spacing may be governed by radiant heat exposure of certain vulnerable items of equipment, rather than personnel. Radiant heat density at the property line must not exceed 1.6 kW/m^2.

In addition, the following minimum spacings apply to burning pit flares:

150 m from property lines, roadways, or any process or storage facilities.

60 m from any source of ignitable hydrocarbons, such as separators or floating roof tanks.

Valves in the inlet, seal water and pilot gas lines should be located according to permissible radiant heat densities for personnel (refer to the last section of this chapter for specific guidelines). Piping to the burning pit should be suitably protected against flame impingement (e.g., by installation below grade).

Inlet Piping and Flashback Protection - In non-freezing climates, for services where incoming liquid or vapor hydrocarbon temperatures do not fall below 0°C, flashback protection is provided by submerging the inlet below a layer of water in the bottom of the pit. The top of the inlet distributor is 100 mm below the water level, and the bottom of the distributor is 150 mm above the bottom of the pit. A continuous flow of at least 1.3 dm^3/s of makeup water into the pit is provided, with the level maintained by an overflow seal discharging to the oily water sewer. Higher water rates may be required on account of ground seepage or evapoaration during maximum release conditions.

For services where ambient or inlet temperatures may fall below 0°C, flashback protection is provided by a special seal drum or loop seal in the inlet line. This equipment is designed specifically for the particular liquid and vapor materials being flared. In these cases a 150 mm minimum water layer is included in the bottom of the pit to prevent oil seepage into the ground, and the hydrocarbon inlet distributor is mounted 150 mm above the water surface. Details of the inlet distributor are shown in Figure 5.

Pilots and Igniters - Two gas-fired pilots with igniters are installed adjacent to the inlet distributor. The igniter assembly and pilot gas valves must be located remote from the flare for protection of personnel and equipment. This restricts igniter selection to the forced air supply type. Location of these components should be such that the calculated radiant heat density at maximum load does not exceed permissible levels for personnel exposure. Because of the potential hazard of release of unignited hydrocarbons at ground level, a flame scanner (suitably shielded and aircooled and connected to an alarm in the control house), is provided for each pilot.

Flare Pilots and Igniters

All flares must be provided with continuous pilots to ensure combustion of any releases discharged to them, and to prevent flame-out from occuring. Various designs of pilot burner are available, and proprietary tips for elevated flares are normally provided complete with pilots.

In addition, an ignition system is required for igniting the pilots when a flare is commissioned. Various proprietary systems are available, but they all function by the application of an electrical spark to a gas/air mixture in an ignition chamber, and allowing the resulting flame front to travel through an igniter tube which terminates with an open end adjacent to the pilot burner tip. In some designs, to improve the reliability of achieving ignition, the igniter tube ignites a raw gas on-off pilot burner provided with a special wide-angle tip, which in turn ignites the continuous pilot.

Air to the ignition chamber may be inspirated through a gas-operated venturi, which is designed to provide the correct gas/air mixture, but this type does not function reliably when horizontal igniter tube runs exceed 6 to 9 m. The ignition chamber and igniter tube venturi must therefore be placed close to the flare; i.e., at the base of an elevated flare or adjacent to the windbreaker of a multijet flare. Remote operation of the ignition controls may however be achieved by locating the spark ignition pushbutton and the igniter gas and pilot gas valves at any required distance from the flare. The inspirating system is not, however, acceptable for burning-pit flares since, even with remote control, the ignition chamber and venturis, etc, would be subject to damage by overheating. Typical designs include an inspirating type or are based on the use of a compressed air supply (from the instrument air header) for the ignition chamber, the gas/air mixture being controlled by restriction orifices. In this latter arrangement, the flame front passes through an igniter tube as for the inspirating type, but there are no limitations on horizontal distance from the flare, and the igniter assembly may therefore be located remotely. The forced air supply type of igniter system is preferred for all flare applications.

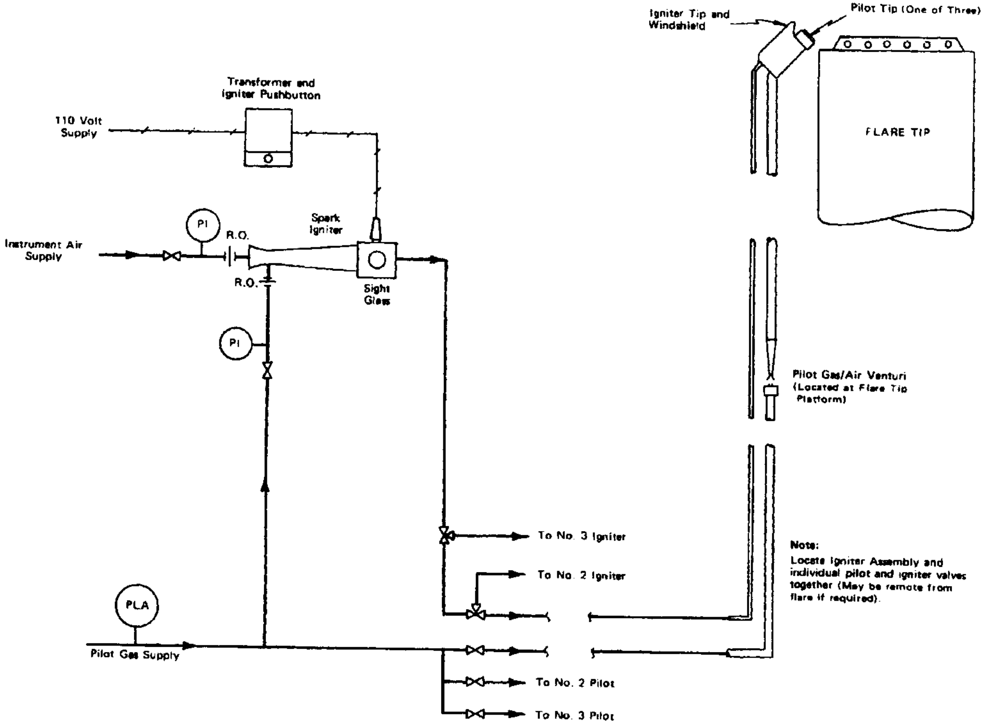
When pilot and igniter systems for flares are being selected, available proprietary systems may be considered but the choice should be based upon proven satisfactory operating experience. A design based on a flame front generator, which uses an instrument air supply to the ignition chamber has been among the most frequently applied igniter system in designs. It is illustrated in Figure 6.

Gas pilots must be provided with a reliable source of gas which will remain available under any single contingency such as power or air failure. Pilots should be designed for the anticipated gas composition, which should be reasonably constant.

Flashback Protection for Flare Systems

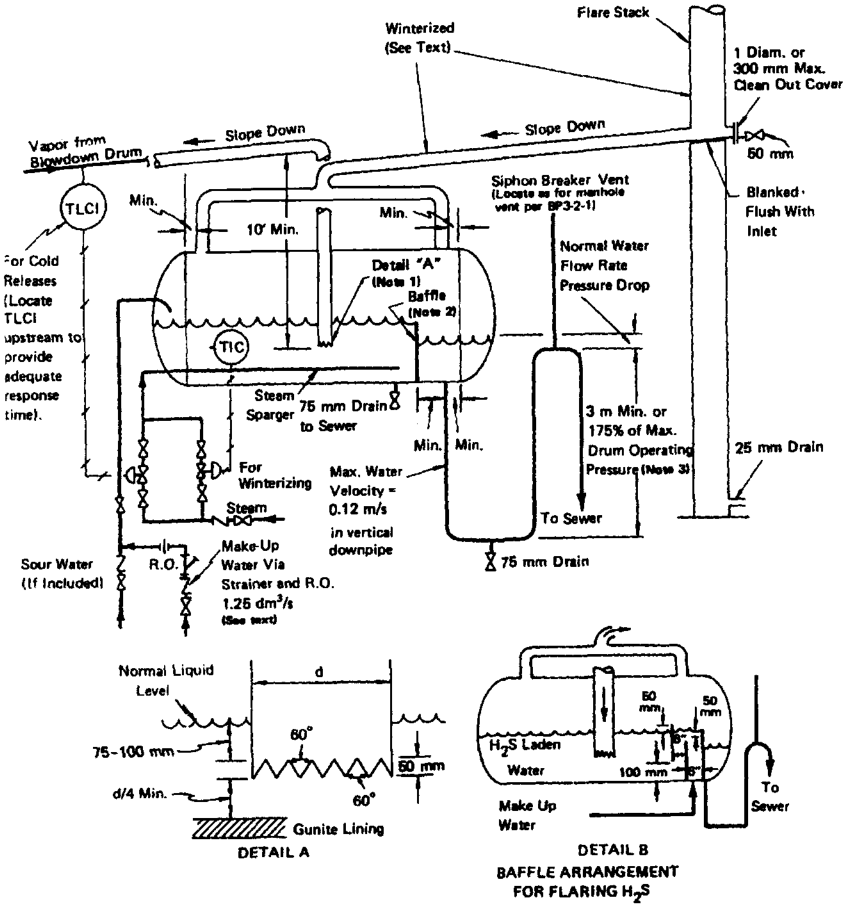
All flares must be provided with flashback protection to prevent a flame front from travelling back to the upstream piping and equipment. A number of different flashback seal designs are available, of which the seal drum is used in nearly all applications. Key design details are summarized below:

The Seal Drum - A typical flare seal drum for an elevated flare stack is illustrated in Figure 7. A baffle maintains the normal water level, and the vapor inlet is submerged 75 mm to 100 mm. Drum dimensions are designed such that a 3 m slug of water is pressured back into the vertical inlet piping in the event of



Note: Locate Igniter Assembly and individual pilot and igniter valves together (May be remote from flare if required).

Figure 6. Typical elevated flare pilot and igniter.



DETAIL A

DETAIL B
BAFFLE ARRANGEMENT
FOR FLARING H_2S

Figure 7. Typical flare seal drum.

flashback, thus preventing the explosion from propagating further upstream.

The vapor space is sized to avoid water entrainment in the flare gas. As a rule, vapor velocities in the drum should not exceed 150 % of critical. This however can be increased to 230 % V_c (V_c, critical velocity) when considering a remote contingency. The critical velocity can be calculated from the following equation :

$$Vc = 0.0479 \sqrt{(\rho_L - \rho_v)/\rho_v} \tag{9}$$

where: Vc = Critical vapor velocity in the drum, m/s

ρ_L = Density of entrained liquid, kg/m^3

ρ_v = Density of vapor, kg/m^3

The seal drum designs described include a serrated edge at the base of the dipleg, as shown in Detail A of Figure 7. Intermittant bubbling through the seal dipleg sometimes causes problems in flame pulsation, low frequency noise and/or smoke pollution. An alternative to this is to design successive diplegs plus the application of a non-pulsing sparger. Seal water should be taken from a reliable source. Salt water should not be used where flare gas components would cause deposition of solids. Winterizing should be provided if required.

The vertical downflow section of the water outlet line from the drum is sized for a maximum velocity of 0.12 m/s, to allow entrained gases to disengage. The seal loop should be sized for the normal water flow of 1.3 dm^3/s. This is an arbitrary rate and is intended to reestablish the seal within a reasonable time after each flaring incident. No specific time limit is intended for resealing since the termination of a flaring incident is of necessity gradual as time is required for natural depressuring of the flare headers. Also a finite time is needed before air would again appear deep down into the flare stack. Where higher water rates are considered necessary then facilities for manually increasing the make up should be provided. The seal depth is equivalent to 175 % of the maximum drum operating pressure, subject to a minimum of 5 m. A seal depth of 110 % of the maximum operating pressure is permitted when applying the 1.5 times design pressure rule to remote contingencies.

The drum is normally gunite-lined for corrosion control and must be designed to meet all the following combinations of extreme temperature and pressure requirements:

1. Internal flashback explosion, i.e., 1055 kPa design pressure at normal

operating temperature (i.e., when no emergency releases to flare are occurring). Experience of actual flashbacks in flares has shown that a vessel with this design pressure can withstand such internal explosions. This design pressure must also be applied to the flare stack and the water outlet seal loop.

2. A design pressure of 345 kPa at the highest possible operating temperature of the entering vapors resulting from a single contingency.

3. A design pressure of 345 kPa at the lowest possible operating temperature of the entering vapors resulting from a single contingency.

For cases (2) and (3), credit may be taken for heat transfer to the atmosphere from the flare header upstream of the seal drum.

A flare seal drum may also be used as a sour water disengaging drum, if economically advantageous. In such cases, special care should be given to ensure that the drum is adequately sized to simultaneously meet all design features required for both functions. Also a separate source of makeup water must still be provided to ensure continuity of the seal.

The drum is usually equipped with steam injection if required for winterizing or cold releases. Refer to Figure 7 for some of the details. If winterizing is necessary, then the steam should be temperature-controlled in order to maintain the seal water temperature at 4 to 10 °C. It is important to note that the drum should be located at a minimum safe distance from the flare.

Y-Leg Seal - The Y-leg seal, which is illustrated in Figure 8, is used for elevated flares in applications where there is no possibility, under any process or ambient conditions, of entrainment or condensation of flammable liquids in the section of the flare header between the blowdown drum and the flare. It is assumed that the blowdown drum is adequately designed to minimize entrainment.

An important design detail is that the inlet line should be sloped at least 30° from the horizontal and the diameters of the vertical and sloped legs of the Y-seal are sized such that a 3 m slug of water is pressured back up the sloping inlet line, without spilling over into the header, in the event of flashback. Note that this requires an enlarged diameter section in the vertical leg of the Y-seal. The seal is maintained by a continuous flow of water at 1.26 dm^3/s and the water makeup is provided with steam injection, if required for winterizing or cold releases.

The 300 mm water seal is maintained by the location of the overflows which discharge to the oily water sewer through the primary seal loop. The water

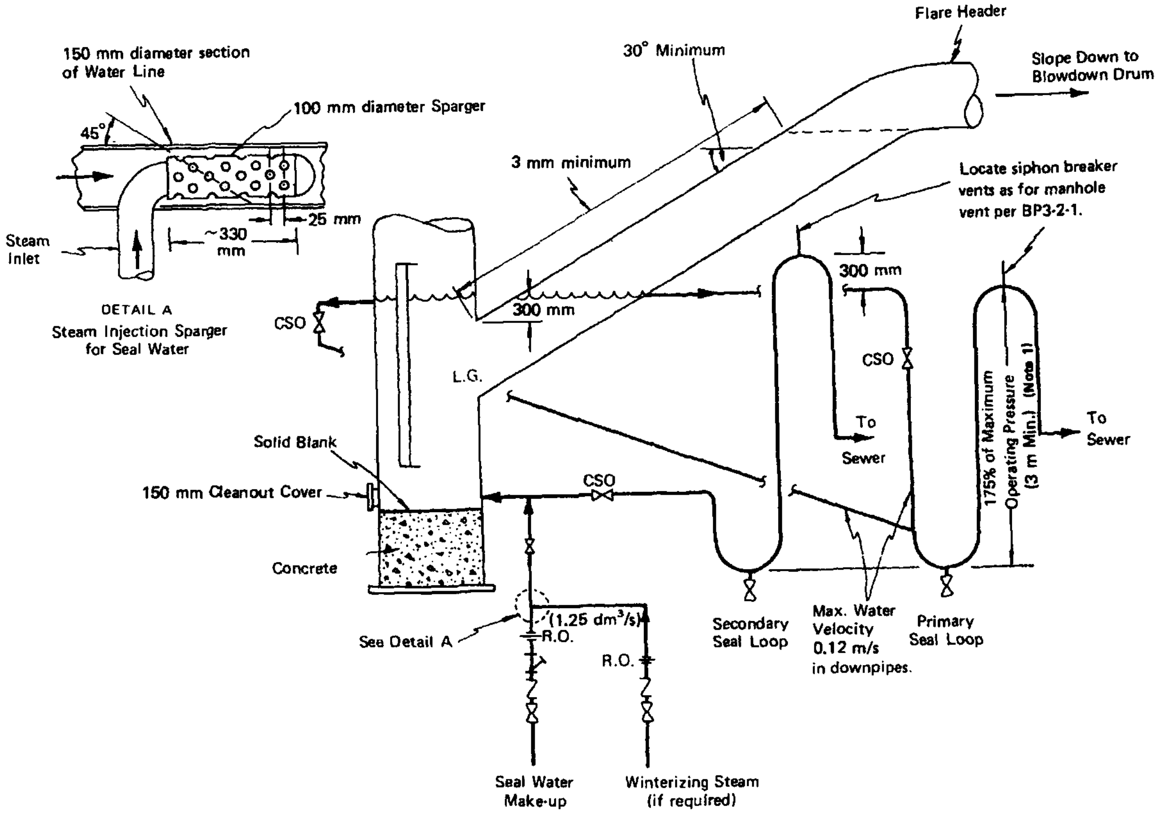


Figure 8. Typical elevated flare with Y-leg seal.

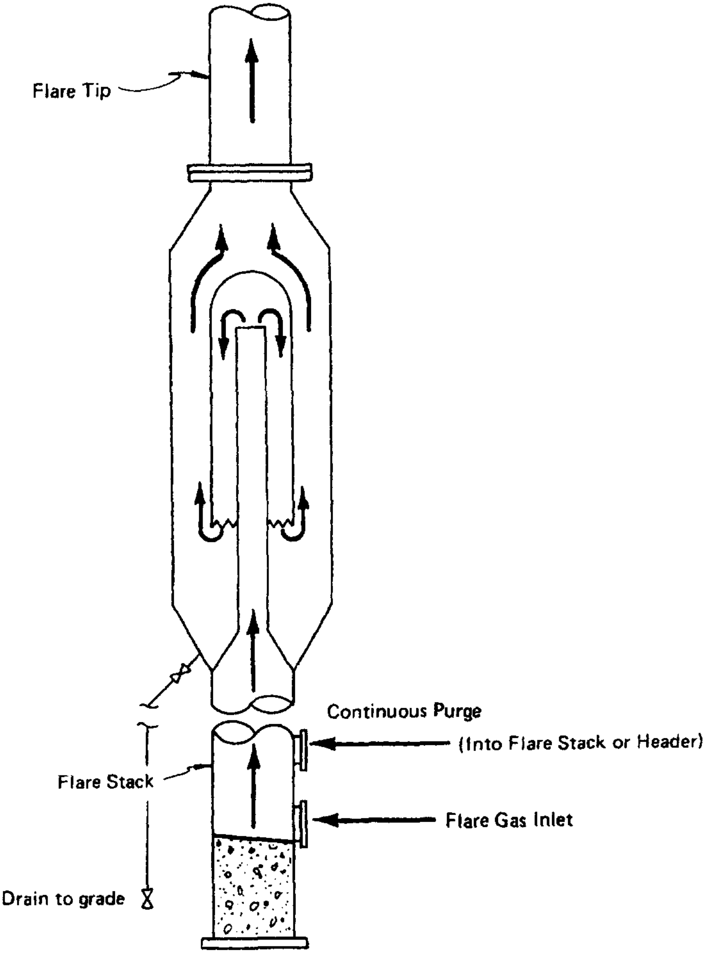


Figure 9. Example of a typical dry seal.

overflow lines, down to the bottom of the primary seal loop, should be sized for a maximum velocity of 0.12 m/s to disengage entrained gases. The height of the primary seal loop should be equivalent to 175 % of the pressure at the base of the stack during a maximum blow, or 3 m, whichever is greater. The overflow line from the downstream side of the Y-seal should be tied into the base of the primary seal loop. A seal depth of 110 % of the operating pressure is permitted when applying the 1.5 times design pressure rule to remote contingencies.

A secondary seal loop is provided for water withdrawal during major blows when turbulence at the downstream overflow connection to the primary seal loop interferes with normal drainage. Extending the base of the flare stack 3 diameters below the sloped inlet line provides vapor disengaging for the secondary seal leg. The bottom of the stack and inlet line up to 1.5 m above the seal water level are gunite lined for corrosion protection.

Gas Purge - As an alternative to a water seal of the drum type or Y-leg type, a continuous hydrocarbon gas purge may be used to provide protection against internal burning or flashback explosion which might result from air backing down from the flare tip at zero or low flaring rates. Application of this method is limited to cases where a continuous low-cost gas supply is available, where there are no public relations or atmospheric pollution objections to continuous flaring, and where problems that would be associated with a water seal (e.g., freezing up by cold vapors) can thereby be avoided. The purge gas must be non-fouling and the source must be reliable and assured during any single contingency such as power or air failure.

Dry Seals - Various proprietary dry seals are available, an example of which is provided in Figure 9. A dry seal may be used in conjunction with a gas purge system of flashback protection when the purge gas is lighter than air. The dry seal functions by trapping a volume of the light gas in the internal inverted compartment, thus preventing air from displacing light gas in the flare stack by buoyancy effects. A continuous purge is still necessary to prevent air diffusion down the stack, although the purge flow need be only 10% of that required for a simple open stack. A typical seal may require approximately 30-40% of the open stack purge rate. The open stack purge rate should be determined from manufacturers literature. This reduced purge requirement is the only incentive for installing a dry seal.

In the case of heavier-than-air purge gas, there is no buoyancy mechanism causing air entry into the stack, and there is thus no incentive to include a dry seal. Unlike a water seal, a dry seal cannot prevent a flashback from traveling upstream if a combustible mixture has been formed by the entry of air into the safety valve or flare headers. It only protects against internal burning flashback which might result from air backing down from the flare tip at zero or low flaring rates. Dry seals are therefore not normally specified for new designs. The location of a molecular seal recommended by manufacturers is approximately 3.6 m below the top of the flare.

Disposal of Seal Water - Effluent water from water seals must be routed to safe means of disposal, considering possible hazards arising from liquid or vapor hydrocarbons or toxic materials that may be entrained or dissolved in the water. Seal water should be discharged as follows:

1. If H_2S is never present in the flare gas, seal water effluent should be routed through an open funnel (to permit checking of seal water flow) to a manhole in a vented section of the oily water sewer. The water inlet to the manhole must be sealed.

2. If H_2S in any concentration is intermittently present in the flare gas, seal water effluent should be routed through a closed connection (incorporating a sight glass for checking flow) to a manhole in a vented section of the oily water sewer. The water inlet to the manhole must be sealed. In addition, the seal drum should include a baffle (refer to detail B in Figure 7) to preferentially route the makeup water to the sewer, while confining H_2S-saturated water to the drum.

However, pollution considerations may make the routing of seal water to the sewer unacceptable, in which case disposal must follow method (3) below.

3. If H_2S is continuously present in the flare gas or if the flare seal drum also functions as a sour water disengaging drum, then the effluent seal water must be routed to a sour water stripper, desalter, or other safe means of disposal. Withdrawal from the drum is by pump in place of the normal loop seal arrangement. Two pumps are provided; one motor driven for normal use, and the other having a steam turbine drive with low pressure cut-in. The seal drum level is controlled by LIC with high and low alarm lights plus an independent high level alarm.

4. Disposal of effluent water from multijet ground flare seal drums should comply with paragraphs (1), (2), and (3) above, except that:

In category (1), closed piping with a sight glass should be used rather than an open funnel.

In categories (1) and (2), the manhole receiving effluent seal water should be located at least 15 m from the windbreaker.

The reader should note that flame arresters are not permitted as a means of flashback protection in flare systems.

Reducing Flare Pulsing and Noise

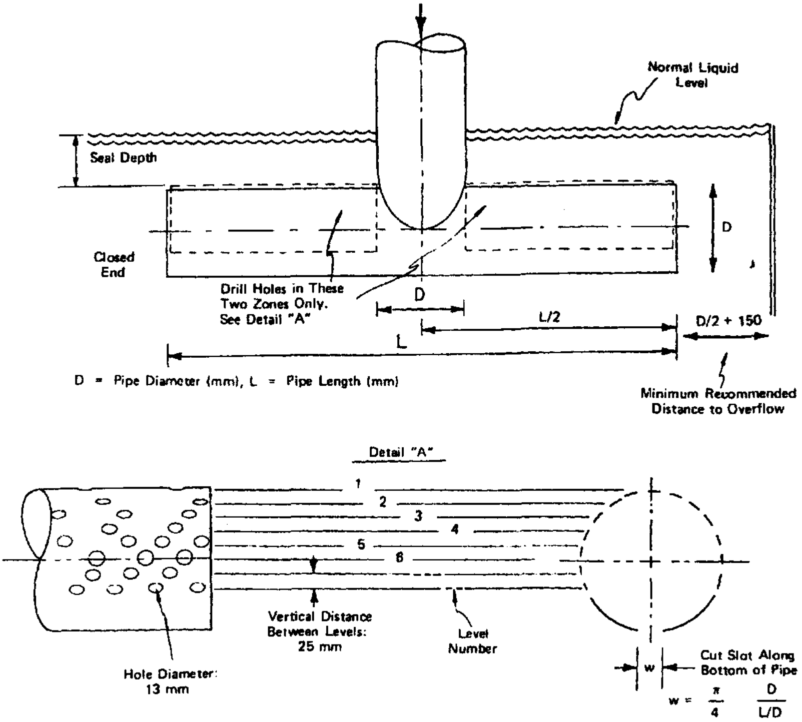
A major cause of pulsing in flare systems is flow surging in the water seal drum. One of several reasons why it is important to eliminate pulsing is to reduce flare noise. Combustion flare noise has been shown to increase as the steam rate increases. Since the amount of steam required to suppress smoke in a flare is set by the flaring rate, flow surges will require a higher steam rate than for a steady flow.

At much higher steam rates than required to suppress smoke, flares occasionally exhibit another type of noise problem, a low frequency rumble that can be sensed in the surrounding community as vibrations. It is theorized that this is produced by portions of the flare gas containing too much steam to burn with a stable flame. By reducing the steam requirement of a flare through eliminating flow surges, there is a greater operating margin between smokeless flaring and high steam rate associated rumble.

Other advantages of eliminating pulsing are reduced steam costs, more accurate flow measurements possible with a steady flow, and reduced incidence of blowing and seal liquid. Also, the size of the seal drum can sometimes be reduced.

Flow surges in the seal drum are likely generated by the cyclic formation of large bubbles as the flare gas is discharged into the drum. These pulsations can be virtually eliminated by the use of a horizontal sparger incorporating many small diameter holes arranged specifically to allow the open area to increase as flow increases. These holes must be spaced sufficiently far apart to avoid interference between bubbles.

An example of a distributor design is shown in Figure 10. Hole density is low at the top of the pipe and is increased lower on the pipe. The maximum open area density of about 10% assures reasonable bubble formation in this design. The average velocity out of the top row of holes starts at about 40 m/s and increases as the pressure rises and total flow increases. Total areas of holes plus bottom slot should be equal to at least two times the cross sectional area of the inlet pipe.



$$\text{Total Number of Levels} = \frac{3D}{(L/D) - 1} + 2$$

Hole Spacing Required on Each Level					
Level Number	1	2	3	4	5 or more
Hole Spacing (mm) Center-to-Center	75	75	50	50	40
Holes/Level per Meter of Pipe	13	26	40	40	50

Note: Minimum Recommended L/D = 7

Figure 10. Typical non-pulsing sparger design.

Flaring Practices for Hydrogen Sulfide

Continuous reliase of H_2S is not permitted in the United States under EPA regulations. Where permitted by the government and local management the following criteria are applicable:

Continuous releases of concentrated H_2S streams must be segregated in a separate flare system to limit the extent of fouling and plugging problems. Releases of H_2S such as diversion of sour gas product to flares during shutdown or upset of a downstream sulfur recovery unit are considered to be continuous, but safety valve releases are not included in this category. However, if a special H_2S flare system is provided for continuous releases, the concentrated H_2S safety valve releases should be tied into it rather than into the regular flare system. Due to the nature of H_2S one should plan on frequent inspection and flushing of H_2S flares to remove scale and corrosion products.

Preferably, the H_2S flare system should consist of a segregated header and separate line routed up the side of a conventional elevated flare stack, sharing the same structure, pilots and igniters. However, the H_2S header may be tied into the regular flare seal drum if there are special mechanical design problems associated with the separate stack; e.g., in the case of a flare which is to be dismantled for overhaul. Flare elevation must be sufficient to meet atmospheric pollution and ground level concentration requirements for the sulfur dioxide produced.

Odor and pollution problems are often experienced due to incomplete combustion when concentrated H_2S is flared. When such considerations are expected to be critical, flare system designs should include a fuel gas connection and equipment sizing sufficient to handle an equal volume of fuel gas when flaring H_2S at design rate. Where flaring of H_2S is intermittent and the fuel gas diluent is continuous then steam injection at the base of the flare may be needed to reduce smoking.

Flashback protection is required for H_2S flaring systems, either by water seal or continuous gas purge. If a water seal is used, special requirements apply to the disposal of the effluent seal water. In the case of an H_2S flaring system handling a flow of H_2S which in uninterrupted throughout the period that a plant is in operation, and which stops only when the producing plant is shutdown, then flashback protection is not required. However, steam or inert gas connections are required to permit purging the flare line startup and shutdown.

Spare Flare Capacity Considerations

Flare capacity must always be available for emergency releases that may occur from the upstream plant facilities which are in operation. Flaring rates corresponding to the design contingency basis of the facilities should be assumed. Thus, when a flare is taken out of service for maintenance, either spare capacity must be available in other flares connected to the same system, or the necessary upstream process units must be shutdown. The justification of spare flare capacity is therefore considered on the basis of an economic evaluation of the debits associated with plant downtime.

As a means of providing spare flare capacity without rendering additional plot area unavailable for plant equipment, a double elevated flare arrangement is permissible. This consists of two flare stacks in a common supporting structure, each sized for the full design flow. Either stack can be dismantled and removed for maintenance while the other is in service. Appropriate platforms and radiation shielding are installed to provide safe working conditions for personnel during the dismantling and reassembly stages, but the mechanical design should be such that this work can be carried out without requiring access above an elevation of 15 m.

Flare Gas Metering

Metering flare gas is important for loss accounting and for control of steam injection. A special requirement for flare gas meters is low pressure drop and the ability to continue functioning in fouling conditions. The flare gas metering methods listed below have been used with varying degrees of success.

Dall Flow Tube - The advantage is this type of flowmeter is that it has a permanent head loss of only 5 % of the measured pressure differential. This is the lowest pressure drop of all orifice meter designs. Flow ratios as high as 1:10 (e.g., 1.0 to 10 kg/s) can be measured within + 2% of actual flow. Dall flow tubes are available in different materials and diameters up to 1500 mm.

Anemotherm - The Anemotherm, a hot wire anemometer, has been extensively applied to flare gas measurement and has been successfully used after modifications to reduce probe fouling. A technique used to reduce fouling is the installation of a steam coil around the tip of the probe, to prevent gases from condensing on the probe.

Pitot Probe - A simple pitot probe with "bouncing ball" flow detector gives a qualitative indication of flare gas flow and can be used in the identification of leakage sources.

Other types of flowmeters that are applicable to flare gas metering include

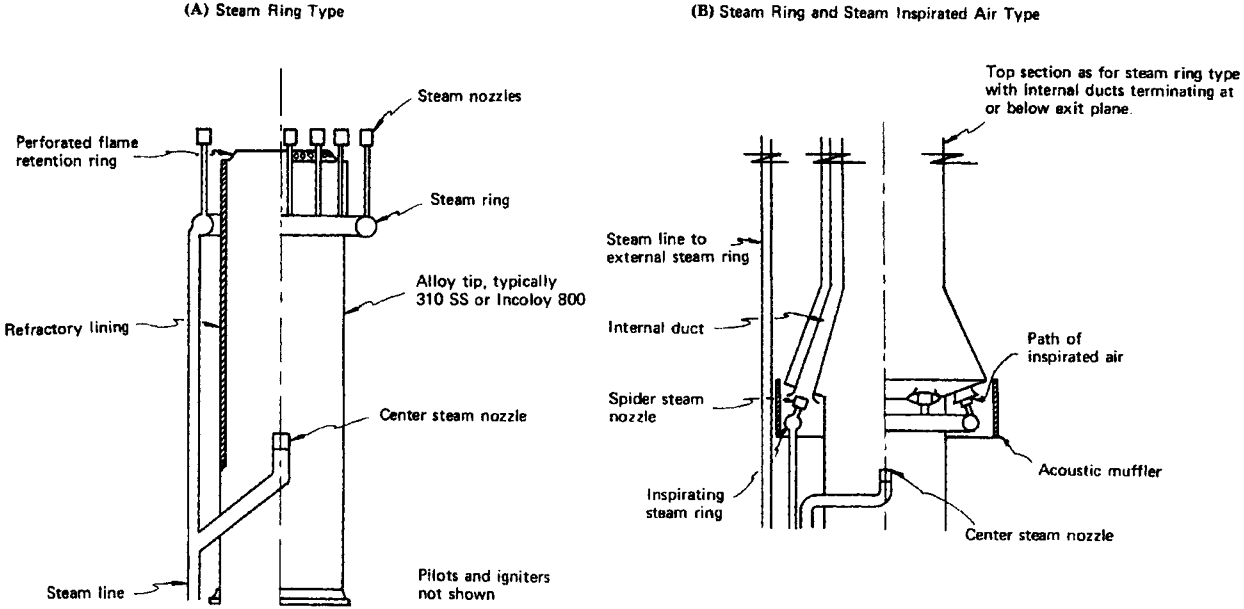
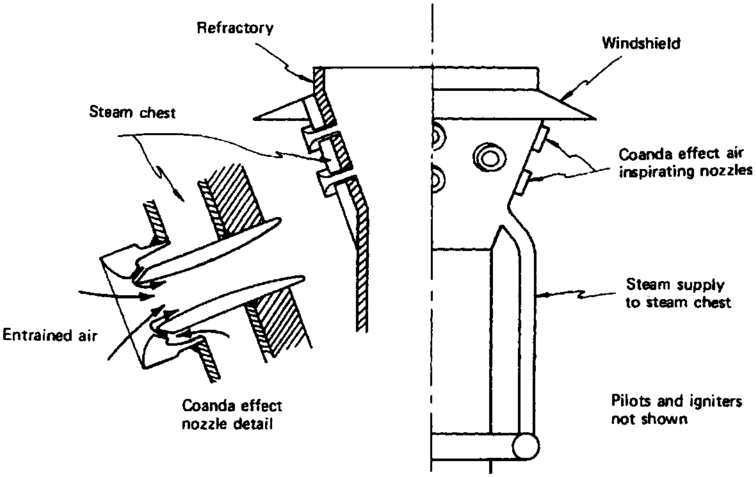


Figure 11. Smokeless elevated flare tip.

(C) Steam Inspirated Air Type



(D) Steam Inspirated Air Type

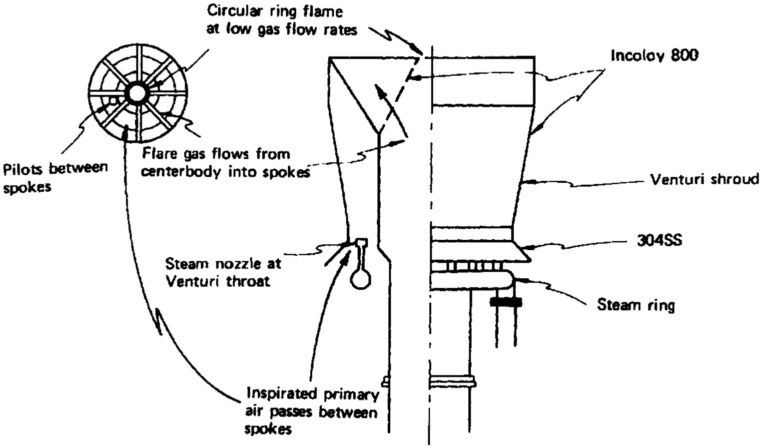


Figure 11. Continued - Other typical smokeless elevated flare tip designs.

fluidic probes, vortex shedding flowmeters, and ultrasonic noise generation flowmeters.

Protection Against Low Ambient or Flare Gas Temperatures

Flare systems must be protected against any possibility of partial or complete blockage by ice, hydrates, solidification, etc. Seal Drums and Y-seals requiring winterizing should be provided with temperature-controlled steam injection to maintain the seal water temperature at 4 to 10°C. This limits the quantity of water vapor entering the flare stack.

When winterizing is required, the steam tracing and insulation should include the first 7.5 m of the flare stack above the vapor inlet; and also in the case of a drum seal, the vapor line from the seal drum to the flare. Where steam injection to one of the smokeless tips , as shown in Figures 11 B and C is used, then the steam ring should remain outside the top of the flare tip (i.e., not internal). Where severe ambient conditions are encountered then it is recommended that the entire seal drum and flare be insulated in addition to steam tracing and open steam injection at base of flare.

RADIATION AND SPACING FROM FLARES AND IGNITED VENTS

The following procedures for determining flame shape, flame center and radiant heat density at any adjacent location are applicable to flares, and to vents (such as atmophseric safety valve discharges) which may inadvertently become ignited. To properly assess radiation effects and to incorpoarte them into the design, the following information is required to be gathered by the designer:

Atmospheric Conditions - Normal average wind speed, ambient air temperature, air density, relative humidity.

Flare Gas - Molecular weight, lower explosive limit, density at flare tip, fraction F of heat release radiated by the flame.

Flare Parameters - Heat release rate, tip diameter, gas discharge velocity, stack height.

A simplified graphical procedure is presented for the determination of X_c (horizontal) and Z_c (vertical), the flame mid-point coordinates for flares and ignited vents.Figures 12 and 13 provide graphs, one for X_c and one for Z_c, which

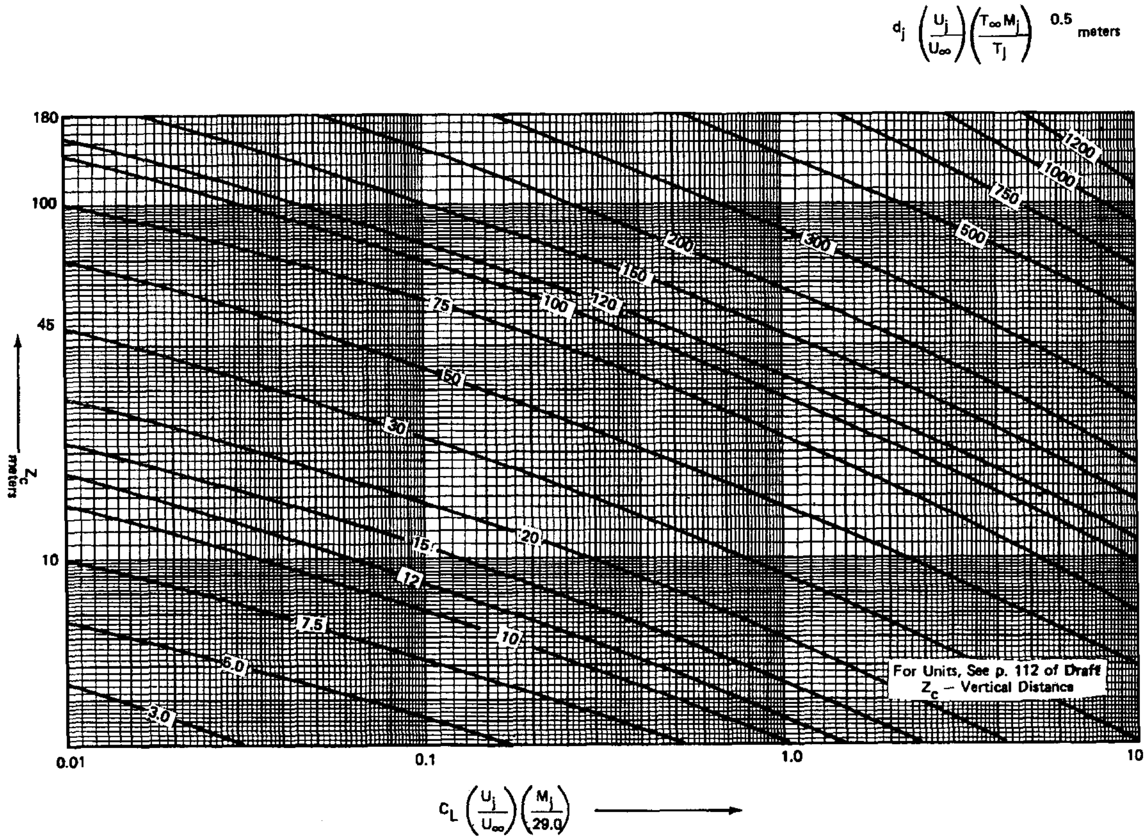


Figure 12. Flare center for flares and ignited vents.

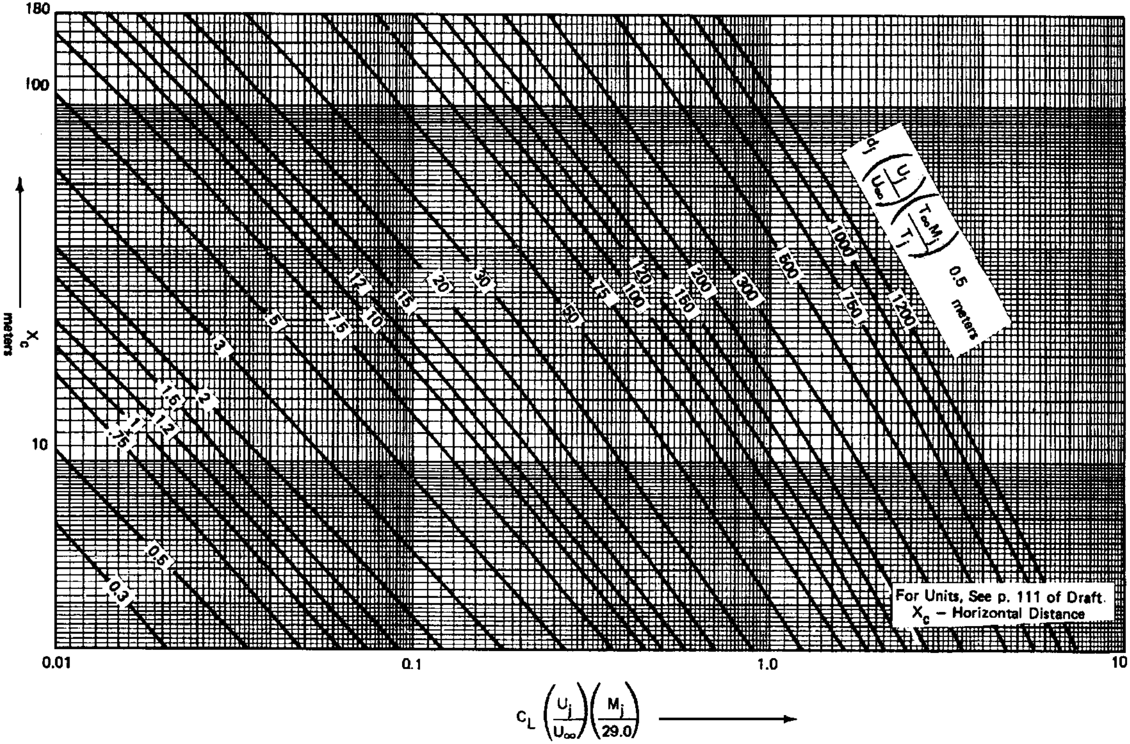


Figure 13. Flame center for flares and ignited vents.

are based on two independent variables: $\overline{C_L}$ and a modified form of d_jR. The latter variable has been modified to include gas and air temperatures and molecular weights instead of densities. The ideal gas law is assumed in the foregoing procedure. The specific information required by the designer is:

C_L = Lower explosive limit concentration of flare gas in air, volume fractions

d_j = Inside diameter of flare tip, m

M_j = Molecular weight of flare gas

M_∞ = Molecular weight of air = 28.97

U_∞ = Normal average wind speed, m/s, based on local meteorological conditions

u_j = Flare gas velocity at the flare tip, m/s

ρ_j = Flare gas density at the flare tip (before ignition), kg/m^3

ρ_∞ = Ambient air density, kg/m^3. If it is not known, use 1.22 kg/m^3

The following is the calculation procedure that can be used:

Step 1: Calculate the concentration parameter $\overline{C_L}$ for the flare gas from the following relationship:

$$\overline{C_L} = C_L\,(u_j/U_\infty)\,(M_j/M_\infty) \tag{10}$$

Step 2 : Using the computed value from Equation 10, determine Z_c from the chart given in Figure 12.

Step 3 : Using the computed value from Equation 10, determine X_c from the chart given in Figure 13.

These coordinates locate the flame center, which is treated as the source of all radiation from the flame.

The following is an alternate calculation procedure to obtain flame shape and length.

Step 1 : Calculate the concentration parameter, $\overline{C_L}$ for the flare gas from Equation 10.

Step 2 : Calculate the value of $\overline{S_L}$ from Equation 11 (a) or 11(c) given below. This term is a dimensionless coordinate of the concentration C_L along the centerline of the flame. From the value of $\overline{S_L}$, calculate the downwind coordinate $\overline{X_L}$ from Equation 11(b), or obtain it from Figure 14, which is a plot of the flame coordinates in dimensionless form.

a. If $\overline{C_L} \leq 0.5$, then

$$\overline{S_L} = 2.04 / (\overline{C_L})^{1.03} \tag{11-a}$$

and

$$\overline{X_L} = \overline{S_L} - 1.65 \tag{11-b}$$

(Refer to Figure 14 for an understanding of these relationships).

b. If $\overline{C_L} > 0.5$, then

$$\overline{S_L} = 2.52 / (\overline{C_L})^{0.625} \tag{11-c}$$

i. If $\overline{S_L} > 2.35$

$$\overline{X_L} = \overline{S_L} - 1.65 \tag{11-b}$$

ii. If $\overline{S_L} \leq 2.35$, obtain X_L from Figure 14. This figure is a plot of the following equation:

$$\overline{S_L} = 1.04 \ \overline{X_L}^{\ 2} - 2.05 \ \overline{X_L}^{\ 0.28} \tag{11-d}$$

The term X (Equation 11b) to locate the tip of the flame.

Step 3 : Calculate the dimensionless rise $\overline{Z_L}$ of the flame tip above the flame tip:

$$\overline{Z_L} = 2.05 \ \overline{X_L}^{\ 0.28}$$

Step 4 : Calculate the scaling parameter R, which accounts for the effect of flame shape of the relative thrusts of the wind and the gas jet discharging from the flare tip:

$$R = \left(\frac{U_j}{U_\infty}\right)\left(\frac{\rho_j}{\rho_\infty}\right)^{0.5} \tag{13}$$

Step 5 : Calculate the full-scale coordinates of the flame tip:

$$Z_L = \overline{Z_L}\,\mathrm{dj}\,R \tag{14-a}$$

$$X_L = \overline{X_L}\,\mathrm{dj}\,R \tag{14-b}$$

where: Z_L = Vertical distance of flame tip above flare tip, m

X_L = Horizontal distance of flame tip from flare tip, m

Step 6: Calculate the location of the flame center, which is treated as the source of all radiation from the flame. Only flames bent over by the wind are considered, since for nearly vertical flames (calm air) the effective center of flame radiation is higher off the ground and therefore not limiting for spacing purposes.

$$Z_C = 0.82 Z_L \tag{15-a}$$

$$X_C = 0.5 X_L \tag{15-b}$$

Table 3. Limits of flammability of gases and vapors, percent in air:

Gas or Vapor	Lower	Upper
Hydrogen	4.00	75.0
Carbon monoxide	12.50	74.0
Ammonia	15.50	26.60
Hydrogen sulfide	4.30	45.50
Carbon disulfide	1.25	44.0
Methane	5.30	14.0
Ethane	3.00	12.5
Propane	2.20	9.5
Butane	1.90	8.5
Iso-butane	1.80	8.4
Pentane	1.50	7.80
Iso-pentane	1.40	7.6
Hexane	1.20	7.5
Heptane	1.20	6.7
Octane	1.00	3.20
Nonane	0.83	2.90
Decane	0.67	2.60
Dodecane	0.60	....
Tetradecane	0.50	....
Ethylene	3.1	32.0
Propylene	2.4	10.3
Butadiene	2.00	11.50
Butylene	1.98	9.65
Amylene	1.65	7.70
Acetylene	2.50	81.00
Allylene	1.74	....
Benzene	1.4	7.1
Toluene	1.27	6.75
Styrene	1.10	6.10
o-Xylene	1.00	6.00
Naphthalene	0.90	....
Anthracene	0.63	....
Cyclo-propane	2.40	10.4
Cyclo-hexene	1.22	4.81
Cyclo-hexane	1.30	8.0
Methyl cyclo-hexane	1.20	....
Gasoline-regular	1.40	7.50
Gasoline-73 octane	1.50	7.40
Gasoline-92 octane	1.50	7.60
Gasoline-100 octane	1.45	7.50
Naphtha	1.10	6.00

Estimating Lower Flammable Limits

Explosivity limits for various pure components are given in Table 3. The limits of flammability (concentration, C) for a mixture of gases can be computed from the following expression:

$$\frac{1}{C_L} = \frac{Y_1}{C_{L2}} + \frac{Y_2}{C_{L2}} + \dots + \frac{Y_n}{C_n} \tag{16}$$

where: C_L = Lower explosive limit concentration of the mixture in air

y_1 = Mole fraction (or volume fraction) of component 1 in the mixture

y_n = Mole fraction of the nth component

C_{L1} = Lower explosive limit concentration of component 1 in air

C_{Ln} = Lower explosive limit of the nth component.

Equation 16 is accurate for mixtures of paraffinic gases or for mixtures of H_2, CO and CH_4. It is only approximate for mixtures of H_2 and C_2H_4, H_2 and C_2H_2, H_2S and CH_4 or CH_4 and $C_2H_2Cl_2$. It is even less accurate for mixtures of flammable gases with steam or inerts. Nevertheless, for the accuracy required here, this simple equation should be tried first in all cases. When the flare gas is a mixture of flammable gases and inerts, the simplest way to approximate C_L is to use the above equation with C_L taken to be equal for the inert components. This step treats the inerts as a simple diluent in an ideal mixture. It would give only a rough approximation for mixtures such as H_2 and H_2O vapor, where a component which is inert in the cold mixture nevertheless enters into the reactions taking place in the flame. In such cases, if a rough approximation of C_L is not good enough, then more accurate methods described below should be used.

Application of this procedure to inadvertently ignited safety valve discharges can involve a special problem. Certain combinations of pressure ratio and length of safety valve riser can result in choked flow, with a pressure discontinuity at the exit. The pressure of the jet then adjusts to atmospheric pressure in a system of shock waves or expansion waves over a distance of a few pipe diameters. These waves can affect the local mixing of the jet with the crosswind. Since the calculation procedure incorporates correlations for subsonic jets, it cannot be expected to be entirely accurate in this case. Nevertheless, since the wave system occupies a very small portion of the flow field influenced by the jet, the procedure can still be counted on to provide a useful approximation of the gross flame length and flame shape when the actual discharge velocity and diameter are used in the calculation.

Credit for additional height of the flame center for multiple valve installations may be taken by clustering the safety valve discharge pipes to the atmosphere. The following procedure should be used for determining equivalent diameter and exit velocity to be used in the flame center calculation. Diameter and velocity are based on the total actual area of the clustered vents.

For Equal Diameter Vents -

$$D_{jequ.} = \left[4*(\text{ sum of vent areas })/\pi\right]^{0.50} = d*n \tag{17a}$$

$$V_{jequ.} = \text{actual velocity of any one vent.} \tag{17b}$$

For Unequal Diameter Vents :

$$D_{jequ.} = \left[4*(\text{ sum of vent areas })/\pi\right]^{0.50} = \sqrt{d_1^2+d_2^2+\cdots+d_n^2} \tag{18a}$$

$$U_{jequ} = \text{Total Gas Rate / Total Vent Area} = \frac{1.273*10^{-3}*Q}{(D_{jequ.})^2} \tag{18b}$$

In the above expressions, the terms are defined as follows:

d, d_1, d_2, d_n = Diameter of individual vents, in meters
D_j = Equivalent vent diameter to be used in flame calculation, in meters
n = Number of vents in cluster ($n > 3$)
Q = Total gas rate of all valves, in dm^3/s at one atmosphere and the release temperature
$U_{j\,equ.}$ = Equivalent vent exit velocity to be used in flame calc., in m/s

Let's now turn attention back to the flammability limit itself. When small increments of a combustible gas are successively mixed with air, a concentration is finally attained in which a flame will propagate if a source of ignition is present. This is referred to as the Lower Flammable Limit (LFL or lower explosion limit) of the gas in air. As further increments of the gas are added, a higher concentration of flammable gas in air will finally be attained in which a flame will fail to propagate. The concentration of gas and air just as this point is reached is referred to as the Upper Flammable Limit (UFL or upper explosion limit, UEL) of the gas in air.

Safety requires that only the most reliable experimentally determined flammable limit data be considered in purging calculations. This is included in Table 3.

Below atmospheric pressure there is no effect on the limits of flammability of natural gas-air mixtures and most other gas-air mixtures. Below about 25 mm absolute pressure, carbon monoxide-air mixtures are not flammable.

From atmospheric pressure, up to 2170 kPa, the lower limit of flammability is not affected, but the upper limit rises as the pressure on the mixture is increased. This widens the limits of flammability as the pressure increases, as shown in Table 4. Above 2170 kPa the lower limit will be reduced.

Table 4. Effect of Increase in Pressure on Raising the Upper Flammability Limit

	Gage Pressure kPa	0	500	1000	1500	2000	2500
Methane and Natural Gas	Approx.	15	17	20	25	31	42
Coce Oven Gas	Approx.	31	35	44	57	72	-
Multipliers*		1	1.2	1.5	2.0	2.6	3.7

*For estimating the upper limit of gases other then those given when the limits at 0 gage pressure are known, apply the multipliers indicated in the table to the percent range in flammability, adding the new range to the lower limit at 0 gage pressure to find the new upper limit.

It is necessary, in approaching most problems to obtain the limits of flammability of the particular gas mixture in question. Few industrial fuel gases are composed of pure gases or vapors, but are mixtures in most cases of many different gases. If the equipment and time are available, a number of mixtures of the fuel gas-air mixtures may be prepared and their flammabilities tested by ignition but it is much easier to determine the flammability limits of complex gas mixtures by calculation. Experience has shown that the results obtained are sufficiently dependable.

A calculation of the flammability limits of complex gas mixtures is carried out by the application of the mixture rule. Stated simply, the mixture rule is that if two limit mixtures of different gases are added together, the resulting mixture also will be a limit mixture. The mathematical statement of this law is as follows:

$$C_L = \frac{1}{\sum (Y_1/C_{L2} + Y_2/C_{L2} + Y_3/C_{L3} + \cdots + Y_n/C_n)} \qquad (19)$$

where Y_1, Y_2, Y_3, etc. , are the proportions of each combustible gas present in the original mixture, free from air and inerts so that $Y_1 + Y_2 + Y_3 + \ldots Y_n = 100$, and C_{L1}, C_{L2}, C_{L3} ... C_{Ln}, etc., are the lower limits of flammability of the mixture. A similar procedure would be applied to determine the upper limit of flammability.

An example of the application of this law is given for natural gas having the following composition:

Gas	% By Volume	%Gas in Air Lower Limit
Methane	80.0	5.00
Ethane	15.0	3.10
Propane	4.0	2.10
Butane	1.0	1.86

$$\text{Lower Limit} = \frac{100}{\frac{80.0}{5.00} + \frac{15.0}{3.10} + \frac{1.0}{1.86}} = 4.30\% \text{ Gas in Air}$$

Any oxygen contained in a mixture may be considered as though it were a part of the air required for the combustion, and the analysis of the flammable mixture should be converted to an air-free basis before the flammable limits are calculated.

When mixtures contain appreciable amounts of nitrogen and carbon dioxide, calculation of the flammability limits becomes more complicated and requires the use of an extension of the mixture rule. In this modified method, the inert gases are taken into consideration by assuming that the original mixture is composed of a number of submixtures of inert gas-combustible gas, the flammability limits of which have been experimentally determined in a similar manner as have the limits for the pure gases as given in Table 3. Figures 14, 15, and 16 provide the flammability limits of mixtures of some common gases present in fuels with CO_2, N_2 and H_2O.

An illustration of the application of this modified mixture rule is given in Table 5. In this example, a producer gas has the composition shown at the top of the table. The CO_2 and N_2 may be apportioned with the different combustibles in any of several ways, two of which are represented by calculations A and B in Table 5.

In these examples, the inerts CO_2 and N_2 are combined with the combustibles H_2 and CO and the small amount of CH_4 is taken alone. Next, the ratio of inert to combustible is obtained for each group, as shown in Table 5, and the flammable limits for each such mixture are obtained from Figure 14. The mixture rule formula is now applied, using the data as just obtained, and the limits are calculated as shown in Table 5.

The summary at the bottom of Table 5 indicates the relative agreement between the calculated data and that experimentally determined for this particular producer gas. It is suggested that the difference between calculated and determined data in this case may be due more to inaccuracies in the analysis of the producer gas (particularly for methane) than to the fault of the mixture rule formula. This points up the fact that reliable gas analyses also are a necessary part of the calculated flammability limit data.

Calculation of Radiant Heat Flux

The following calculation procedure is used to predict the heat flux K incident on a surface normal to the direction of radiation, at any distance from the flare stack and any elevation above grade. The information required for this procedure is as follows:

x_c = Downwind distance from flare stack to flame center, m

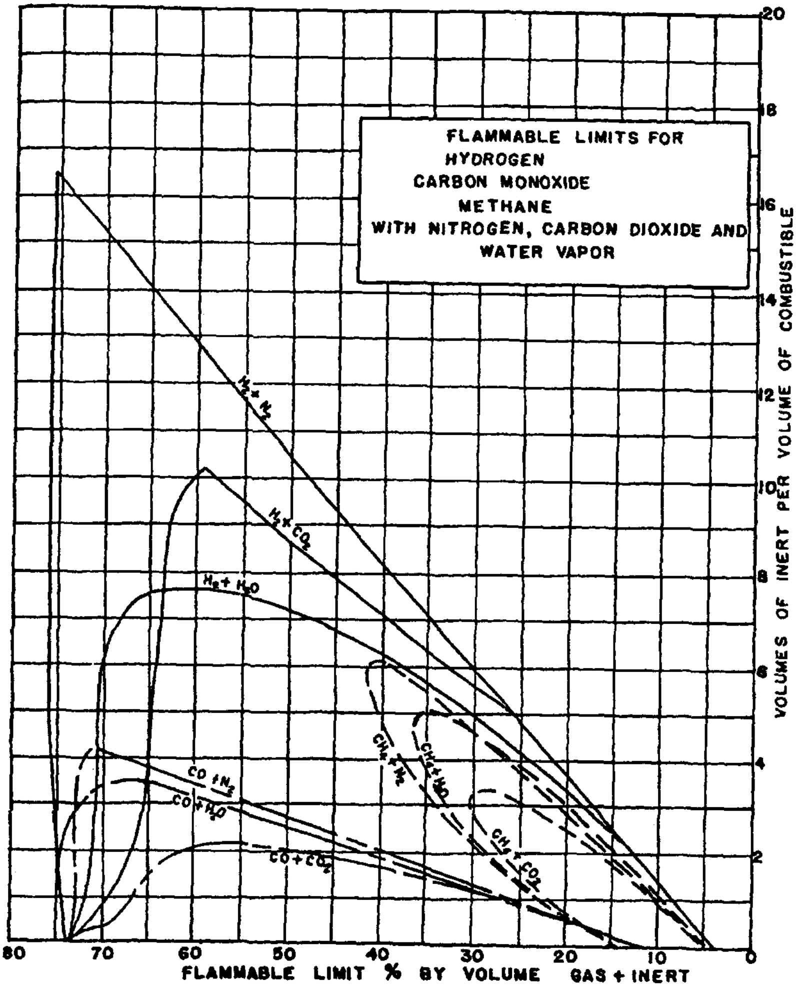


Figure 14. Flammable limits for hydrogen, carbon monoxide, methane, with nitrogen, carbon dioxide and water vapor.

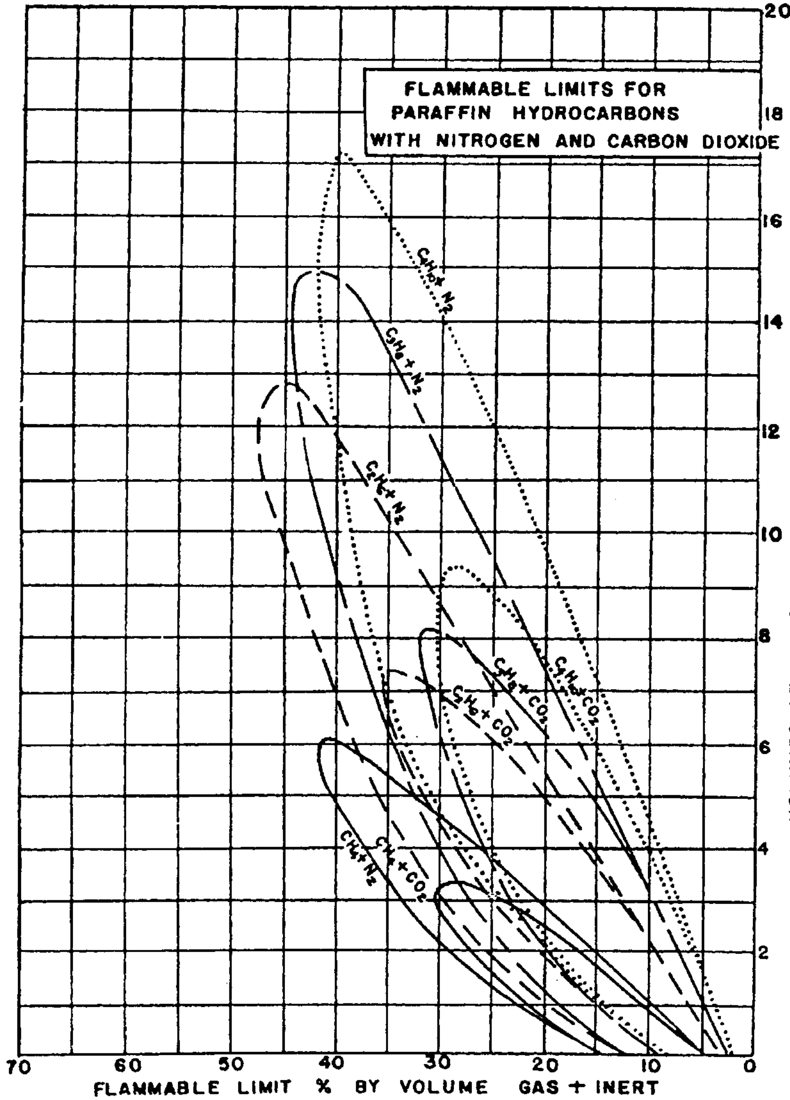


Figure 15. Flammable limits for paraffin hydrocarbons, with nitrogen and carbon dioxide.

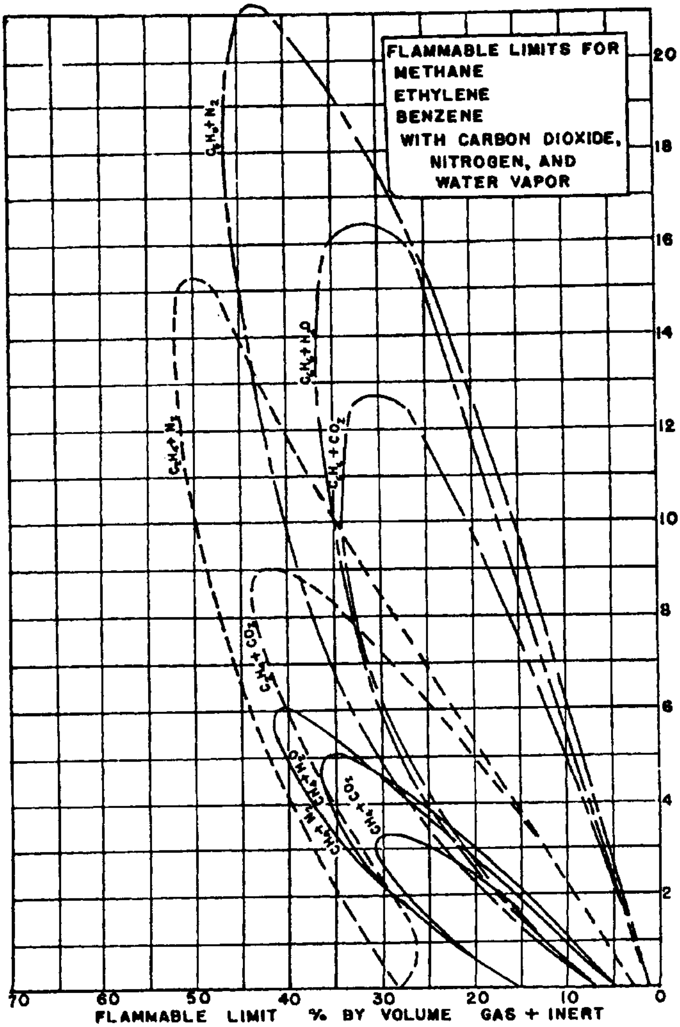
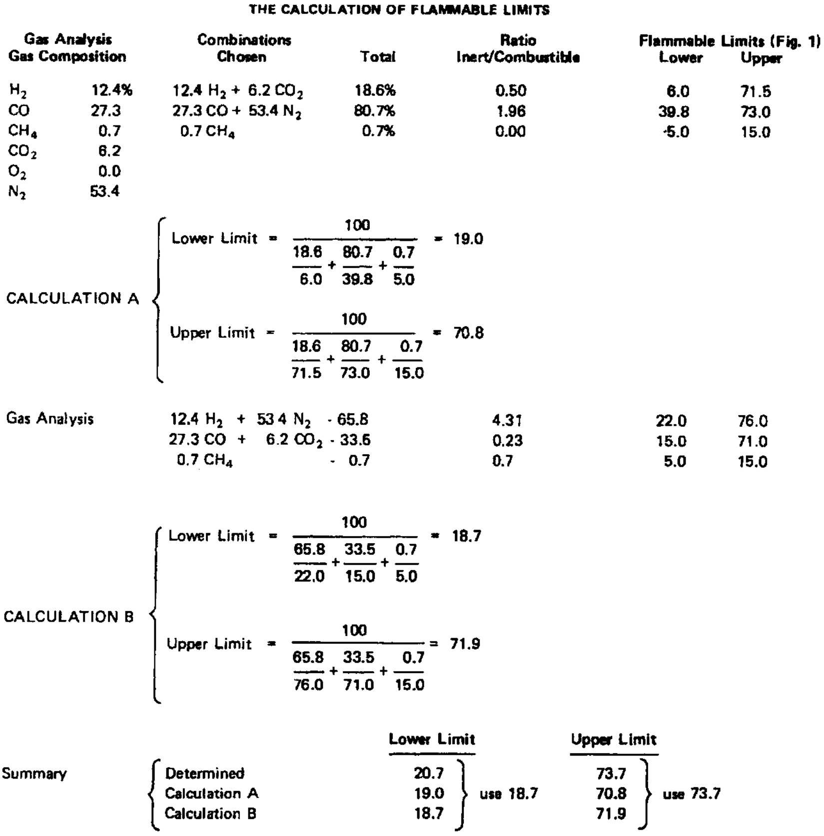


Figure 16. Flammable limits for methane, ethylene, benzene, with nitrogen, carbon dioxide and water vapor.

Table 5. The calculation of flammable limits.



THE CALCULATION OF FLAMMABLE LIMITS

Gas Analysis Gas Composition		Combinations Chosen	Total	Ratio Inert/Combustible	Flammable Limits (Fig. 1) Lower	 Upper
H_2	12.4%	$12.4\ H_2 + 6.2\ CO_2$	18.6%	0.50	6.0	71.5
CO	27.3	$27.3\ CO + 53.4\ N_2$	80.7%	1.96	39.8	73.0
CH_4	0.7	$0.7\ CH_4$	0.7%	0.00	5.0	15.0
CO_2	6.2					
O_2	0.0					
N_2	53.4					

CALCULATION A

$$\text{Lower Limit} = \frac{100}{\frac{18.6}{6.0} + \frac{80.7}{39.8} + \frac{0.7}{5.0}} = 19.0$$

$$\text{Upper Limit} = \frac{100}{\frac{18.6}{71.5} + \frac{80.7}{73.0} + \frac{0.7}{15.0}} = 70.8$$

Gas Analysis				
	$12.4\ H_2\ +\ 53.4\ N_2$ - 65.8	4.31	22.0	76.0
	$27.3\ CO\ +\ 6.2\ CO_2$ - 33.6	0.23	15.0	71.0
	$0.7\ CH_4$ - 0.7	0.7	5.0	15.0

CALCULATION B

$$\text{Lower Limit} = \frac{100}{\frac{65.8}{22.0} + \frac{33.5}{15.0} + \frac{0.7}{5.0}} = 18.7$$

$$\text{Upper Limit} = \frac{100}{\frac{65.8}{76.0} + \frac{33.5}{71.0} + \frac{0.7}{15.0}} = 71.9$$

Summary	Lower Limit		Upper Limit	
Determined	20.7	use 18.7	73.7	use 73.7
Calculation A	19.0		70.8	
Calculation B	18.7		71.9	

z_c = Height of flame center above flare tip, m

h = Height of flare tip above grade, m

F = Fraction of heat release radiated from the flame

m = Mass flaring rate, kg/s

H = Lower heating value of the flare gas, MJ/kg

r = Relative humidity, percent

The following are the calculation steps :

Step 1 : Calculate the rate of heat released Q in units of Watts (W):

$$Q = 10^6 mH \tag{20}$$

In general, $H \simeq 46.6$ MJ/kg for most paraffins, olefins and diolefins,
$H \simeq 41.9$ MJ/kg for aromatics, and
$H \simeq 120$ MJ/kg for hydrogen.

Step 2 : Select the appropriate flame emissivity factor F, based on flare gas composition. For ignited vents, lower values are recommended. The following are emissivity values reported from the literature:

Flare Gas Component	F
Hydrogen, Methane	0.2
C_2 Hydrocarbons	0.25
C_3 and C_4 Hydrocarbons	0.3
C_5 and heavier	0.4

Step 3 : Calculate the distance from the center of the flame to a position with coordinates x and z, representing personnel or equipment exposed to radiation from the flame. (Since the limiting case is the one in which the flame is blown over in the direction of the exposed personnel or equipment, they are in the same plane and a two-coordinate system is adequate for describing their relative locations). Note also that all wind directions must be considered, even in locations which have a prevailing wind direction.

$$D = \sqrt{(X-X_C)^2 + (Z-Z_C-h)^2} \tag{21}$$

where: D = Distance from the flame center to point (x, z), m

x_c = X coordinate of the flame center, m

$h + z_c$ = Z coordinate of the flame center, m

Step 4 : Calculate the atmospheric transmissivity, ξ

$$\xi = 0.79\,(100/r)^{1/16}\,(30.5/D)^{1/16} \tag{22}$$

Note that ξ is dimensionless. D and r are defined as earlier.

Step 5 : Calculate K from the following equation. K is incident on a surface normal to a line from point (x, z) to the flame center. This is the maximum value of K at that location and is the value used when personnel exposure is being considered.

$$K_{(x,z)} = \xi\, FQ/4\pi D^2 \tag{23}$$

where: K = Heat radiated to the surface, kW/m^2.

Spacing criteria based on heat radiation from flares is based on criteria established to protect personnel and to protect neighboring equipment. For personnel exposure, in most cases, heat radiation is the controlling factor in determining the spacing of elevated or burning-pit flares, considering personnel exposures at grade under maximum heat release conditions. In the case of inadvertent ignition of atmospheric vents, such as safety valve releases, higher levels of radiant heat exposure for personnel are used for determining elevation and spacing requirements. Maximum permissible flare radiation levels for personnel exposure are defined based on OSHA standards and are summarized in Table 6.

In contrast, most equipment can safely tolerate higher degrees of heat density than those defined for personnel. However, if anything vulnerable to overheating problems is involved, such as low melting point construction materials (e.g., aluminum or plastic), heat-sensitive streams, flammable vapor spaces, or electrical equipment, then the effect of radiant heat on them may need to be evaluated. When this evaluation is required, the necessary heat balance is performed to determine the resulting surface temperature, for comparison with acceptable temperatures for the equipment.

Table 6. Maximum permissible flare radiation levels for personnel.

Maximum Permissible K, kW/m^2	Conditions
9.45	Maximum value of K at design flare release at any location where personnel have access; e.g., at grade below the flare, or on a service platform on a nearby tower. Exposure must be limited to a few (approx. six) seconds, sufficient for escape only. On towers or other elevated structures, ladders must be provided on the side away from the flare, so that the tower or structure can provide some degree of shielding.
6.3	Areas where emergency actions lasting up to 1 minute may be required by personnel without shielding.
4.75	Areas where emergency actions lasting several minutes may be required by personnel without shielding.
1.6	Maximum value of K at design flare release for continuous exposure of personnel. (By comparison, solar radiation in a hot climate may be as high as 1 kW/m2).

FLARE SYSTEM SPECIFICATION

This section provides a detailed example of a *flare system specification.* This is an example of a formal design document that would be provided by the plant designer to a flare system vendor. The reader should consider this section as an example of the types of information and level of detail needed in specifying the system, however the format and organization might certainly be changed to suit the particular needs of the plant operations.

It is recommended that eleven sections be included in such a document. These sections are:

Scope - An overall or brief description of the design and requirements.

Applicable Codes, Standards and References - This information will help the supplier in his specification of design persformance and guarantees.

Technical Requirments - These will define the performance specifications for the equipment supplied.

Scope of Supply - This information will define the overall application and performance requirements

Preparation of Shipment - Defines the conditions and terms under which equipment will be deleuvered to the plant.

Vendor's Responsibility - Defines the terms of contractual commitments on the part of the vendor.

Warranty and Guarantee - Ensures that the vendor stands behind the design specification and will replace any defective parts.

An example of each of these sections follows. The discussions below may be viewed as a single specification document for a particular system, and may therefore serve as a boiler-plate example.

Scope

This specification covers the minimum requirements for the process, mechanical and structural design, materials, fabrication, shop inspection testing and supply for a flare system. The knock-out drum will be by N&P Refinery.

This specification presents the minimum requirements. The Vendor shall be deemed to be an expert on his equipment and shall be solely responsible for all aspects of design, materials, fabrication and installation instructions regardless of the requirements stated herein.

It is incumbent on the Vendor to comply with all legislation applicable to the Project. Venezuelan Government requirements, laws and custom shall prevail.

The flare system furnished as standardized or proprietary equipment shall in general follow the intent of this specification, but may differ in some details by following Supplier's standards for design conditions. However, any difference in detail shall be in accordance with established Oil and Gas Industry Standard Practice.

The requirements of this specification are to apply where they are more exacting or do not conflict with applicable local codes, or regulations.

No deviation from this specification shall be acceptable unless notified to and accepted by N&P Refinery in writing.

Applicable Codes, Standards and References

The following Codes, Standards and Specifications apply to this specification. When an edition date is not indicated for a code or standard, the latest edition and addendum in force in time of purchase shall apply.

ME-O-JBOOI Structural Engineering Design Criteria
ME-0-JDOO2 Noise Control
ME-0-JGOO2 Painting
ME-0-JDOO3 Export Packaging
ME-0-JDOO5 Welding - Shop & Field Fabricated Piping
ME-0-JFO40 Electric Motors
ME-O-JF 115 Electrical Design Requirements for Mechanical Equipment
ME-0-JN400 Instruments Furnished with Packaged Equipment
MD-0-JJOO3 Vendor Drawing and Data Submittal Requirements
ME-O-JSO 10 General Specification for Structural Steel

American Society of Mechanical Engineers (ASME)

ANSI-B 1.20.1 Pipe Threads, General Purpose
ANSI-B46.1 Surface Texture
ANSI-BI6.5 (1988) Pipe Flanges and Flanged Fittings
ANSI-B3 1.3 (1990) Chemical Plant and Petroleum Piping

American Petroleum Institute (API)

API-605 Large Diameter Carbon Steel Flanges
API RP 520 Recommended Practice for the Design (Part 1) and Installation (Part 11) of Pressure Relieving System in Refineries
APIRP521 Guide for Pressure Relieving and Depressuring Systems

National Fire Protection Association (NFPA)

NFPA 70 (1990) National Electrical Code

Precedence

In case of conflict between this specification and other specifications listed in the material requisition, the following priority of documents shall apply in descending order:

Purchase order and variations there to,
Data requisition sheets and drawings,
This specification,
Other project specifications,
Other National and International Standards, e.g., ANSI.

General Requirements

The flare shall provide safe combustion of gases up to the relief conditions shown on the data sheet without exceeding the allowable system pressure drop.

The flare shall provide smokeless combustion of gases up to the relief rates shown on the data sheet.

Equipment shall be suitable for the site conditions as identified on the data sheet.

The flare system shall be designed for zero flow to maximum flow.

The flare system with steam shall not be used.

Technical Requirements

Flare Tip Assembly

The flare tip assembly shall be furnished complete with smoke suppression system, pilot and igniter assemblies, windshield, and flame retention assembly.

The flare tip assembly shall be flange mounted on a standard pipe flange according to Specification ME-0-JDOO5. Igniter piping and pilot gas piping shall have flanged joints at the elevation of the tip mounting flange. N & P Refinery shall furnish all riser piping form the vendor's flanged connection to grade.

The flare tip assembly shall be provided with a sufficient number of adequately sized lifting attachments for proper handling of the assembly.

The flare tip assembly shall be provided with at least two pilot assemblies to assure complete ignition of gases leaving the flare tip.

Each pilot shall be spark operated complete with a windshield, pilot tip and pilot tube. It shall also have an igniter tube, thermocouple and remote venturi, if applicable.

The pilots shall be designed for stability with wind conditions up to 160 mile per hour.

Each pilot burner shall be equipped with a chrome/alumel thermocouple or similar device to sense the status of the pilot flame.

If provided, thermocouples shall be protected by inconel sheathing and shall terminate in a steel or cast iron juntion box located four (4) feet above grade at the base of the flare stack.

Design and materials of the flare tip assembly shall ensure that burning of the tip will not occur during low relief rates. The following are minimum acceptable materials:

a. Flanges at the elevation of the flare tip mounting flange shall be carbon steel.

b. The body of the flare tip and the air duct mixing section shall be high temperature austenitic. Types 309, 310, 321 stainless steel are acceptable; other materials require N & P Refinery's authorization.

c. Piping, fittings, manifolds, clips, and shields above the flare tip mounting flange but not directly exposed to flame shall be high temperature austenitic. Type 308, 310 and 321 stainless steel are acceptable; other materials require N & P Refinery's authorization.

d. Pilot burners and other materials in the flame zone shall be high temperature, high alloy steel. Proposed materials require specific authorization.

e. All materials shall be ASTM materials.

Vendor remains responsible for calculations.

The ground flare shall satisfy the maximum radiation at grade, excluding solar radiation, and shall not exceed 1500 Btu/hr/ft^2 at the edge of the sterile radius: 200 ft.

The ground flare shall not exceed radiation level of 500 BTU/hr ft^2 at third persons of the radius of 500 ft, excluding solar radiation.

Flare radiation level calculations are to be based on the worst case wind velocity, a design wind velocity of 40 Kph and a maximum of 100 Kph. Vendor shall provide radiation plots for each case.

Gas Seal

A gas seal shall be furnished if required to reduce purge gas requirements. The gas seal shall be self draining. The gas seal must be qualified by test data showing oxygen concentration below the seal at proposed purge rate and at least 50 mph simulated wind condition.

Pilot Ignition and Monitoring System

Pilot ignition shall be electric spark. One free standing pilot ignition control panel shall be provided with the necessary valves, piping and wiring to ignite each pilot on the flare tip. The panel shall be complete with gas valves, gauges and automatic ignition controls. The vendor shall specify if purge and pilot valves are in or out of control panel.

The pilot monitoring system shall include locally mounted warning lights to indicate which pilot has been extinguished. The panel shall also indicate when all pilots are operating. A set of contacts shall be provided for activating a remote trouble alarm. Under normal operation, contacts are to be closed, and will deenergize and open to alarm condition.

Ignition controls shall include upstream (in flare header prior to knockout drum) dual flow sensing equipment which shall start the automatic flare purge, pilot ignition and the flare ignition cycle. N & P Refinery will be responsible for the wiring between the flow sensors and the ignition control panel.

N & P Refinery will be responsible for the wiring and piping materials from the ignition panel to the bottom of the flare tip assembly.

All pilot and igniting piping shall be provided with break flanges located just below the burner tip flange. If the pilot is equipped with an air inspirator, the latter shall be located as low as practicable but not within 500 mm of the burner tip flange. Screwed connections shall not be used for pilot and ignition piping.

A separate ignition piping must be provided for each pilot.

Piping

Piping shall be in accordance with the requirements of ASME B3 1.3.

Flange faces shall be finished in accordance with ANSI 646.1

Flanges 24 inches and smaller shall be in accordance with ANSI 816.5.

Flanges larger than 24 inches through 60 inches shall be in accordance with API 605.

Carbon steel piping shall include at least 1/8 inch allowance for corrosion.

Cast iron piping materials shall not be used.

Minimum nominal pipe size shall be 3/4 inch. Nominal pipe size of 1-1/4, 2-1/2, 3-1/2, 5 and 7 inch shall not be used.

Use of threaded connections shall be minimized and in no case shall be larger than 1-1/2 inches. Pipe threads shall conform to ANSI 61.20.1.

Welding

All welding shall be done in accordance wiht specification ME-0-JDOO5.

Electrical

All electrical equipment shall be constructed, wired, and tested in accordance with the NFPA 70 National Electrical Code and Specificaton ME-O-JFI 15.

Enclosures for all equipment and panel boards shall be suitable for outdoor installation in areas designated Hazardous Class 1, Group D, Division 2.

Ignition panel shall be furnished with a local power disconnect.

Conduit entrance shall be from the bottom of terminal junction boxes.

All wiring shall be completely enclosed in conduit or suitable boxes. Furnished wiring shall terminate in enclosed terminal boxes.

Instruments and Controls

The vendor shall be responsible for the selection of proper range, pressure rating, materials of construction and electrical enclosure based on process conditions and electrical area classification. All instruments and controls shall be in accordance with Specification ME-0-JN400.

The flare stack design shall be in accordance with Specifications ME-O-JBOOI and ME-O-JSOIO.

Painting

All painting preparation, priming, finishing and protective coatings shall be in accordance with Specification Number ME-0-JGOO2.

Technical Documents

Services - Documents

The vendor shall supply the following documents and data:

- Flare staging calculation sheets,
- Mechanical and thermal calculation sheets,
- Checking calculation for operating conditions of equipment regarding process guarantees,
- Technical data sheets for considered equipment,
- Recommendations and drawing regarding packing and equipment transportation,
- Recommendations for extensive storage of equipment and materials,
- Instructions for erection and detailed welding specifications with recommendations for checking of site work,
- Instructions for start-up, operation and shutdown as well as those regarding prevention and suppression of failures, if any,
- Maintenance schedule indicating the estimated operating time between two interventions,
- Construction, commissioning and start-up assistance.

Drawings and Data

The Vendor shall provide general arrangement, detail and other drawings as listed in the material requisition.

Following information have to be supplied at earliest stage of engineering studies:

- Staging system PID (if applicable),
- Pressure drop through tip and gas seal,
- Purge gas consumption for gas seal,
- Minimum purge gas flow rate required to prevent tip degradation when operating at turn down condition, even under maximum wind velocity,
- Thermal radiation map for various Isopleths including 500 and 1500

BTU/hr/ft^2, (excluding solar),

- Storage blocks, cradles; stiffeners and any necessary supply for handling and transport of equipment,
- Clamps and support required for assembling and final erection of prefabricated parts,
- Spare parts for erection, pre commissioning and start-up following a list prepared by Vendor,
- Priced list of spare parts for two years operation,
- Special tools (if any),
- Vendor name plates,
- Checks in shop,
- Testing, analyses and acceptance of materials in forge or in shop welding procedure, acceptance of electrodes,
- All piping, wiring within the limits of supply.
- Flare staging control valves and rupture discs,
- Cabling and piping between burner header base and ignition/control panel and to the control room,
- Anchor bolts,
- Transport to the job site.

Limits of Supply

Termination points for the ground flare system shall be the burner tips / junction boxes at burner bottom (ground level) and on the ignition/control board.

Termination points for the flare stack shall be the flare / juction boxes at flare bottom (ground level) and on the ignition / control board.

To flare stack riser :

- Platforms, caged ladders as well as relevant guides and supports for associated piping,
- Ancillary pipeworks down to flare base,
- Cabling down to flare base with junction boxes (located four (4) feet above grade),
- Flare supports (Guy wires, etc.),
- All other items which may be required for operation of the flare stacks shall be included into the supplier's scope.

Combustion air blower located at the base of the flare stack, complete with the electric motor drive and proper control system.

The flare system shall be an automatic/manual ignition panel wholly prefabricated and wired for flame or electronic front generation ignition, complete with:

- Ignition and pilot monitoring systems,
- Fire standing panel framework including a weather protection canopy (against rain and sun),
- Lifting lugs,
- Grounding connections,
- Pressure regulators/gauges,
- Air and ignition gas feed lines with stariner, valves and bleed,
- Gas pilot,
- Gas purge line,
- Ignition chamber, if required,
- Ignition pulse generator including pulse generator with button, low tension cable, coil electrode and sight glass, if required,
- Trip amplifiers with facilities to wire back alarms to the control room,
- Terminals for cables from thermocouples,
- All controls, safety devices and items required for the proper operation of the flare.

The following shall also be provided by the Vendor :

- Flare staging design (if applicable)
- Electrodes for welding on site plus 30% excess quantity to balance job site losses.

Scope of Supply

The scope of supply includes all the required equipment, piping and instrumentation for proper flaring of gaseous effluents with smoke suppression system. The scope of supply shall include at least:

- To Ground Flare System, (if applicable):
- Nozzles for burner tips,
- Flame retention device, (if applicable),
- Flanged inlet connection,
- Lifting lugs, (if applicable),
- One set ofpipework for each pilot burner header,
- One set of ignition piping for each pilot burner header,
- One thermocouple for each pilot burner header for its monitoring, (if applicable),
- Cabling down to burner base with junction boxes (located four (4) feet

above grade),
- All other items which may be required for operation of the ground flare system shall be included into the supplier's scope.

To sonic flare tip assembly:
- Nozzle for flare tip,
- Flame retention device, (if applicable),
- Windshield,
- Flanged inlet connection,
- Lifting lugs,
- Three self inspirating pilot burners arranged at 120°,
- One set of pipework for each pilotburner,
- One set of ignition for each pilot burner,
- One thermocouple for each pilot burner for its monitoring,
- Flare smoke suppression system consisting of an air duct mixing section, air nozzle ring header and associated duct work from ground level to fire tip.

The elevated flare shall meet the noise level as specified in the Noise Control Specification ME-O- JD002.

Seismic Loading

Seismic loads shall be as specified PDVSA SPECIFICATION-JA-221 "Diseflo Antisismico de Instalaciones Industriales".

Performance

The Vendor shall guarantee that the flare system will meet the following operating characteristics.

Maximum specified system pressure drop at maximum relief rate.

Smokeless capacity for specified relief gas compositions.

Noise limit at all smokeless rate of flow.

Specified pilot gas consumption rate.

Specified purge gas consumption rate.

Requirement 20 % excess air or 35 % excess air (67 mol % unsats.)

Flashback protection of flare systems.

Documents required for engineering work, namely:

- General arrangement drawings with location and characteristics of terminal points,
- Foundation loading diagrams.

Documents shall be detailed enough to allow engineering work to start immediately with efficiency.

Documents which are not in compliance with above requirements will not be reviewed by N & P Refinery.

Drawings and data requirements for proposals and purchase orders shall be in accordance with Specification MD-0-JJOO3.

English language and the conventional American system of measurement shall be used throughout all document and tagging.

Utilities and Site Data

Utilities

The following utilities are available at the battery limits of the flare:

-Instrument Air: Normal pressure, psig : 100

-Ignition gas:

Pressure:	450 psig
Temperature:	80-140°F
Heating value LHV:	1023 Btu/SCF
MW	22,40

-Electricity: 480 V - 3 phase - 60 Hz for motor less than 200 hp.

Vendor shall advise in his quotation under a section labeled "Utilities", all utility consumption required for the system.

Meteorological Design Basis

Wind Load Design Criteria: Wind load criteria for design of structure, buildings, towers, etc, and their parts shall conform with the requirements of the latest edition of the Uniform Building Code. Basic wind speed at standard height of 33 feet shall be 100 Kph. Prevailing Wind Direction and Speed: NNE(22%) ENE (13,90/0) N (16,40/0) NE (15,3%)

Rainfall: Annual 39.6 inches. The design rainfall rates for a 5 year return period are the following: Time of Concentration= 15min. Intensity = 4.0"/hr; Time of Concentration = 1 hour Intensity = 2.0"/hr

Air Temperature: Design temperature for equipment and outdoor transformers shall be 98°F. Design Temperature for the Installation of Service Equipment - Maximum Temperature 100°F - Minimum Temperature 60°F; Design Temperature for Calculation of the Heat Loss - Normal Daily Average 79°F. The maximum mean relative humidity is 97%. Mean relative humidity - 82%

The Seismic Zone: Basic design information follows:

- a* Parameter (Seismic Map): 51gal
- y Parameter (Seismic Map): 4
- Useful Life of Installation: 20 years
- Annual Probability of Exceeding Maximum Acceleration: $\rho_1 = 0.002$
- Maximum Horizontal Ground Acceleration: $A_0 = 0.25$ g
- Soil Profile: S2
- g is Acceleration due to Gravity.

In accordance to PDVSA-JA-221 specification, Horizontal Earthquake Force (design Base Shear) is given by:

- $V_0 = A_0 * W$

- $V_0 = 0.25 * W$

Where: Ao= is the maximum horizontal ground acceleration.
W = total weight of the structure.
This force should be distributed over all system components in proportion to the mass of each component.

Others: Annual Average Normal Pressure is based on plant elevation. Site elevation is 515 ft above mean sea level.

Inspection and Testing

In addition to Vendor's shop inspection, materials and workmanship shall be subject to inspection by N & P Refinery or its authorized agent in the shop and/or in the field.

Client's Inspector will check, as a minimum, equipment components, dimensions, connection sizes and locations, and quality of workmanship against certified drawings, data sheets, and purchase order documents.

Instruments and control shall be checked for continuity and operation at the factory.

Preparation for Shipment

After final inspection, exposed machined surfaces shall be coated with a rust preventative.

Openings shall be provided with substantial plastic or metal covers. Screwed connections shall be provided with steel plugs. Loose parts and controls shall be boxed in wooden boxes, tagged, and strapped to skid.

The entire unit must be protected for six (6) months extended storage in a tropical marine enivoronment. If storage for more than six (6) months is required, N & P Refinery shall consult with the Vendor regarding proper protection.

Equipment preparation for export shall be in accordance with Specification ME-0-JDOO3, Export Packaging.

Equipment tagging and marking shall conform to the following requirements:

- The flare tip and associated equipment shall be piece marked and the electrical terminals shall be ceded to facilitate field erection.
- Instruments shall be tagged with the instrument tag numbers which will be identified by N & P Refinery on Vendor's drawings.
- In addition to the Vendor's standard identification, tags shall be furnished to show the service, N & P Refinery's tag number, and N & P Refinery's order number for each tag item listed on the Purchase Order.
- Instrument and equipment tags shall be impression stamped stainless steel with 3/16 inch (minimum) characters and permanently mounted with stainless steel drive screws or 16 gauge (minimum) wire.

- Tagging shall be in English units and language.

One complete set of the installation, operating, and maintenance instructions, shall be packed in the boxes or crates and shipped with the equipment.

Vendor's Responsibility

The Vendor shall be fully responsible for the trouble-free operation of the flare system constructed by him throughout the entire guaranteed period.

The basic code, local regulations and this specification cannot be referred to in case of a component failure.

Comments by N & P Refinery on the Vendor's documents do not relieve him of his responsibility.

Warranty and Guarantee

Supplier guarantees that equipment supplied shall be suitable for continuous outdoor operation and holding under continuous operation at design flow.

Each equipment shall be guaranteed against materials, workmanship and fabrication defects.

The Vendor shall guarantee that the flare system will meet the following operating characteristics.

- Heat intensity at grade level
- Maximum specified system pressure drop at maximum relief rate.
- Smokeless capacity for specified relief gas compositions.
- Noise limit at all smokeless rates of flow.
- Specified pilot gas consumption rate.
- Specified purge gas consumption rate.
- Specified area required.

List of Documents Required from Supplier

1. Complete Technical Data Sheet
2. General Arrangement Drawings
3. Foundation Loading Diagrams
4. Structure Calculation Notes
5. Pressure Parts Calculation Notes
6. Schedule for Studies / Orders and Fabrication
7. Engineering Drawings for Various Parts

8. Sub Suppliers Drawings
9. Erection Manual
10. Operating and Maintenance Manual
11. Drawing List
12. Refractory Materials Characteristics
13. Spare Parts List for Commissioning and Start up
14. Radiation Maps
15. Noise Maps
16. P&I Diagrams
17. Logic Diagrams
18. Erecton Schedule
19. As Built Drawings
20. Mechanical Data Book
21. Material Certificates
22. Welding Procedures
23. Welders Qualification Reports
24. Hydrostatic Test Reports
25. Shop Inspection and Test Reports
26. Performance Test Reports
27. Welds Radiographs
28. Procedure Qualification Records
29. Manufacturing Schedule Including Design, Procurement, Manufacturing Testing and Inspection, Prep. For Shipments Transportation
30. Recommendation for Protection During Transport & Storage
31. Subvendor's List
32. Main Sub-orders Copy Without Prices
33. Vendor's Procurements Status
34. Preliminary Packing List
35. Drawing for Heavy and / or Large Dimesions Equipment with Weight and Dimension
36. Shipping Documents
37. List of Commissioning Spare Parts Included in Main Purchase Order
38. Priced List of Recommended Spare Parts for 2 Years Operation
39. Proposed Quality Assurance System

DESIGN CONSIDERATIONS FOR THE FLARE HEADER

Relieving vapors from various pressure-relief and depressuring valves in the system must be collected in individual flare headers that should be appropriately located near each process area. Subheaders must be interconnected to a main flare header which feeds to a knock- out drum and disposal system. Condensates that are carried over by vapors are separated in the knock-out drum. The vapors that exit the vessel go to the flare stack where they are burned.

The number of flare headers and individual subheaders connected to them depends upon the type of vapors handled, process temperature conditions, and the available back-up pressure or limitations of the pressure receiving devices specified for the system. This section reviews some of the important design criteria and considerations for the headers and subheaders, which is an integral part of the overall flare system design.

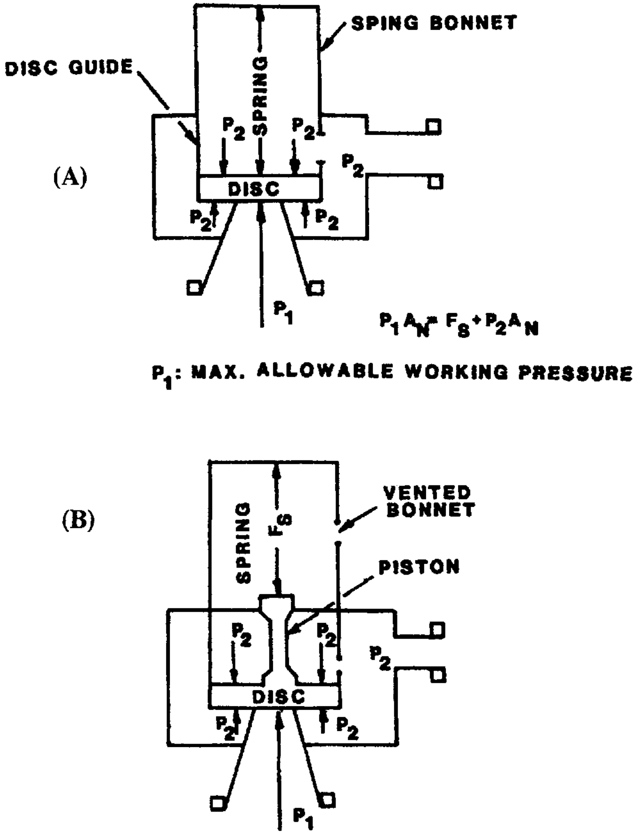
Main Flare Header and Subheader Pressure Levels

Pressure levels at the flare header depend largely on the type and nature of pressure relieving devices that have been specified to protect upstream equipment, as well as the pressure levels of all equipment connected directly to the flare stack.

In some conventional PR (pressure-relief) valves the disk of the valve is held tight against the inlet nozzle by means of a spring. A schematic of such a valve, indicating a simplified view of the forces involved in the valve's operation is given in Figure 17. As a general rule, this type of valve is limited by a maximum back pressure of 10% of the maximum allowable working pressure. The reason for this limitation is as follows. If P_1 is the maximum allowable working pressure of the vessel to which the safety valve is connected, then this is also the normal set pressure of the safety valve. Hence the spring is so loaded that the total downward force on the valve disk is exactly equal to the total upward force exerted on it by the process vapor when it reaches a pressure equal to P_1. Note that the spring force, F_s, is related to the applied force P_1 through the inlet nozzle area A_n as follows:

$$F_s = P_1 x A_n \tag{24}$$

A slight increase in pressure above P, from within the vessel will lift the valve disk up and hence relive the vapor through the discharge nozzle of the valve.



Figures 17. Effect of back pressure on (A) conventional safety valve and (B) piston type valve.

Accumulation or overpressure above the maximum allowable working pressure (P_1) within the vessel is safe up to about 10 %, provided the overpressure persists for only a short period of time. Under normal operating conditions when the vessel's operating pressure is much below P_1, the downward force exerted by the spring on the valve disk is much greater than the upward force exerted by the pressure within the vessel. As such the disk is held tight against the inlet nozzle. When the back pressure exceeds atmospheric pressure, the combined downward force exerted by the spring and the force attributed to the back pressure is the sum of F_s and P_2A_n, where P_2 is the back pressure (refer back to Figure 17).

In order to lift the valve disk against this combined downward force, the inlet vapor inside the vessel must be pressurized to a level higher than P_1.

Hence, $P_1' A_n = F_s + P_2A_n$, where P_1' is the new pressure developed.

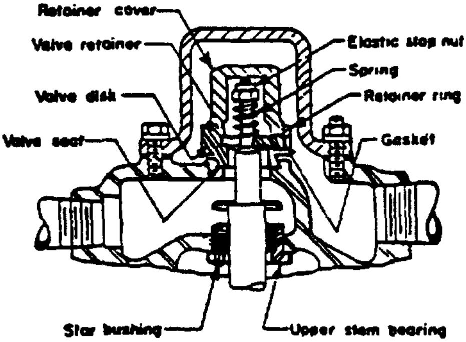
We may note that $P_1' = 1.1\ P_1$ and therefore, $P_2 = 0.1P_1$.

Thus, the maximum allowable back pressure is 10 percent of the maximum allowable working pressure of the vessel.

Other types of pressure-relief valves do not depend upon the back pressure for their performances. However, to ensure that the safety valves work at their maximum capacity, back pressure is limited to 50 percent of the relief valve set pressure. In the balanced bellows type valve, the spring does not act directly on the disk. Instead, it serves on a bellows first, which in turn acts on the disk. In case of the piston type, it works on the same principle as the bellows type, except that the bellows is replaced by a piston (see Figure 17B). The cross-sectional area of both the piston and the bellows is the same as the inlet nozzle of the valve and the effect of the back pressure on the top and the bottom of the disk creates equal balancing forces. That is, P_1A_n is always equal to F_s, as shown in Figure 17B.

Pilot-operated valves have a pilot valve combined with the main valve. The spring of the main valve provides 75 percent loading on the disk and the remaining 25 percent is offered by the gas or vapor through the pilot valve. When the vessel reaches the maximum allowable working pressure, the pilot valve relieves the gas pressure, which contributes to the disk load. Thus the safety valve becomes wide open. This is illustrated in Figure 18.

With all nonconventional valves, the maximum allowable back pressure may be taken as high as 50 percent of the valve set pressure. This pressure value



DETAIL "A"-POPPET ASSEMBLY

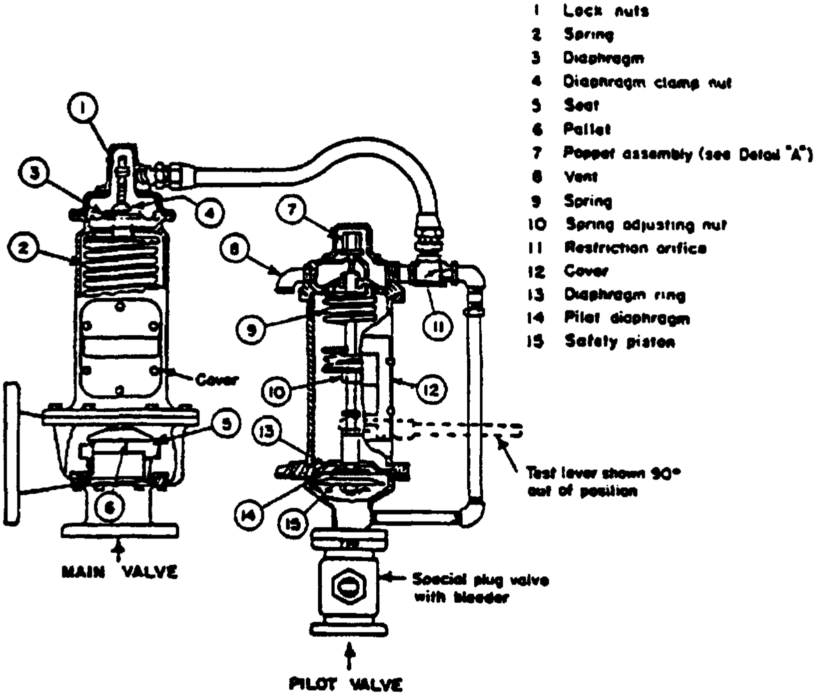


Figure 18. Pilot Operate safety valve.

approaches the critical flow pressure. If the back pressure becomes greater than the critical flow pressure corresponding to the set pressure of the safety valve, the total pressure drop available for flow of the vapors through the safety valve decreases. This state can potentially lead to over-pressurizing the vessel. The recommended back pressure is, therefore, a maximum 40 to 50 percent of the set pressure in psia. This level ensures that the back pressure is below the critical pressure.

The type of safety valves employed (either conventional or others) in a specific collection system dictates the level of back pressure in that system. In flare headers where multiple discharges exist, each safety valve must be checked so that it does not exceed its allowable back pressure.

The design pressure levels of process equipment connected to a flare system are also important to determine flare header pressure levels. In some cases, pressure varies widely from one process vessel to another. Hence, it is not always economical to combine systems with a common header. For example, if the set pressures are 20 psig and 200 psig for two different systems, for the low pressure system the maximum back pressure attainable is 17 psia by using a balanced type safety valve. The high-pressure system is capable of withstanding a back pressure as high as 34 psia even with a conventional type safety valve. But if these two valves are connected to the same header, the maximum pressure level of the header will be 17 psia in order to protect the equipment from overpressure, and this requires a much larger pipe diameter. Consequently, it is often more economical to have flare headers of two pressure levels, one connecting the low-pressure system and the other connecting the high-pressure system.

The basis for collection philosophy of a flare system is based upon an economic evaluation. The methodology is outlined in the next subsection.

Determining the Number of Flare Headers

The number of flare headers required depends upon an economic evaluation of system combinations that will result in the minimum piping cost. The following steps outline the procedure for comparative estimations:

1. Study of the Plot Plan Layout: From the plot plan layout the number of safety valves in different process areas, the set pressures of safety valves, individual relieving loads of safety valves, relieving temperature of vapors, the nature of vapors (i.e., whether corrosive, condensable, or dry) are noted.

2. A single subheader in each process area is drawn up, connecting area pressure-relief valves or depressuring valves.

3. The subheaders are then connected to give a single main flare header based upon the shortest routing.

4. The equivalent length of the main flare header is then calculated from the flare stack to the last safety valve, taking into consideration the straight length of the pipe and approximate equivalent lengths for bends, etc. If the actual location of the flare stack is not known by that time, it maybe assumed to be 500 ft from the last piece of equipment. Later on, even if it varies from 500 ft, it will not affect the pressure drop calculation at all compared with the entire length of the pipe.

5. A trial and error estimate is made for determining the diameter of the flare header based upon the maximum relieving flare load and considering the back pressure limitation of 10 percent for conventional valves and 40 percent for balanced type valves. Note, however, a single main header in most cases turns out to be too large to be economically feasible. Line sizing procedures are discussed in detail in the next subsection.

6. A second trial is required for two main flare headers, one collecting the low-pressure flares (usually 5 to 10 psig) and the other collecting relatively high-pressure flares (usually 15 to 20 psig). The two headers are connected to their individual knock-out drums. The vapor lines from the knock-out drums are combined into a single header connected to the flare stack.

The maximum simultaneous load in each header must be calculated separately and the pressure drop must also be computed for the entire length of the pipe including the combined length from the knock-out drum to the stack.

The subheaders in each process area similarly will have two levels of flare headers. The line sizing of each level of subheader in an individual area will depend upon the maximum simultaneous flow in that particular area. Thus the line sizing criterion of a subheader may be the largest single flow due to a blocked outlet condition. This flow may not necessarily be the controlling load for the flare stack.

7. The next consideration is the cost of construction materials. This determines the final number of flare headers. Vapors that normally require expensive materials may be listed as follows:

   a. Corrosive vapors, e.g., SO_2, H_2S.

b. Very high temperature vapors, e.g., high temperature gases used for regeneration of catalysts in reactors.

c. Very low temperature vapors, e.g., vapors generated due to flashing across a control valve or a safety valve in a cryogenic system.

Of the three, corrosive vapors are usually piped up in a separate header right up to the flare stack since such lines are normally very small and if combined with other streams may run the risk of corroding the much larger and more expensive pipelines.

For a high-temperature system, a separate subheader may be run up to the point where the temperature drops down to the allowable limit of a less expensive material. It may then be connected to the main flare header (either low pressure or high pressure).To properly evaluate this a heat loss calculation is needed. As a rule of thumb a heat loss of 10 BTU/hr/ft^2 may be assumed for a quick estimate for bare pipe. Consideration should also be given to the need for expansion joints. Main flare headers may be as large as 36 to 42 inches in diameter for a large-capacity plant. Expansion joints of such magnitudes may be so expensive as to call for a separate small header for the hot flare system.

The flare subheaders carrying very low temperature vapors (temperatures ranging from 50°F and below) may similarly be combined into a single low temperature flare header and piped all the way up to the flare stack. Again, since the atmosphere is warmer than the pipes, a heat gain calculation may indicate that the cold subheaders after running a certain distance by themselves may be safely combined either with the low-pressure main flare header or the high-pressure main flare header, depending upon their operating pressure.

8. *Wet Flare and Dry Flare:* Sometimes relatively hot vapors carrying condensates may be separated from the dry cold vapors. They do not run as separate headers but either low-pressure or high-pressure flare headers may be associated with any one of them. Thus a wet flare header may be, in fact, the low-pressure header, and the dry flare header maybe the high-pressure flare, or vica-versa.

9. After the total number of flare headers has been established, it may be necessary to recheck the vapor load in individual headers since introduction of a separate header may allow subtraction of the flow quantity from the low-pressure header to which it was added initially.

Two examples are an ethylene plant and a coal gasification plant.

A typical ethylene plant usually has the following flare headers:

a. Low-pressure wet flare header containing hot gas and water.

b. High-pressure dry flare header containing cold gas and devoid of water.

c. Liquid drain header containing low-temperature vapors after flashing across valves.

A typical coal gasification plant has the following flare headers:

a. High-pressure dry flare header.

b. High-pressure wet flare header.

c. An H_2S header containing vapor of more than 5 percent (volume of H_2S).

Line Sizing for Flare Headers

When the maximum vapor-relieving requirement of the flare system has been established and the maximum allowable back pressure (as just described) has been defined, line sizing reduces to standard flow calculations.

The flare piping system can be divided into the following sections

1. Individual discharge lines from the pressure-relief valves.

2. The subheaders in each area connecting the discharge lines.

3. The main flare header connecting the subheaders leading to the knock-out drum.

4. The final header connecting the vapor line (or lines) from the knock-out drum leading to the flare stack.

Since vapors in the flare headers are relieved from a high pressure system to almost atmospheric pressure, there is an appreciable kinetic energy change throughout the pipeline. The flow condition is that of compressible flow. The nature of compressible flow in the case of flare headers may be assumed to be isothermal since flare lines are normally long and not fully insulated. Short and properly insulated vapor lines are close to an adiabatic flow. In general, all vapor flows that normally occur in process plants are somewhere in between adiabatic and isothermal flow. It has been observed that for the same flow rate and pressure drop, line sizing calculations based on compressible isothermal flow conditions always give an equal or a larger diameter pipe. Hence, flare headers should be sized based upon isothermal compressible flow for a more conservative design. The following criteria are used in sizing flare headers:

1. The back pressure developed at the downstream section of any pressure-relief valve connected to the same headers should not exceed the allowable limit, i.e., 10 percent of the set pressure in psig for the conventional type and 40 to 50 percent of the set pressure in psia for the balanced type valve.

2. Since the pressure drop is quite high, there is a possibility of approaching sonic velocity in the line. This will result in a potential noise problem. Hence, it is a good practice to limit the velocity to 60 percent of the sonic velocity or a 0.6 Mach number.

A quick method for sizing compressible isothermal flow is offered by the following method developed by Lapple.

This method employs a theoretical critical mass flow based on an ideal nozzle and isothermal flow condition. For a pure gas, the mass flow can be determined from one equation:

$$G_{ci} = 12.6\, P_0 \sqrt{\frac{M}{(2Z-1)T_0}} \tag{25}$$

where:

G_{ci} = maximum mass flow or critical mass flow, in lbs per sec per sq. ft
P_0 = upstream pressure, in lbs per sq in, absolute
M = molecular weight of the gas
T_0 = upstream temperature in °R
Z = compressibility factor

The actual mass flow rate through a pipe, G, in lbs per sec per sq ft is a function of critical mass flow G_{ci}, line resistance, N, and the ratio of downstream to upstream pressure. These relationships are plotted in Figure 19. In the area below the dashed line in Figure 19 , the ratio of G to G_{ci} remains constant, which indicates that sonic flow has been established. Thus, in sizing flare headers the plotted point must be above the dashed line. The line resistance, N, is given by the equation:

$$N = \frac{4fL}{D} + \sum K_i \tag{26}$$

where:

L = equivalent length of line, in ft
D = line diameter, in ft
f = Fanning friction factor (obtained from Figure 21)
N = line resistance factor, dimensionless

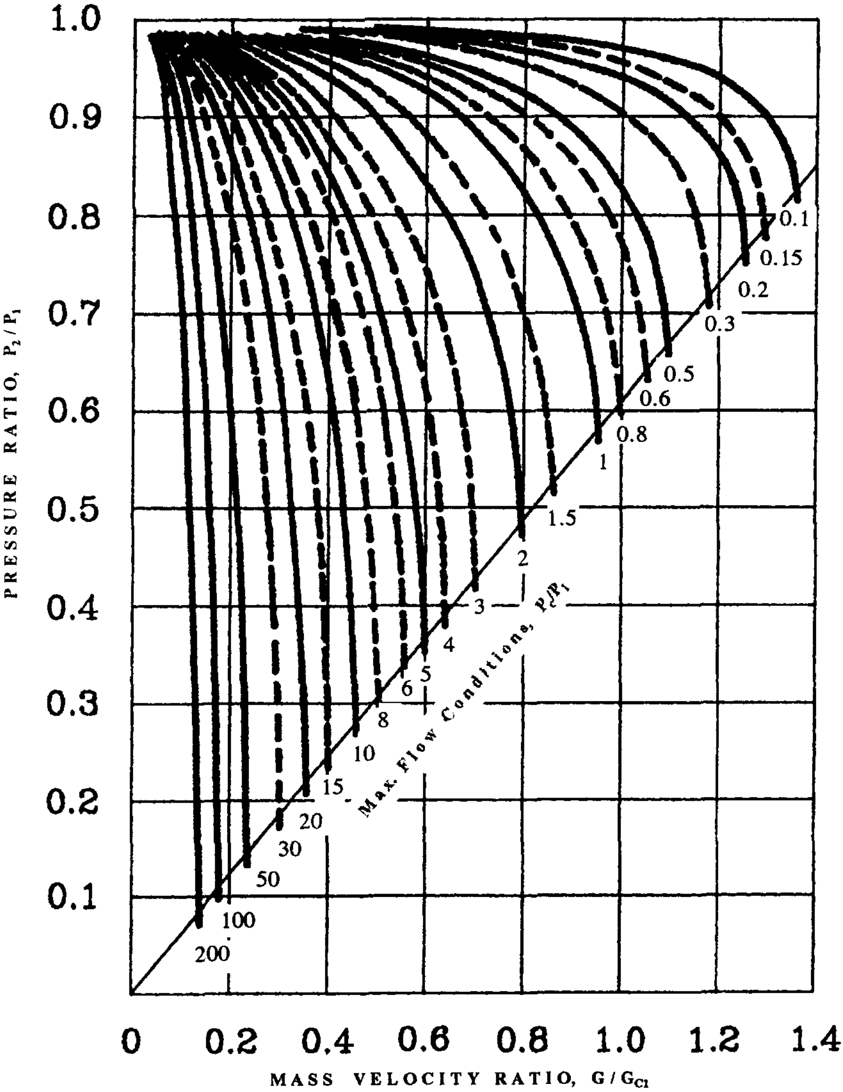


Figure 19. Pressure drop chart - after Lapple.

K_i = resistance coefficients for pipe fittings (see Table 6).

Lapple's method is useful when the upstream pressure of a header is known and the downstream pressure has to be calculated. However, it is often required to develop the pressure profile of the flare headers as a function of the distance from the stack. For this reason, it is more convenient to calculate the pressure drop backward, starting from the flare stack exit where the pressure is atmospheric. Figure 20 provides another plot which enables the pressure loss calculation when the downstream pressure is known.

Although Figures 19 and 20 can be used for line sizing, it should be noted that Figure 19 requires more extensive trial and error calculations.

The following steps summarize the procedure for sizing flare headers:

1. The pressure at the base of the flare stack is approximated as 2 psig. Depending on the type of seal used in the flare stack, the pressure at the base may vary slightly although 2 psig is generally a good approximation:

Pressure at the base = Atm pressure at the flare exit + 0.5 psi flare tip.

This can be rationalized as:

ΔP + 0.5 psi mol seal ΔP + 1 psi ΔP due to flow through the stack height = 2 psig

2. Compute the pressure in the knock out drum:

= 2 psig + ΔP required for the flow of the full load vapors from the knock-out drum to the stack + 0.5 psi ΔP assumed inside the knock - out drum.

In the absence of actual line distances, a conservative estimate of the distance between the knock-out drum and the flare may be taken as 500 ft.

3. As a first trial, an inside pipe diameter is assumed based on 60 percent of the sonic velocity corresponding to the pressure and temperature at the base of the stack, i.e., at 2 psig and temperature $= T_0$ (upstream temperature since isothermality is assumed).

The sonic velocity can be computed from:

$$V_s = 223 \sqrt{\frac{KT}{M}} \qquad (27)$$

where: V_s = sonic velocity in ft/sec

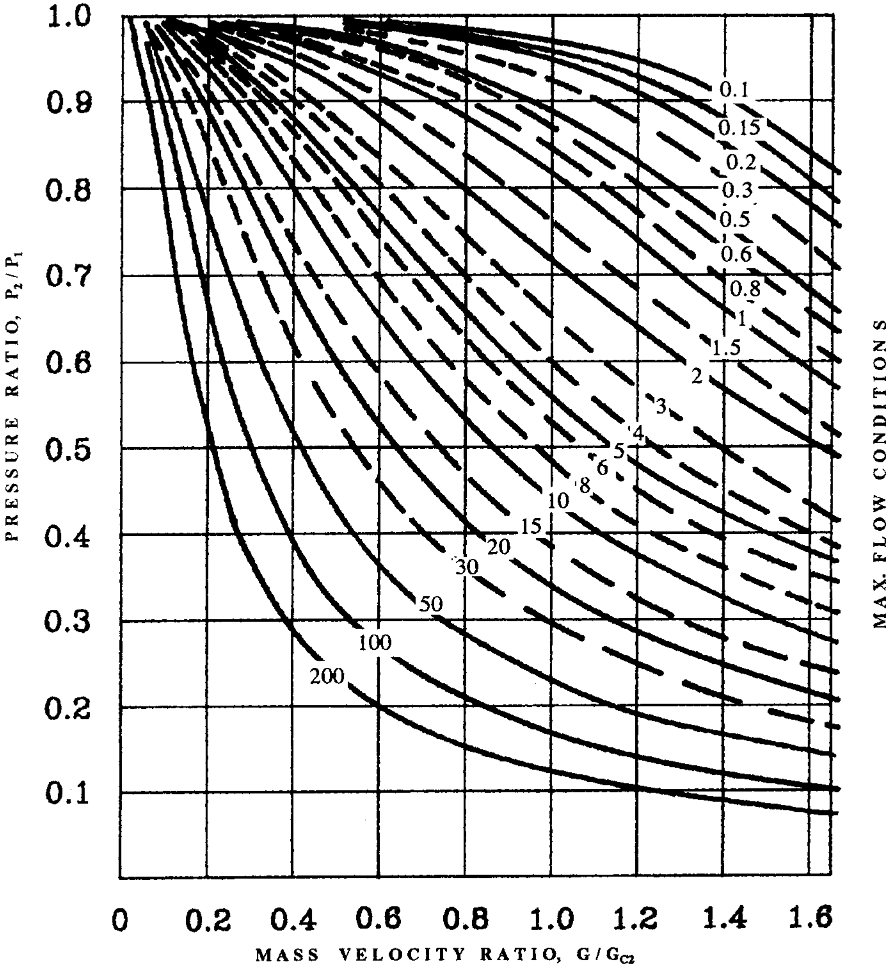


Figure 20. Pressure drop chart -after Loeb.

$K = C_p/C_v$ of the gas, normally between 1 to 1.8
T = temperature in °R
M = molecular weightof the gas.

The flare load, W, lbs/sec, is known. The density of the vapor at 2 psig and T_0 (°R) is determined from the ideal gas law:

$$P = \frac{MP}{RT}$$

where: M = molecular weight.
P = pressure in psia
T = temperature, °R
R = gas constant

Hence,

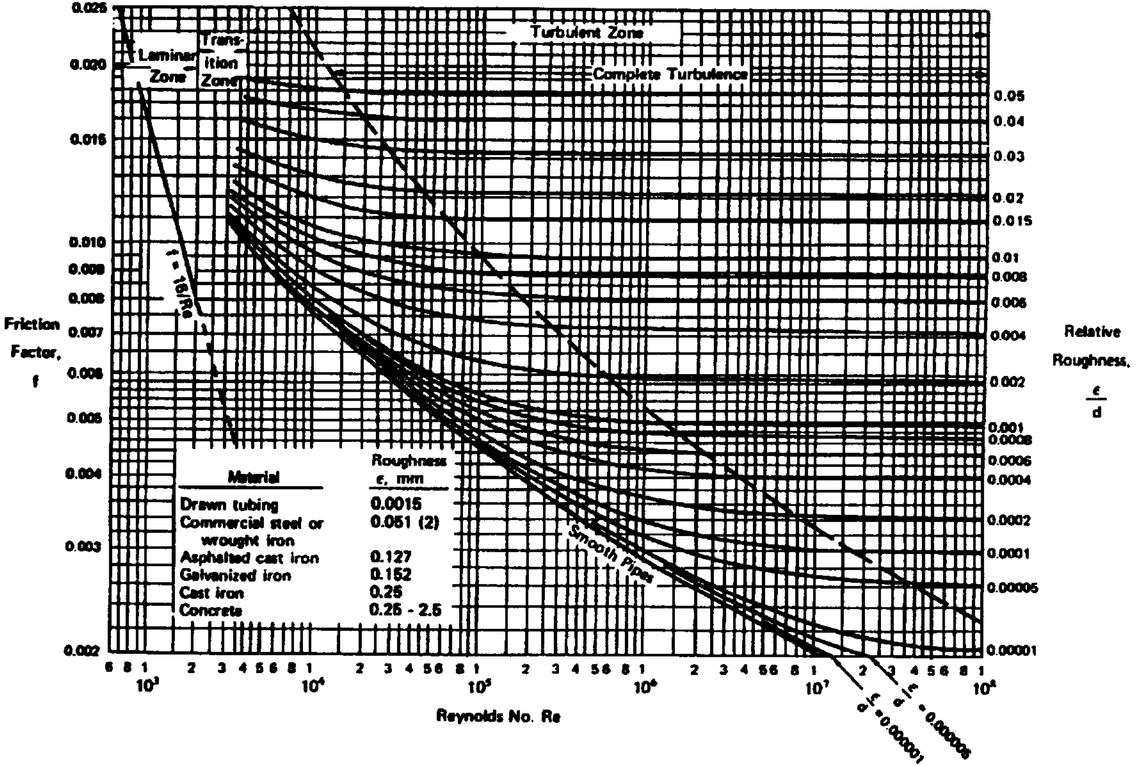
$$\frac{W}{\pi d^2/4} = 0.6x223\sqrt{\frac{KT}{M}} \tag{28}$$

where: d = pipe ID in ft

The inside diameter d can be calculated from Equation 28 since everything else is known.

Once the diameter is known, the Reynolds number, Re can be computed and the friction factor f obtained from Figure 21. Assuming a straight length of pipe for L = 500 ft, N (line resistance factor) can be calculated. Next G_{ci} is calculated based on the downstream pressure and G/G_{c2} evaluated. From Figure 20 the ratio P_2/P_0 can be obtained. Since P_2 is known, P_0 can then be calculated. The pressure at the inlet of the knock-out drum is given by P_0 + 0.5 psi. Table 6 provides typical values of resistance coefficients for various pipe fittings.

4. From the knock-out drum, the individual flare headers can now be sized in the same manner. Based on a Mach number of 0.6 and the density corresponding to (P_0 + 0.5) psia, a trial diameter can be established. The pressure at every intersection between the subheader and the main header must be calculated,with the downstream pressure being (P_0 +0.5) psia. Knowing the pressure at the intersection of the subheader and the main header, the pressure at the intersection of the subheader and the discharge pipe of the safety valve is computed. Usually the discharge pipe of the safety valve is small and hence, a separate pressure drop calculation is not necessary. However, the velocity at the discharge pipe should be checked to ensure that it is below sonic velocity. If the discharge pipe runs a considerable distance before it ties in with the subheader, a separate pressure drop calculation will be required. Figure 22 shows a typical



Material	Roughness ε, mm
Drawn tubing	0.0015
Commercial steel or wrought iron	0.051 (2)
Asphalted cast iron	0.127
Galvanized iron	0.152
Cast iron	0.25
Concrete	0.25 - 2.5

Figure 21. Generalized Moody plot for obtaining friction factor.

Table 6. Resistance coefficient K for various pipe fittings.

Fitting	*K*	Fitting	*K*
Globe valve, open	9.7	90° double-miter elbow	0.59
Typical depressuring valve, open	8.5	Screwed tee through run	0.50
Angle valve, open	4.6	Fabricated tee through run	0.50
Swing check valve, open	2.3	Lateral through run	0.50
180° close screwed return	1.95	90° triple-miter elbow	0.46
Screwed or fabricated tee through branch	1.72	45° single-miter elbow	0.46
90° single-miter elbow	1.72	180° welding return	0.43
Welding tee through branch	1.37	45° screwed elbow	0.43
90° standard screwed elbow	0.93	Welding tee through run	0.38
60° single-miter elbow	0.93	90° welding elbow	0.32
45° lateral through branch	0.76	45° welding elbow	0.21
90° long-sweep elbow	0.59	Gate valve, open	0.21

d/d':	*0*	*0.2*	*0.4*	*0.6*	*0.8*
Contractions (USASI)	—	—	0.21	0.135	0.039
Contractions (sudden)	0.5	0.46	0.38	0.29	0.12
Enlargements (USASI)	—	—	0.9	0.5	0.11
Enlargements (sudden)	0.0	0.95	0.74	0.41	0.11

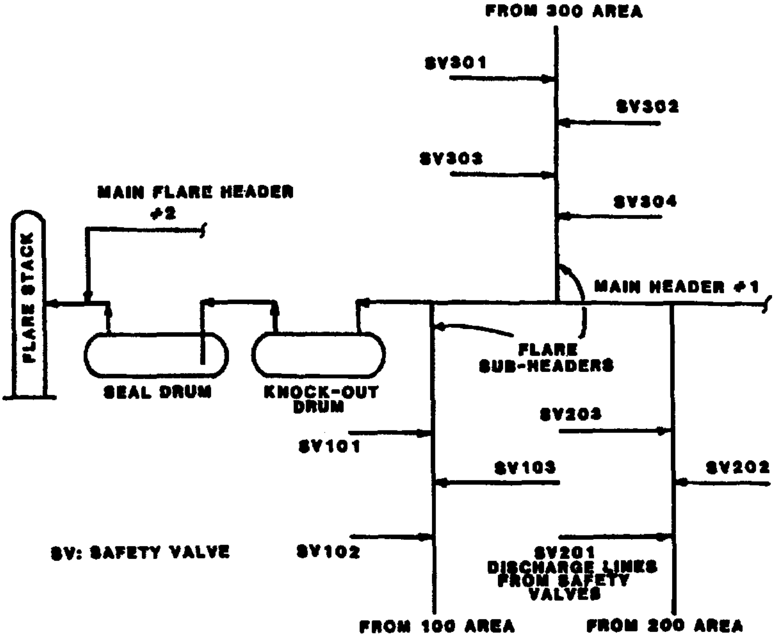


Figure 22. Typical layout of a flare collection system.

layout of the flare piping.

The sum of all the pressure losses starting from the flare stack up to the safety valve yields the total back pressure in psig. This back pressure must be lower than the maximum back pressure allowed in the system and corresponding to the lowest set pressure of the safety valve.

The following examples illustrate the calculations needed to properly size lines for flare headers.

Example: The maximum flare load of a system is 1,000,000 lbs/hr of vapor. The pressure at the base of the flare stack is 2 psig, the average molecular weight of the vapor is 50, at a temperature of 200°F at the combined header to the flare stack. The distance from the drum to the stack is 500 ft. The line consists of two 90° welding elbows and an orifice for a flow controller. The total pressure drop at the knock-out drum is 0.5 psi. Determine the pressure at the inlet of the knock-out drum.

Additional Data:
$K = C_p/C_v = 1.2$
Z (compressibility factor) ≈ 1

Trial No. 1 (From Equation 28):

$$\frac{W}{\pi d^2/4} = 0.6 \mathrm{x} 223 \sqrt{\frac{KT}{M}}$$

$$\frac{1{,}000{,}000 \mathrm{x} 4}{3600 \mathrm{x} 0.125 \mathrm{x} 3.14 d^2} = 0.6 \mathrm{x} 223 \sqrt{\frac{1.2 \mathrm{x} 660}{50}}$$

$$\rho = \frac{MP}{10.73T} = 0.125 \text{ lbs/cu. ft}$$

d = 2.3 ft = 27.7 in. ≈ 28 in, corresponding to standard pipe 30 in., 20 schedule.

Compute Line Resistance Factor (N):

Since the Fanning friction factor "f" does not vary much, f =0.004 is a good approximation. However, it can be calculated from Re by the standard method from Table 6.

$$\sum K = 0.32x2 + 0.2$$

This calculation includes the contribution from the elbow and the sudden enlargement.

$$N = \frac{4fL}{D} + \sum K_i = \frac{4x0.004x500x12}{29} + 0.84 = 4.04$$

Determine G/G_{ci}:

$$G = \frac{1,000,000x4x12x12}{3600x3.14x29x29} = 56.6 \text{ lbs/sec/ft}^2$$

$$G_{ci} = 12.6\ P_0 \sqrt{\frac{M}{(2Z-1)T_0}} \quad \text{(From Equation 25)}$$

The upstream pressure P_0 will be replaced by the downstream pressure = 2 + 14.7 = 16.7

$$Z = 1,\ T_0 = 200+460 = 660°R \text{ (isothermal)}$$

Hence, $G_{ci} = G_{c2} = 12.6x16.7\sqrt{\frac{50}{660}} = 57.97 \text{ lbs/sec/ft}^2$

(G_{c2} represents mass flow rate based on downstream conditions)

$G/G_i = 56.6/57.92 = 0.977$

$P_2/P_1 = 0.59$ (From Figure 20)

$P_1 = 16.7/0.59 = 28.3$ psi

$P = P_1 - P_2 = 28.3 - 16.7 = 11.6$ psi which is too high (as a rule of thumb, ΔP should not exceed 3 psi). This means that a larger line size should be chosen.

Trial No. 2

Assume an inside diameter of 35.25 in. (i.e., nominal 36 in.pipe, standard thickness).

$N = 3.5$, $G/G_{c2} = 0.678$
$P_2/P_1 = 0.79$, $P_1 = 21$ psi, $\Delta P = 4.3$ psi

Trial No. 3

ID = 39.25 in., (40 in. std.)
$N = 3.24$,$G/G_{c2} = 0.55$
$P_2/P_1 = 0.85$, $P_1 = 19.6$ psi, $\Delta P = 3.3$ psi.

Trial No. 4

ID= 41.25 in., $N = 3.125$, $G/G_{ci} = 0.483$ (Corresponding to 42 in., standard thickness)

$P_2/P_1 = 0.87$, $P_1 = 19.2$, $\Delta P = 2.5$ psi

Hence, total pressure drop

$= 2.5 + 0.5 + 0.25$

= Line ΔP + K.0. drum + orifice ΔP

$\Delta P = 3.25$ psi

The pressure at the inlet of the knock-out drum is $16.7 + 3.25 = 19.95$ psi or 20 psi

Example: Calculate the dry flare header size, which is connected to the above knock-out drum. Previously noted conditions remain the same. Additional data are:

W = 720,000 lbs/hr, Temperature = 100°F, and the maximum allowable back pressure at the safety valve (as shown in Figure 23) is 34 psia. The discharge rate from the safety valve is 30,400 lbs/hr.

Trial No. 1

Assuming a Mach number of 0.6 and d = 21.25 in. (22 in.standard).

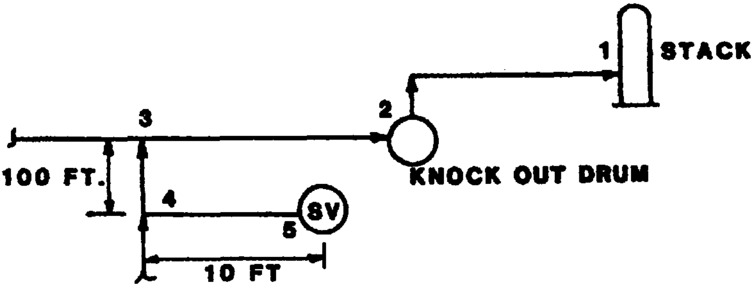


Figure 23. System definition for Example.

$$\rho = \frac{\mathbf{50x20}}{\mathbf{10.73x560}} = \mathbf{0.169\ lbs/cu.ft}$$

$G/G_{ci} = 0.988$, $N = 4fL/D = 17.45$, P_2/P_1 from Figure 20 $= 0.36$, $P_1 = 56.5$ psi, which is too high.

Trial No. 2

Use ID = 29 in.
$G/G_{ci} = 0.53$, $N = 12.8$, $P_2/P_1 = 0.66$, $P_1 = 30.8$ psi,
$\Delta P = 30.8 - 20$ ft= 10.8 psi

Trial No. 3:

ID= 27.25 in. (28 in. Standard)
$G/G_{ci} = 0.568$, $N = 13.7$, $P_2/P_1 = 0.61$, $P_1 = 33.36$ psi

Hence, either a 28 or 30 in, diameter is acceptable for the main header size depending on the ΔP at the subheader.

We must now consider the pressure drop between points 4 and 5 in Figure 23. Since the distance between points 4 and 5 is only10 ft, ΔP calculations are not warranted. Instead, we can assume the same diameter for the subheader and the discharge line of the safety valve. It is quite possible that because of some other safety valves discharging to the same subheader, the load on the subheader is much higher than the given rate from the safety valve, i.e., 30,400 lbs/hr. As a result, the subheader size should actually be larger than that specified based on 30,400 lbs/hr. Thus, if the size of the subheader is not known at the time, a conservative estimate of ΔP can be made assuming a smaller diameter for the subheader corresponding to 30,400 lbs/hr.

Trial No. 1

$\rho = \frac{\mathbf{50x33.36}}{\mathbf{10.73x560}} = \mathbf{0.277\ lbs/cu.ft}$ @33.36 psia and 100°F corresponding to

28 in. ID for main header

For a Mach number of 0.6, d = 3.3 in.
But the outlet nozzle size of the safety valve is 6 in. Schedule 40. Hence, 6 in., will be minimum line size. (This is normal practice.)

ID = 6.065 in.
G/G_{ci} = 0.34, N = 3.52, P_2/P_1 = 0.94, and P_1 = 35.4 psi

Trial No. 2:

ID = 7.981 in., corresponding to 10 in., Schedule 40 pipe
G/G_{ci} = 0.192, N = 2.65, P_2/P_1 = 0.98, P_1 = 34.1 psia

In summary, we have obtained the following:

Combination No. 1 - 42 in. of 500 ft, 28 in. of 2,000 ft, 8 in, of 110 ft, pressure at point 5 = 34.1 psia

Combination No. 2 - 40 in. of 500 ft, 30 in. of 2,000 ft, 8 in, of 110 ft, pressure at point 5 = 32.15 psia

Combination No. 3 - 40 in, of 500 ft, 32 in, of 2,000 ft, 6 in, of 110 ft, pressure at point 5 = 32.82 psia

Combination No. 1 appears to be the most economical since the 28 in. line has the maximum run. Next is No. 2, compared to No. 3 for the same reason.

11

STACK DESIGN OVERVIEW

INTRODUCTION

This chapter explores the design of stacks from the point of view of the downwind observer whose task is to determine the connection between stack design, process emissions, meteorology, and, most important, environmental effects. Stacks must be designed to specifications based on meteorological conditions and environmental air quality standards, which may be quite unrelated to process requirements

The principal factors which must be accounted for when designing a stack for air pollution control purposes are the dispersion and transport of the pollutants and the performance criteria against which the stack will be compared. These factors include: (1) air quality standards, (2) meteorological conditions, and (3) topographical peculiarities.

The problem of designing a stack to exploit its air pollution control potential largely reduces to a problem of determining a stack height which will assure nonpolluting performance. This means designing a stack to meet some performance standard (usually legally binding) given the meteorological conditions, topographic influences, and process exit conditions. Usually, the exit gas conditions are unalterable and the topographic influences are unknown or speculative. This leaves the meteorology and the air quality standards as the governing design criteria.

Historically a 2-1/2 design rule, which states that the height of a stack should be 2-1/2 times higher than the nearest surrounding structure, represented the most reliable way to design stacks for avoidance of ground-level pollution problems. As regulators became more concerned with the effects of increasing pollutant concentrations in the atmosphere, ground-level ambient air standards were adopted which prescribed maximum tolerable ambient air concentrations for a variety of substances. The Clean Air Acts authorized the U.S. Environmental Protection Agency (EPA) to promulgate Primary and Secondary National Ambient Air Quality Standards to protect the health and welfare in the United States.

Source operators must, therefore, be certain that they are familiar with the standards applicable to plant operations, for the performance of the source may ultimately be compared against a variety of standards. The rule of thumb in determining which state or federal standard applies when they appear to conflict is that the more stringent standard prevails.

In addition to environmental regulations governing the concentration of ground-level pollutants, there also exist Federal Aviation Administration regulations on the maximum permissible stack height at a given distance from an airport or along air corridors. These regulations may limit the air pollution control potential of a stack by restricting upper limits on stack heights.

The first step in designing a stack for air pollution control purposes is to determine exactly what regulatory constraints and requirements exist at the particular site. These constraints and requirements may be so severe that alternative means of air pollution control may have to be sought. In any case, the regulations specify a performance standard to which the stack must be designed, and against which the design can be evaluated.

METEOROLOGY

Meteorological conditions, as much as any other consideration, determine how a stack should be designed for air pollution control purposes. Operating transport mechanisms are determined by the micro meteorological conditions, and any attempt to predict ground-level pollutant concentrations is dependent on a reasonable estimate of the convective and dispersive potential of the local air. The following are meteorological conditions which need to be determined:

1. Mean wind speed and direction: the air flow is assumed to be horizontal, but the flow may be tilted (to yield a vertical component) due to local topographic effects. The mean wind speed determines the convection of the stack emissions.

2. Intensity of turbulence: these factors, represented by the standard deviations of the horizontal wind direction, σ_y , the standard deviation of the vertical wind component, σ_z, and the gustiness as measured by the standard deviation of the wind speed, all have significant bearing on the dispersion of emissions from a stack.

3. Vertical temperature gradient: the lapse rate (rate of decrease in temperature with increases in height) must be taken into account because it affects the final height to which a buoyant plume rises.

These meteorological parameters, with the possible exception of the mean wind speed and direction, are not generally available for inclusion in calculations. Even wind speed measurements, which are usually taken at 20 ft above grade, must be corrected to the release point elevation. The correction applied to the wind speed depends on the turbulence of the air. The wind speed is the key determinant of the convection of pollutant in a plume.

The vertical temperature gradient (the lapse rate) is usually not monitored by routine meteorological observation, and it, too, must be approximated from estimates of solar insolation, solar angle, and differential heating due to uneven cloud cover. For purposes of diffusion analyses, the lapse rate is usually approximated by a constant.

The parameters about which the least is known are the diffusion parameters which govern diffusion transport of pollutants within a plume. These parameters are not monitored by meteorological stations and must always be approximated through indirect methods. Figure 1 illustrates the role each of these parameters has in the transport of airborne pollutants.

Experimental works have shown that the vertical distribution of diffusing particles from an elevated point source is a function of the standard deviation of the vertical wind direction at the release point. It is known that the standard deviations of the vertical and horizontal wind directions can be related to the standard deviations of particle concentrations in the vertical and horizontal directions within the plume itself. This is equivalent to saying that fluctuations in stack top conditions control the distribution of pollutant in the plume. Also it can be noted that the plume pollutant distributions follow a diffusion relation that can be approximated by a Gaussian distribution.

The Pasquill and Gifford approach described later, removes the need to concentrate on determining σ_z and σ_y (refer to Figure 1) directly from weather data. In order to do this, Pasquill introduced the concept of the atmospheric stability class.

Pasquill defined six stability classes ranging from highly stable, low-turbulence Class F, to unstable, highly turbulent Class A, and he identified the surface wind speed, intensity of solar radiation, and nighttime sky cover as being the prime factors controlling atmospheric stability. Pasquill then correlated observations of the behavior of plumes in terms of their dispersion with the

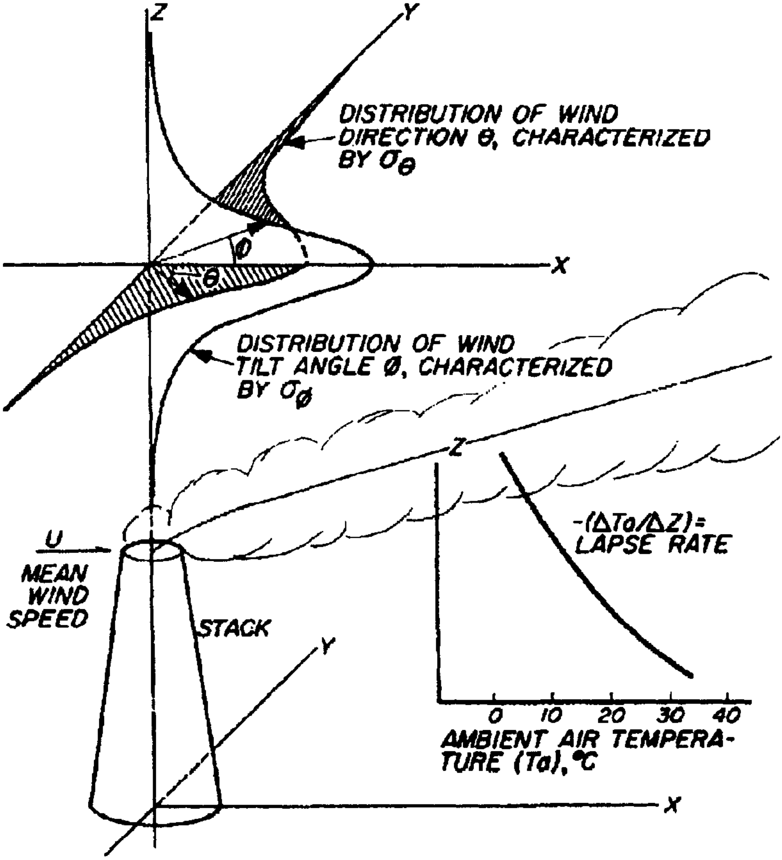


Figure 1. Role of meteorological parameters in transport of airborne pollution.

identified prime meteorological factors.

For each stability class a value for the standard deviations was determined as a function of downwind distance These standard deviations may be used to estimate the concentration of pollutant at any point in the plume.

LOCAL TOPOGRAPHICAL FACTORS

The following discussion shows the effect that topographic factors can have on the transport of pollutants and, consequently, on the design and especially the siting of the stack.

Effects of Nearby Ridges. An example of the effect that nearby mountain ridges may have on pollutant transport is given in Figure 2. High concentrations of pollutants emitted from the source stack would probably be expected to occur on the windward slope of the ridge shown, but, in fact, observations at smelting plants and thermal generating stations located in valleys indicate that the leeward side of the ridge often experiences the more severe pollution due to wind induced eddy formation.

Probably there is little that can be done in the design of a stack to avoid this problem, although raising the elevation could be of value. In such cases, air pollution control can probably be implemented only by process reduction during unfavorable wind conditions or by permanent stack gas cleaning devices. In the lee of a cliff there may be eddies of the type shown in Figure 3.

This example points out one of the central problems in stack design for pollution control: local, short-term effects may be the most important stack design consideration, but will usually be the aspect of the problem about which the least information is available.

Diurnal Air Circulation Effects in Valleys. Figure 4 shows some of the complications associated with siting a stack in a valley. When the winds are light, the air in a valley becomes stagnant and stratified, especially at night and particularly in winter. Pollutants emitted at night may be drawn down to the valley floor by cold air. During the day most of the air in the valley remains stably stratified while warm air flows up the hillsides, receiving heat from them as it ascends. As shown in Figure 4, a circulation pattern is set up where the plume is forced to fan out.

Effects of Large Local Water Bodies. The siting of a stack along the shoreline of a large lake or along an ocean front may be accompanied by unique air

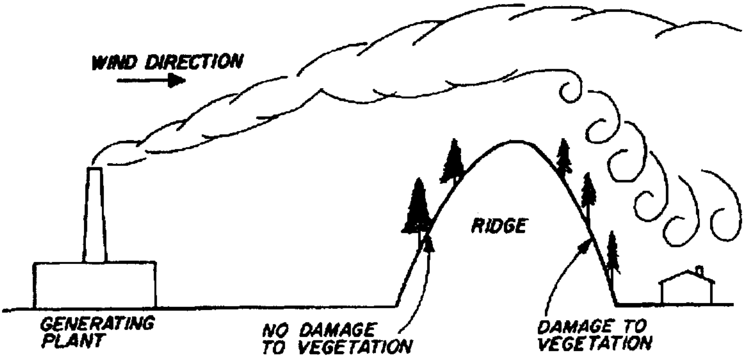


Figure 2. Effects of nearby mountain ridges on pollutant transport.

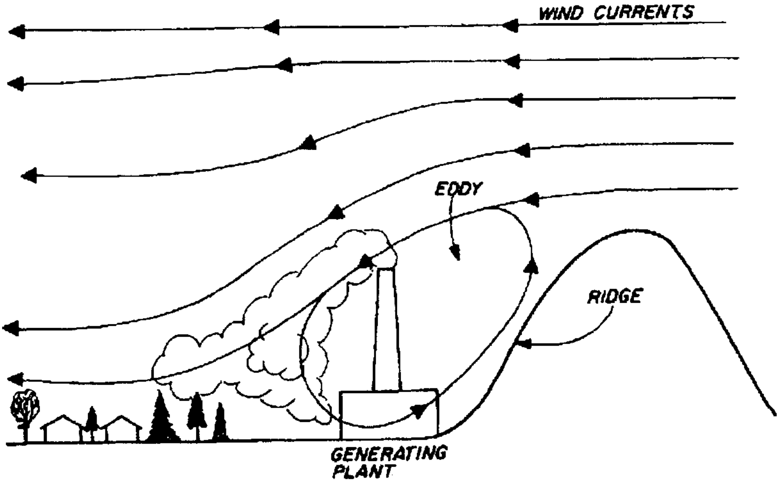


Figure 3. Eddy information in the lee of a cliff.

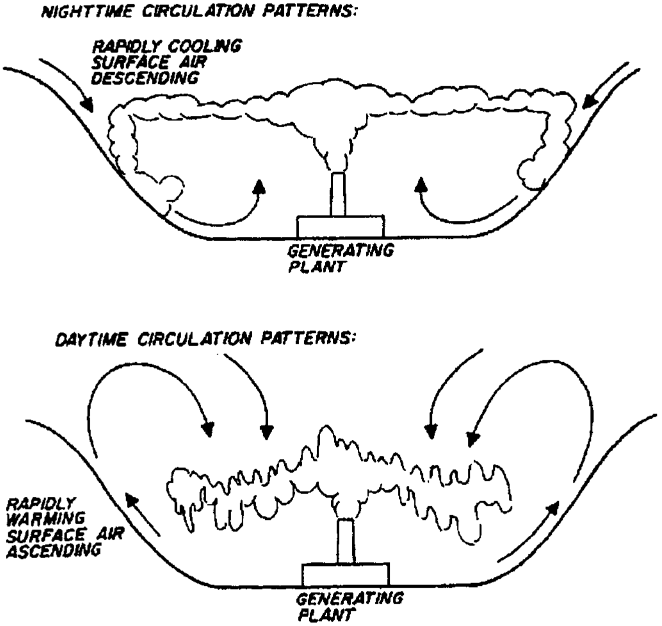


Figure 4. Diurnal air circulation effects associated with valleys.

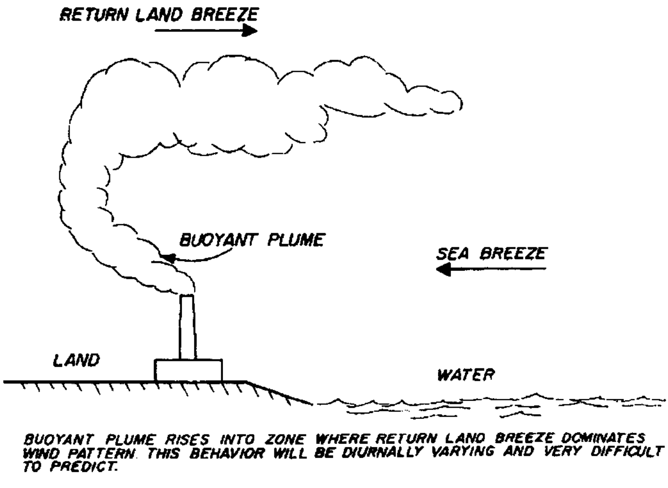


Figure 5. Effects of large water bodies on pollutant transport.

pollution transport phenomena. Figure 5 shows how thermal convection wind currents arising from differences in the temperatures of land and water masses can influence transport properties. Such convection phenomena may be highly dependent on meteorological and seasonal conditions and may defy the available methods to estimate them quantitatively. The pattern shown in Figure 5 is appropriate for an occasion when the land mass is warmer than the water, a frequently encountered summer air circulation phenomenon which reverses in the evening as the land cools off faster than the lake or ocean.

It is doubtful that stack design would have any significant influence on controlling this type of atmospheric transport problem, and it is more unlikely that such an influence could be quantified. Hence, if sea breeze- induced air pollution effects were found to be objectionable, the only control options remaining to the source operator would be: (1) process reduction during periods of high air pollution potential, or (2) installation of a stack gas cleaning device or continuous flaring operations at high burn rates.

SIMPLIFIED CALCULATION METHOD FOR DISPERSION

As noted earlier, atmospheric dispersion of a pollutant largely depends on:

1. meteorological conditions such as ambient temperature, wind speed, time of day, insulation, cloud coverage; all of which may be classified as atmospheric stability parameters, and

2. pollutant stack emission parameters such as velocity, temperature and the molecular weight of discharging vapors.

The stability of the atmosphere is largely controlled by the atmospheric thermal gradient. This is normally described by the term lapse rate, which is basically the temperature change of the surrounding air as a function of altitude. The dry adiabatic lapse rate is referred to as a neutral stability and as a reference, is a temperature gradient of -1^0 C/100 meters (i.e., air temperature decreases 1^0 C for every 100 meters of altitude). This condition implies that a volume of pollutant in air would neither gain or lose its buoyancy upon being emitted to the atmosphere. Unstable conditions with lapse rates greater than -1^0 C/100 meters add to the buoyancy of an emission, and stable conditions or inversions (lapse rates less than -1^0 C/100 meters) tend to inhibit verticle motion of the pollutant gases or plume.

Any stack should be designed based on a knowledge of prevailing meteorological conditions and stack emission criteria based on years of operating experience. Highly stable atmospheres or inversions can trap waste gases below the mass or relatively hot air, thus limiting dispersions and allowing pollutant concentrations to build up. Of greatest concern from a health risk standpoint, are high ground level concentrations (GLC's) that may occur over the short term and generally are a result of unstable atmospheric conditions. Under unstable conditions, it is more likely for atmospheric turbulence and crosswinds to carry the plume to the ground, thus exposing humans to potentially toxic emissions. These GLC's can be referred to as critical values.

Critical GLC's can usually be calculated based on a unstable atmosphere, thus enabling the designer to determine a worst case scenario. For any given day, typical atmospheric stability data can usually be obtained from a local weather bureau, or may be estimated from the so-called Pasquill chart for the appropriate Atmospheric Stability Class (refer to Table 1).

Table 1. The Pasquill Chart for Determining the Atmospheric Stability Class

Surface Wind Speed (at 10m), m sec^{-1}	**DAY**			**NIGHT**	
	Incoming Solar Radiation			**Thinly Overcast or 4/8 Low Cloud**	**3/8 Cloud**
	Strong	**Moderate**	**Slight**		
2	A	A-B	B		
2-3	A-B	B	C	E	F
3-5	B	B-C	C	D	E
5-6	C	C-D	D	D	D
6	C	D	D	D	D

The neutral class, D, should be assumed for overcast conditions during day or night

Stability Class	**Class Description**
A	Extremely unstable
B	Unstable
C	Slightly unstable

D	Neutral
E	Slightly stable
F	Stable to extremely stable

Dispersion from an elevated source is effected by the mixing and dilution of waste gases within the atmosphere. This is generally accomplished by the turbulent action of the existing gases, and the crosswind, turbulent eddy currents, wind shear, etc. At the effective stack height, pollutant gases are diluted further by increased wind speeds. Higher wind speeds make available more volumes of air to be mixed with the plume in a shorter time period. However, higher wind speeds also tend to "bend" a plume, retarding the plume's verticle motion and increasing downwind pollutant concentrations. GLC's are greater for higher wind speeds since the plume is forced to ground level before the pollutants can be dispersed over a much broader region and atmospheric volume. The need to increase the area over which a pollutant is dispersed, as well as remove the emissions from harming surrounding structures results in tall stacks and the desirability for large plume rises.

The Dispersion Process. The calculation methods to predict ambient pollutant concentrations are based on a two-step process for dispersion. First, the pollutant gases from a stack rise as a result of their own conditions of release, and then they are dispersed approximately in accordance with a Gaussian or normal distribution.

Meteorology plays an important role in determining the height to which pollutants rise and disperse. Wind speed, wind shear and turbulent eddy currents influence the interaction between the plume and surrounding atmosphere. Ambient temperatures affect the buoyancy of a plume. However, in order to make equations of a mathematical model solvable, the plume rise is assumed to be only a function of the emission conditions of release, and many other effects are considered insignificant.

Plume Rise

The vertical motion of the plume to the height where it becomes horizontal is known as the plume rise, (refer back to Figure 1). The plume rise is assumed to be a function primarily of the emission conditions of release, (i.e. velocity and temperature characteristics). A velocity in the vertical plane gives the gases an upward momentum causing the plume to rise until atmospheric turbulence disrupts the integrity of the plume. At this point the plume ceases to rise. This is known as the momentum plume rise. Stack gas exiting temperatures are usually much greater than ambient making them less dense than the surrounding air. This difference in densities gives the gases buoyancy, allowing the plume to rise until it is cooled by the atmosphere, reducing the density differential to zero. This is known as the thermal plume raise. The momentum and thermal plume rises combine to produce the plume rise of an emission. These effects are not independent: gases with a high exiting velocity are cooled faster as a result of more atmospheric mixing of the plume. The thermal buoyancy contribution to plume rise can, therefore, be lessened by increasing exiting velocities. Low exiting velocities can cause the plume to become trapped in the turbulent wake along the side of the stack, and fall rapidly to the ground (referred to as fumigation). Fumigation can usually be prevented by keeping the emission velocity greater than 10 meters/second. An emission velocity that is one and one-half times greater than the atmospheric crosswind is generally accepted as a safety factor to prevent fumigation.

In many calculation methods, the momentum contributions to plume rise are considered negligible when compared to the thermal plume rise, and hence are ignored.

Effective Stack Height

The importance of plume rise is that it determines the effective stack height, or the height at which most calculation procedures assume dispersion to initiate. The plume rise added to the actual height of the stack is known as the effective stack height, described by the following expression:

$$h = h_s + \Delta h \tag{1}$$

where h = effective stack height (m)
h_S = actual stack height (m)
Δh = plume rise (m)

Plume rise can be estimated from empirical relationships; some of the more well known expressions are summarized in Table 2.

At the effective stack height, the dispersion of the pollutants are assumed to spread out as a Gaussian distribution. The basic dispersion equation considers

Table 2. Maximum Ground Level Concentration for A 5-min. Blow (Worst Case)

Author	Expression	Comment
1. Holland	$\Delta h = \frac{V_s D K_s}{\overline{u}_s}\left(1.5 + 2.71 D\left(\frac{T_s - T_a}{T_s}\right)\right)$	Highly empirical formulation of limited applicability.
2. Concawe	$\Delta h = \frac{5.53\, Q_h^{1/2}}{u^{3/4}}$	Basically a regression equation. Suited more to large buoyant plume applications.
3. Stümpke	$\Delta h = \frac{d}{u}\left(1.5 V_s + 65 d^{1/2}\left(\frac{T_a - T_s}{T_s}\right)\right)$ T_a = ambient air temperature (K) T_s = stack gas temperature (K)	Essentially the same as Holland's formula, except buoyancy term depends on 1/4 power.
4. Lucas-Moore-Spurr	$\Delta h = \frac{135\, Q_h^{1/4}}{u}$	Regression equation based on work of Priestly.
5. Rauch	$\Delta h = \frac{47.2\, Q_h^{1/4}}{u}$	Same as Lucas et al., but different data base.
6. Stone-Clark	$\Delta h = \left(104.2 + 0.171 h_p\right)\frac{Q_h^{1/4}}{u}$ h_p = physical stack height (m)	A modification of the Lucas-Moore-Spurr equation to account for the effect of the physical stack.
7. Moses and Carson	$\Delta h = \frac{A}{u}\left(-0.029 V_s d + 5.53\, Q_h^{1/2}\right)$ A = coefficient dependent on atmospheric stability **Stability** **A** Unstable 2.65 Neutral 1.08 Stable 0.68	Regression equation developed from many data sources.

Table 2 continued.

8. Briggs	a. Unstable, neutral conditions: $\Delta h = 0.25\, Q_h^{1/3} h_p^{2/3} u$ b. Stable conditions: $\Delta h = 0.296 \left(\frac{Q_h}{u \left(\frac{\partial \theta}{\partial z} \right)} \right)^{1/3}$ $\frac{\partial \theta}{\partial z} =$ variations of potential temperature with height ≈ 0.03 K/m	Nonempirical formulation.

a continuously emitting point source emanating through a coordinate system with its origin at the base of the source as shown in Figure 1.

Emissions of gases or particles less than 20 microns (larger particles settle more quickly due to gravitational effects) disperse with an origin and plume centerline at the effective stack height. Pollutant concentrations are greatest within one standard deviation of the plume centerline. Thus, the determination of the value of these standard deviations is an important factor in calculating ambient concentrations.

The standard deviations in the vertical direction σ_z and in the horizontal direction σ_y, of the dispersing plume along the centerline are functions of meteorological conditions and depending on the evaluation method, downwind distance also. These dispersion parameters and the effective stack height can be calculated in a number of ways, using various empirical constants. Each manner of calculating these parameters, σ_z, and σ_y defines a different calculation method without disrupting the basic Gaussian calculation procedures. These methods are summarized as follows: The basic dispersion equation to calculate ground level concentrations directly downwind from a point source is:

$$\chi = \frac{Q}{\pi \bar{u}_s \sigma_z \sigma_y} \exp - \left(\frac{h^2}{2 \sigma_z^2} \right) \qquad (2)$$

where: χ = ground level concentration (gm/m^3)

Q = pollutant exiting rate (gm/sec)

σ_z, σ_y = horizontal and vertical plume standard deviations (m), hence are a function of distance x downwind from the point source.

$\overline{u}_s$ = mean wind speed at height of stack (m/sec)

h = effective stack height (m)

The maximum GLC is of major importance to the refinery in determining whether or not emissions meet maximum permissible discharge concentrations. It can be calculated as follows:

$$\chi_{max} = \frac{2Q}{e \pi \overline{u}_s h^2} \frac{\sigma_z}{\sigma_y} \qquad (3)$$

where: e = 2.7181

Three commonly used dispersion calculation methods for the prediction of ground level concentrations are based on the above expression. The variance in each method is the calculation of plume rise, Δh , and the horizontal and vertical plume dispersion parameters. These methods are:

1) The ASME plume rise equations and the ASME dispersion parameters.

2) The Pasquill-Gifford dispersion parameters and Brigg's plume rise equations.

3) The Pasquill-Gifford dispersion parameters and Holland's plume rise equations.

The parameters σ_Z and σ_y are determined through the use of one of the above calculation procedures and substituted into equations 1, 2 and/or 3 to obtain GLC's .

The ASME Dispersion Calculation Method

ASME Plume Rise Equations

The ASME method is one of the few calculation methods to consider emissions having relatively low exiting temperatures and relatively high exiting velocities. Under these conditions of release the momentum effects of the plume dominate over thermal, thus the momentum plume rise should be used in the GLC calculations.

If $V_5 \geq 10$m/sec and $T_S > 50^0\,K + T_a$ then

$$\Delta h = D \left(\frac{V_s}{\overline{u}_s} \right)^{1.4} \tag{4}$$

where Δh = the plume rise (m)
D = stack diameter (m)
V_s = stack gas exiting velocity (m/sec)
u_s = mean wind speed at height of stack (m/s)
T_s = stack gas exiting temperature °K
T_a = ambient temperature °K

The thermal plume rise is based on the relative buoyancy of a plume to the surrounding atmosphere. When the stack gas exiting temperatures are 50° C greater than ambient temperature and relatively large volumes of gases are being discharged, the following set of equations should be used to determine plume rise.

For stable atmospheric conditions:

$$\Delta h = 2.9 \left(\frac{F}{\overline{u}_s G} \right)^{\frac{1}{3}} \tag{5A}$$

where:

$$F = 2.45\ V_s D^2 \left(\frac{T_s - T_a}{T_s} \right) \tag{5B}$$

$$G = \frac{9.8}{T_a} \left(\frac{\Delta T}{\Delta z} + 0.98 \right) \tag{5C}$$

$\frac{\Delta T}{\Delta z}$ = The atmospheric temperature gradient, °K/100m

For neutral or unstable atmospheric conditions:

$$h = \frac{7.4\, h_s^{2/3} F^{1/3}}{\overline{u}_s} \tag{6}$$

The ASME Dispersion Parameters. The horizonal and vertical dispersion parameters are represented by the following empirical power law equation

$$\sigma_z,\ \sigma_y = a x^b \tag{7}$$

where a,b= the empirical ASME dispersion constants given in Table 2.

x = downwind distance from the source

Table 3. ASME Dispersion Constants, a and b.

Atmospheric Stability	σ_y		σ_z	
	a	b	a	b
very unstable	0.40	0.91	0.40	0.91
unstable	0.31	0.86	0.33	0.86
neutral	0.32	0.78	0.22	0.78
stable	0.31	0.71	0.06	0.71

The Pasquill-Gifford-Briggs dispersion method:

Brigg's Plume Rise Equation
for $x \leq 3.5W$

$$\Delta h = \frac{1.6\, F^{1/3} x^{2/3}}{\overline{u}_s} \tag{8}$$

and $x > 3.5W$

$$\Delta h = \frac{1.6\,(F^{1.3})(3.5\,W)^{2/3}}{\bar{u}_s} \tag{9}$$

$W = 14F^{5/8}$ if $F \leq 55$

$W = 34F^{2/3}$ if $F > 55$

where $$F = 2.45\,V_s\,D^2 \left(\frac{T_s - T_a}{T_s} \right)$$

x = distance between source and receptor.

The Pasquill-Gifford dispersion parameters are functions of downwind distance and meteorological conditions. The parameters, σ_z and σ_y may be obtained from Figure 6(a) and 6(b) respectively. The user must know the atmospheric stability as well as downwind distance from the source to select the appropriate dispersion parameters

The Pasquill-Gifford-Holland Dispersion Method.

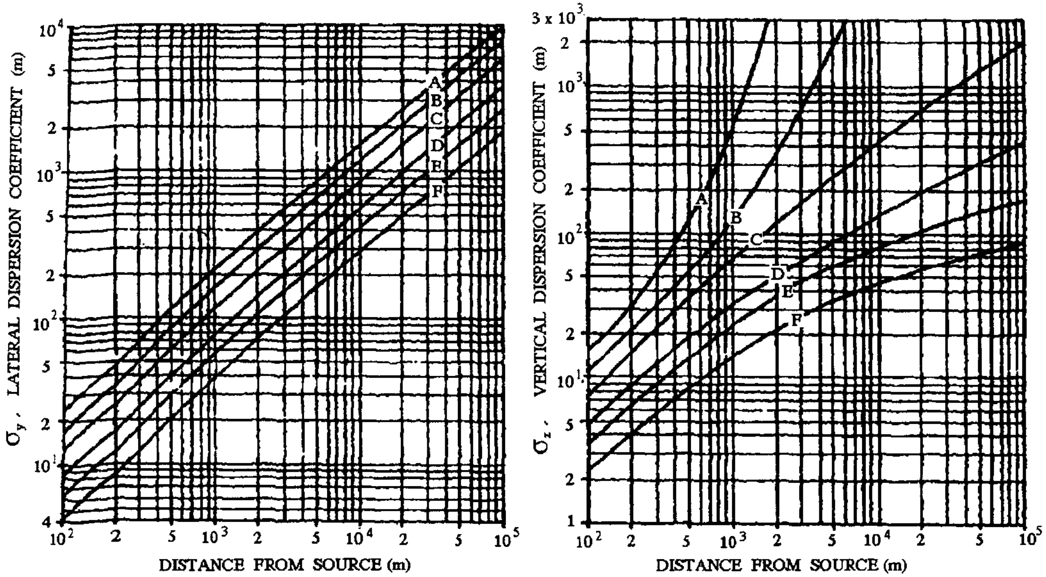
Holland's Plume Rise equation: for $x > 300$ meters

$$\Delta h = \frac{V_s D K_s}{\bar{u}_s} \left(1.5 + 2.71\,D \left(\frac{T_s - T_a}{T_s} \right) \right) \tag{10}$$

where K_s = stability class factor given in Table 4.

Table 4. Atmospheric Stability Class Factor.

Atmospheric Stability	K_s
slightly unstable	1.1
unstable	1.2



(a)

(b)

Figure 6. The Pasquill-Gifford dispersion coefficients versus downwind distance for various dispersion classes. (*a*) The lateral dispersion coefficient and (*b*) the vertical dispersion coefficient are plotted against x.

Table 4 continued.

neutral	1.0
slightly stable	0.9
stable	0.8

This calculation method uses the same Pasquill-Gifford dispersion parameters as the Pasquill-Gifford-Briggs calculation method. Therefore, the dispersion parameters σ_y and σ_z for this method may be obtained from Figures 6(a) and 6(b) respectively.

Averaging Time

The calculation procedures just described are methods for predicting short-term GLC's for single stack cases. In the calculation of short-term GLC's it is assumed that meteorological conditions are constant throughout the measuring time period. Short-term time periods are usually considered to be 1 hour or less. The averaging times for the various methods discussed are as follows:

Method	**Averaging Time**
ASME	60 minutes
Pasquill-Gifford-Briggs	10 minutes
Pasquill-Gifford-Holland	10 minutes

The problem the plant operator faces is that governmental regulations may specify 10, 15, 30 or 60 minute averaging times and one of the above calculation procedures. A simple means of converting from one averaging time to another in order to compare methods, regulations or calculated and measured GLC's of different averaging times is given below:

$$\frac{\chi_1}{\chi_2} = \left(\frac{t_1}{t_2} \right)^{-\alpha} \tag{11}$$

where
χ_1 = GLC of averaging time period t_1
χ_2 = GLC of averaging time period t_2
t_1 = averaging time period 1
t_2 = averaging time period 2

α = an empirical constant equal to 0.17

The above calculation procedures can be used to assess several dispersion problems including the determination of:

1) GLC's effected by the current refinery practices.
2) GLC's effected by plant process modifications, i.e. emission quantities, velocities and temperatures.
3) GLC's effected by plant process additions.
4) stack blower requirements and process operating conditions.
5) minimum stack height requirements to meet safe exposure limits.

The maximum GLC as determined by equation 3 is dependent on specific meteorological conditions and downwind distance from the source. The critical GLC of a source is the maximum possible GLC that can occur regardless of meteorological conditions and downwind distance.

The determination of the critical GLC is a trial and error computation of GLC's due to various wind speeds, atmospheric stabilities and downwind distances. The maximum value obtained from these procedures is the critical GLC. Because of the number of computations involved, calculations should be performed on the computer. Software simulation is also necessary to calculate GLC's due to multiple stack cases. Wind direction is an additional variable that must be taken into account with multiple stact cases.

Long-Term Average GLC's

Long-term averaging periods are normally considered to be 24 hours or greater which is too long a time interval to assume constant meteorology. Therefore, long-term average GLC's are calculated with the use of actual meteorological data to predict monthly, seasonal or annual average ambient concentrations.

The weather data are based on thousands of observations of wind speed, wind direction and atmospheric stability taken over the desired averaging interval at local weather bureau stations.

Software system generally include one or more of the GLC's calculation methods previously discussed. Input requirements would also include:

1) A coordinate system to identify stack and receptor site positions.
2) Stack coordinates.
3) Average stack emission parameters.
4) Meteorological data for the desired averaging period.
5) Receptor site coordinates.

Hundreds of computations are needed to generate the long-term average ambient concentration at various receptor sites due to one or more continuously discharging stacks.

EXAMPLE OF REACTOR VENTING EPISODE

This section illustrates by way of example, the application of simplified dispersion estimates to assessing a catastrophic venting operation. In this example, an analysis was performed to predict the fate of air pollutants, specifically vinyl chloride monomer (VCM), originating from an episode type upset (reactor blow) condition from a reaction vessel.

Principal emphasis of the analysis is an estimation of ground level conditions within proximity of the plant and the intermediate surrounding community within a 5-mile radius of the plant, and potential community impact.

The basis of the study was to predict ground level concentrations based on a worst/worst case or "episode" condition scenario, i.e., the entire contents of a reactor venting in three (3) and five (5) minutes and estimating the concentrations over a broad range of weather conditions and surface wind speeds. Ground level concentrations were calculated by predictive means of computation and compared on several bases. The figures presented are based on such computations and best professional judgment analysis to show the degree to which ground level concentrations may be anticipated under the extreme conditions assumed, and from which a relative risk factor can be assigned. Values computed and worst case conditions summarized are supported in data shown in the discussions below.

In general, ground level concentrations resulting from stacks are a function of meteorological conditions such as stability of the atmosphere, wind speed and direction, atmospheric mixing height and ambient air temperature, stack height, stack diameter, exit stack gas speed and temperature and other factors. The peak emission rate over a three and five minute venting of an entire reactor contents (extreme condition of failure) is defined in Table 5.

Based on these peak emission rates and other parameters the data shown and computed were developed.

Table 5. Initial Data Sheet for Reactor Venting Episode

Stack Height 45 ft. (13.72m) (above grade)
Stack Diameter 8 in. (0.203m)
VCM Specific Gravity ≈ 2.2/Density = 2.467 kg/m^3

Item	**Charge to Reactor**		**Q(kg/s)**	
	(Lbs)	**(kg)**	**3 Min. Venting**	**5 Min. Venting**
VCM	≈13,000	5,900	32.6	19.6
V. Acet.	≈2,300	1,000*	5.7	3.5
Chain Modifier	≈200	100*	0.59	0.35
H_2O	≈21,600	9,800*	54.5	32.7
Total	≈37,000	16,800	93.4	56.0
Discharge Velocity (m/s)			1,170	700

Mean Ambient Temp. 50^0 F (283^0 K)
Stack Effluent Temp.155^0 F (341.3^0 K)
Molecular Wt. VCM 62.50.
To convert mg/m^3 multiply by 0.3584 at STP. to ppm.
To convert mg/m^3 multiply by 0.3584 at to ppm.
*Considered as inerts from an air pollution standpoint.

Tables 6 provides a summary of the predicted GLC's under worst case meteorological conditions. Note that the ground level concentrations summarized do not consider wind direction nor variation. More specifically, two points within the plant vicinity are presented to summarize wind persistence as related to known landmarks:

Direction (Wind From)	**% of Time**	**Landmark**	**Distance**
NE	5.2	High-school	<0.5 mi.
NW	7.7	Health Center	1-1.5 mi.

Maximum ground level concentrations estimated from the (episode) case condition assumed herewith within close proximity of the plant (up to 5-miles) range as follows:

Within one hour of incident 1.6 to 47.0 ppm.
Within three hours of incident 1.3 to 39.5 ppm.
Within twenty-four hours of incident 0.3- 9.9 ppm.
This additionally assumes no change in weather conditions or compensatory diffusion/dispersion within the atmosphere.

Alternative Considerations and Qualifications

Because of extreme venting conditions assumed, effective stack heights and resultant plumes from both 3- and 5-minute discharge conditions attain heights beyond the micro-meteorological conditions assumed in accepted computation models. It is therefore highly probable there will be considerably further atmospheric dispersion and diffusion of the VCM than predicted in the results shown. That is, the ground level concentration can be expected to be considerably lower than the values shown in Table 6.

Table 6. Summary of Predicted/estimated Ground Level Concentration at Worst (Episode) Case Conditions and Weather.

Stability Class	Class Description	Surface Wind Speed (MPH)	Worst Case Ground Level Concentration (ppm)			Distance From Plant (miles)
			1 hour	3 hours	24 hours	
B	unstable	10.2	34.6	29.1	8.7	0.5
B	-//-	4.08	23.6	19.8	5.0	1.0
B	-//-	2.04	14.9	12.5	3.1	1.5
B	-//-	1.0	11.6	9.7	2.4	2.0
B	-//-	1.0	1.6	1.3	0.3	5.0
C	slightly unstable	16.4	47.0	39.5	9.9	0.5
C	-//-	10.2	30.7	25.8	6.5	1.0
C	-//-	6.1	20.7	17.4	4.4	1.5
C	-//-	10.5	16.9	14.2	3.6	2.0
C	-//-	4.08	6.5	5.5	1.4	5.0
D	neutral	30.7	5.9	5.0	1.3	0.5
D	-//-	20.5	41.8	35.1	8.8	1.0
D	-//-	16.4	30.5	25.6	6.4	1.5

Table 6 continued.

Stability Class	Class Description	Surface Wind Speed (MPH)	Worst Case Ground Level Concentration (ppm)			Distance From Plant (miles)
			1 hour	3 hours	24 hours	
D	-//-	10.2	23.2	19.5	4.9	2.0
D	-//-	6.1	12.3	10.3	2.6	5.0

There is some question as to whether the total reactor contents will be discharged as assumed. If the reaction is in progress, then material already will have polymerized and will most probably not vent to the atmosphere, thus resulting in a significantly smaller emission.

In the calculations that were made to predict ground level concentrations from a VCM reactor blow off, the Pasquill-Gifford-Holland dispersion model was used as a basis for these estimations. Calculations were made for six different stability classes and ground level concentrations, and at various distances from the point source of emission.

No correlation was made initially as to wind direction, nor to probability of any wind direction/weather condition or percent time of occurrence, however, this is certainly an important factor in the probability of the pollutant being at a concentration predicted. The greatest significance is attached to predicting an ultimate ground level concentration from any potential episode.

The range of ground level concentrations varies widely depending on wind speed, distance from the emission source, duration of the emission.

Assumptions and Model Predictions

The Pasquill-Gifford-Holland dispersion method was used for the above example. Note that vinyl chloride monomer (VCM) constitutes the primary active ingredient in the reactor. It was therefore assumed that:

(a) the density of the gaseous effluent mixture was that of VCM,

and

(b) ground level concentrations and maximum GLC's were computed for VCM alone.

Calculations were performed for two different venting rates, namely 3 and 5 minutes. Note that the calculation procedure predicts short-term GLC's for the single stack. This means that meteorological conditions are assumed to be constant throughout the measuring time period.

Tables 7 and 8 give the short-term ground level concentrations for a 3-minute venting at 1/2, 1.0, 1-1/2, 2.0, and 5 miles from the discharge point respectively. The calculations assume meteorological conditions to be constant for approximately 10-minutes. Values in Tables 7 and 8 were computed for the six stability classes over a range of wind speeds.

Table 7. Short-term (10 Min. Period) Ground Level Concentration for a 3-min. Blow

(Shaded boxes indicate worst case conditions)

Stability	Δ h (m)	h_e (m)	Wind Speed (m/s)	Ground Level Concentrations (mg/m^3)				
				0.5 mi.	1.0 mi.	1.5 mi.	2.0 mi.	5.0 mi.
A	907.7	921.4	0.5	5.79	30.66	20.74	11.00	0.787
A	567.3	581.0	0.8	44.57	20.73	13.82	7.07	0.492
A	453.8	467.5	1.0	63.74	16.89	11.22	5.70	0.394
A	302.5	316.2	1.5	75.82	11.47	7.59	3.82	0.262
A	226.9	240.6	2.0	69.84	8.66	5.72	2.87	0.197
A	181.5	195.2	2.5	61.54	6.95	4.59	2.30	0.157
B	907.7	921.4	0.5	2.3×10^{-18}	0.139	22.35	30.64	4.09
B	567.3	581.0	0.8	5.37×10^{-6}	10.90	38.86	29.48	2.58
B	453.8	467.5	1.0	3.61×10^{-3}	26.86	39.44	26.07	2.07
B	302.5	316.2	1.5	1.98	54.92	33.33	19.21	1.38
B	226.9	240.6	2.0	16.05	61.31	27.19	14.93	1.04
B	181.5	195.2	2.5	39.34	59.14	22.63	12.14	0.831
B	151.2	164.9	3.0	60.78	54.63	19.27	10.21	0.692
B	113.4	127.1	4.0	85.12	45.48	14.17	7.73	0.519
B	90.7	104.5	5.0	91.58	38.23	11.95	6.21	0.415
C	208.0	221.7	2.0	0.301	26.70	42.32	41.14	17.33
C	166.4	180.1	2.5	3.81	50.10	51.86	44.74	14.84
C	138.7	152.4	3.0	14.55	66.79	54.66	44.15	12.84
C	104.0	117.7	4.0	51.39	80.81	52.07	39.34	10.01
C	83.2	96.9	5.0	85.97	81.18	46.68	34.16	8.15
C	69.3	83.0	6.0	108.31	76.85	41.46	29.80	6.86
C	52.0	65.7	8.0	124.28	65.72	33.21	23.43	5.20
D	189.1	202.8	2.0	1.25×10^{-9}	0.0168	1.79	4.26	23.04
D	126.1	139.8	3.0	7.95×10^{-4}	3.83	23.87	29.28	32.51

Table 7 continued.

Stability	Δ h (m)	h_e (m)	Wind Speed (m/s)	Ground Level Concentrations (mg/m^3)				
				0.5 mi.	1.0 mi.	1.5 mi.	2.0 mi.	5.0 mi.
D	75.5	89.4	5.0	0.755	52.23	71.26	61.26	29.13
D	47.3	61.0	8.0	7.197	103.61	80.61	60.76	21.11
D	37.8	51.5	10.0	11.38	110.67	74.81	54.56	17.53
D	25.2	38.9	15.0	15.69	100.39	58.42	41.14	12.16
D	18.9	32.6	20.0	15.68	85.05	46.64	32.40	9.26
E	170.2	183.9	2.0	2.70×10^{-17}	4.72×10^{-5}	0.112	0.439	13.25
E	113.5	127.2	3.0	7.33×10^{-7}	0.304	8.81	13.45	36.31
E	68.1	81.8	5.0	0.224	25.43	68.74	63.27	46.59
E	42.5	56.3	8.0	18.41	99.32	111.35	84.92	38.61
E	34.0	47.7	10.0	50.82	126.54	113.45	82.49	33.18
E	22.7	36.4	15.0	126.36	138.31	97.80	67.59	23.87
E	17.0	30.7	20.0	160.96	126.57	81.34	55.09	18.46
F	151.3	165.0	2.0	1.23×10^{-37}	2.9×10^{-15}	$1.45 \times 10^{-}$	$5.08 \times 10^{-}$	0.239
F	100.9	114.6	3.0	1.46×10^{-16}	5.42×10^{-6}	0.0199	0.271	8.66
F	60.5	74.2	5.0	2.78×10^{-5}	0.421	8.97	21.08	45.13
F	37.8	51.5	8.0	0.349	21.48	66.76	81.31	63.33
F	30.3	44.0	10.0	3.36	51.91	99.46	102.71	62.07
F	20.2	33.9	15.0	34.40	116.53	131.3	113.09	51.72
F	15.1	28.8	20.0	78.32	143.1	129.95	104.0	42.47
F	10.1	23.8	30.0	130.1	143.2	108.85	82.0	30.50

Table 8. Short-term (10 Min. Period) Ground Level Concentration for 5-min. Blow

Stability	Δ h (m)	h_e (m)	Wind Speed (m/s)	Ground Level Concentrations (mg/m^3)				
				0.5 mi	1.0 mi.	1.5 mi.	2.0 mi.	5.0 mi.
A	544.6	558.3	0.5	48.56	19.99	13.31	6.80	0.479
A	340.3	354.0	0.8	75.41	12.86	8.52	4.29	0.299
A	272.2	285.9	1.0	74.63	10.35	6.85	3.44	0.239
A	181.5	195.2	1.5	61.57	6.90	4.59	2.30	0.160
A	136.1	149.8	2.0	49.85	5.22	3.45	1.73	0.120
A	108.9	122.6	2.5	41.35	4.18	2.76	1.38	0.096

Table 8 continued.

Stability	Δ h (m)	h_e (m)	Wind Speed (m/s)	Ground Level Concentrations (mg/m^3)				
				0.5 mi	1.0 mi.	1.5 mi.	2.0 mi.	5.0 mi.
B	544.6	558.3	0.5	2.23×10^{-5}	9.77×10^{-5}	39.29	28.93	2.48
B	340.3	354.0	0.8	0.530	0.822	35.65	21.16	1.55
B	272.2	285.9	1.0	4.996	6.08	31.12	18.81	1.24
B	181.5	195.2	1.5	39.34	37.53	22.64	12.15	0.831
B	136.1	149.8	2.0	71.59	62.59	17.52	9.23	0.623
B	108.9	122.6	2.5	87.1	73.07	14.22	7.43	0.499
B	90.71	104.4	3.0	91.7	75.18	11.95	6.21	0.416
B	68.0	81.7	4.0	87.35	69.94	9.04	4.68	0.312
B	54.4	68.1	5.0	78.42	62.08	7.26	3.75	0.249
C	124.8	138.5	2.0	25.56	73.90	54.55	42.81	11.75
C	99.8	113.5	2.5	57.93	81.53	51.25	38.45	9.65
C	83.2	96.9	3.0	85.99	81.19	46.69	34.16	8.16
C	62.4	76.1	4.0	116.9	73.17	38.38	27.37	6.21
C	49.9	63.6	5.0	124.8	63.99	32.11	22.61	5.00
C	41.5	55.3	6.0	122.4	56.08	27.44	19.19	4.18
C	31.2	44.9	8.0	109.1	44.35	21.13	14.67	3.15
D	113.4	127.1	2.0	6.22×10^{-3}	8.74	0.111	6.63	32.92
D	75.6	89.3	3.0	0.764	53.23	6.160	25.92	29.15
D	45.3	59.02	5.0	8.04	107.57	41.94	41.07	20.44
D	28.4	42.1	8.0	14.97	106.79	66.13	37.17	13.56
D	22.7	36.4	10.0	15.93	96.42	67.39	32.75	11.02
D	15.1	28.8	15.0	14.56	73.50	58.73	24.31	7.47
D	11.3	25.0	20.0	12.44	58.26	49.20	19.05	5.64
E	102.1	115.8	2.0	3.06×10^{-5}	1.157	16.76	22.09	40.80
E	68.1	81.8	3.0	0.224	25.44	68.76	63.29	46.60
E	40.8	54.5	5.0	23.30	105.8	112.8	85.15	37.67
E	25.5	39.2	8.0	106.05	139.4	103.9	72.64	26.41
E	20.4	34.1	10.0	142.4	135.5	91.98	63.02	21.77
E	13.6	27.3	15.0	169.2	112.2	68.65	46.0	15.00
E	10.2	23.9	20.0	161.5	92.15	53.97	35.83	11.41
F	90.7	104.4	2.0	3.07×10^{-13}	1.53×10^{-4}	0.125	1.014	14.68
F	60.5	74.2	3.0	2.78×10^{-5}	0.421	8.98	21.09	45.14

Table 8 continued.

Stability	Δ h (m)	h_e (m)	Wind Speed	Ground Level Concentrations (mg/m^3)				
				0.5 mi	1.0 mi.	1.5 mi.	2.0 mi.	5.0 mi.
F	36.3	50.0	5.0	0.569	26,09	73.2	86.04	63.49
F	22.7	36.4	8.0	21.03	99.98	126.8	113.8	55.37
F	18.2	31.9	10.0	48.91	128.5	132.5	110.7	48.33
F	12.1	25.8	15.0	110.7	147.5	119.9	92.4	35.60
F	9.06	22.8	20.0	137.7	138.5	102.0	76.05	27.82
F	6.06	19.78	30.0	143.5	112.6	76.05	55.0	19.24

Tables 9 and 10 give the maximum ground level concentrations expected for worst case conditions for the 3-min. and 5-min. venting rates, respectively; worst case conditions were obtained from Tables 7 and 8 (i.e., values were calculated based on the largest concentration found in Tables 7 and 8 for each distance within a stability class).

Table 9. Maximum Ground Level Concentration for a 3-min. Blow (Worst Case)

Stability	Wind Speed (m/sec)	Maximum Concentrations (mg/m^3)	Distance of Maximum		h_e (m)	Plume Height (m)
			(m)	(mi.)		
A	1.5	93.12	805	0.5	316.2	302.5
A	0.5	98.14	1609.7	1.0	921.4	907.7
A	0.5	79.96	2415	1.5	921.4	907.7
A	0.5	89.96	3219	2.0	921.4	907.7
A	0.5	276.8	8094	5.0	921.4	907.7
B	5.0	101.1	805	0.5	104.5	90.7
B	2.0	72.25	1609.7	1.0	240.6	226.9
B	1.0	51.4	2415	1.5	467.5	453.8
B	0.5	32.22	3219	2.0	921.4	907.7
B	0.5	90.05	8049	5.0	921.4	907.7
C	8.0	130.8	805	0.5	65.7	52.0
C	4.0	85.3	1609.7	1.0	117.7	104.0
C	3.0	61.4	2415	1.5	152.4	138.7
C	2.5	50.1	3219	2.0	180.1	166.4
C	2.0	38.8	8049	5.0	221.7	208.0
D	15.0	15.70	805	0.5	38.9	25.2

Table 9 continued.

Stability	Wind Speed (m/sec)	Maximum Concentrations (mg/m^3)	Distance of Maximum		h_e (m)	Plume Height (m)
			(m)	(mi.)		
D	10.0	117.9	1609.7	1.0	51.5	37.8
D	8.0	96.21	2415	1.5	61.0	47.3
D	5.0	61.9	32.19	2.0	89.4	75.6
D	3.0	34.73	80.49	5.0	139.8	126.1
E	20.0	167.3	805	0.5	30.7	17.0
E	15.0	147.1	1609.7	1.0	36.4	22.7
E	10.0	125.5	2415	1.5	47.7	34.0
E	8.0	90.35	3219	2.0	50.3	42.5
E	5.0	54.6	8049	5.0	81.8	68.1
F	30.0	174.0	805	0.5	23.8	10.1
F	15.0	142.4	1609.7	1.0	33.9	20.2
F	15.0	131.3	2415	1.5	33.9	20.2
F	15.0	118.11	3219	2.0	33.9	20.2
F	8.0	65.73	8049	5.0	51.5	37.8

Table 10. Maximum Ground Level Concentration for A 5-min. Blow (Worst Case)

Stability	Wind Speed (m/sec)	Maximum Concentrations (mg/m^3)	Distance of Maximum		h_e (m)	Plume Height (m)
			(m)	(mi.)		
A	0.8	83.6	805	0.5	354.0	340.0
A	0.5	160.4	1609.7	1.0	558.3	544.6
A	0.5	137.4	2415	1.5	558.3	544.6
A	0.5	154.5	3219	2.0	558.3	544.6
A	0.5	237.7	8049	5.0	558.3	544.6
B	3.0	101.3	805	0.5	104.4	90.71
B	3.0	153.5	1609.7	1.0	104.4	90.71
B	0.5	43.25	2415	1.5	558.3	544.6
B	0.5	52.66	3219	2.0	558.3	544.6
B	0.5	147.0	8049	5.0	558.3	544.6
C	4.0	117.0	805	0.5	76.1	62.4
C	2.5	88.0	1609.7	1.0	113.5	99.8

Table 10 continued.

Stability	Wind Speed (m/sec)	Maximum Concentrations (mg/m³)	Distance of Maximum		h_e (m)	Plume Height (m)
			(m)	(mi.)		
C	2.0	66.9	2415	1.5	138.5	124.8
C	2.0	63.6	3219	2.0	138.5	124.8
C	2.0	59.7	8049	5.0	138.5	124.8
D	10.0	16.1	805	0.5	36.4	22.7
D	5.0	107.8	1609.7	1.0	59.02	45.3
D	10.0	129.7	2415	1.5	36.4	22.7
D	5.0	85.2	3219	2.0	59.02	45.3
D	2.0	37.8	8049	5.0	127.1	113.4
E	15.0	169.3	805	0.5	27.3	13.1
E	8.0	142.7	1609.7	1.0	39.2	25.5
E	5.0	115.4	2415	1.5	54.5	40.8
E	5.0	92.6	3219	2.0	54.5	40.8
E	3.0	54.6	8049	5.0	81.8	68.1
F	30.0	151.1	805	0.5	19.78	6.06
F	15.0	147.5	1609.7	1.0	25.8	12.1
F	10.0	133.4	2415	1.5	31.9	18.2
F	8.0	115.3	3219	2.0	36.4	22.7
F	5.0	66.9	8049	5.0	50.0	36.3

SAMPLE CALCULATION

We shall now provide a second example to illustrate step-by-step calculations. In this example a flare stack is estimated to be 80% efficient in combusting H_2S off-gas. The total off-gas through the stack is 400,000 kg/hr, of which 7.0 weight percent is H_2S. The physical stack height is 250 m, the stack diameter is 5.5 m, and the stack emission velocity is 18 m/s. The stack emission temperature is 15°C. The meteorological conditions may be described as a bright sunny day with a mean wind speed of 3 m/s.

a)We wish to determine the ground level concentration of H_2S at a point 2000 meters downwind from the source.

b) Determine the maximum GLC for the above emission parameters and atmospheric conditions.

Part(a) Solution:

Step 1 - Determine the pollutant quantity source strength

$$Q = 400{,}000\ \frac{kg}{hr} \times 0.070 \times \frac{288^o\ K}{273^o\ K} \times \frac{l\ hour}{3600 sec} \times 0.80$$

$$= 6.564 kg\ H_2 S \times 1000 gm/kg = 6{,}564 gm\ H_2 S/sec$$

Step 2- Determine the atmospheric stability class:

From Pasquill's chart (Table 1), for a bright summer day with windspeeds of 3 m/s, the atmospheric conditions can be described as Class B, unstable.

Step 3- Determine the plume rise:

In this example, we shall apply the Holland plume rise equation (Equation 10). There are, however other expressions that may be used (refer to Table 2 for other formulae):

$$\Delta h = \frac{V_s D K_s}{\overline{u}_s} \left(1.5 + 2.71 D \left(\frac{T_s - T_a}{T_s} \right) \right) =$$

$$= \frac{(18\, m/s)\, (5.5m)\, (1.2)}{3\, m/s} \left(1.5 + (2.71)\, (5.5\, m) \left(\frac{(140 - 15)^o\ K}{(140 + 273)^o K} \right) \right)$$

$$= 238.0\ m$$

Step 4- Determine the effective stack height

$$h = \Delta h + h_s = 238.0 + 250 = 488\ m$$

Step 5- Obtain the dispersion standard deviation coefficients and compute the GLC downwind from the source.

From Figures 2 and 3, the Pasquill-Gifford dispersion coefficients are obtained for a downwind distance of 2000 meters and for atmospheric stability Class B.

$\sigma_Z = 230$ m and $\sigma_y = 300$ m.

The ground level concentration (GLC) 2000 meters downwind from the source can now be computed from Equation (2):

$$\chi = \frac{Q}{\pi \bar{u}_s \sigma_y \sigma_z} \, exp - \left(\frac{h^2}{2\sigma_z^2} \right) =$$

$$= \frac{6564\, gm/s}{\pi\,(3m/s)\,(300m)\,(230m)} \, exp - \left(\frac{(488\,m)^2}{2\,(230m)^2} \right) =$$

$$= 0.0101 \times exp - (2.2509) \times 10^6 \; \mu g/gm.$$

Part (b) Solution:
The maximum GLC for the above meteorological conditions is estimated by applying Equation 3:

$$\chi_{max} = \frac{2Q}{e \pi \bar{u}_s h^2} \frac{\sigma_z}{\sigma_y}$$

$$= \frac{2\,(6{,}564\, gm/s)\,(230\,m)}{(e)\,(\pi)\,(3\,m/s)\,(488\,m)^2\,(300\,m)}$$

$$= 0.00165\, gm/m^3 \;\; H_2S$$

or $\chi_{max} = 1.65 \times 10^3 \mu g/m^3 \;\; H_2S$

The Pasquill-Gifford-Holland method predicts GLC's based upon 10 minute averaging times. However, what is also of interest is the concentration of H_2S

based on a 3 hour averaging time. The results may be converted as follows:

$$\frac{\chi_1}{\chi_2} = \left(\frac{t_1}{t_2}\right)^{-0.17}$$

$$\chi_1 = \left(\frac{3\ hour \times \dfrac{60\ minutes}{1\ hour}}{10\ minutes}\right)^{-0.17} (1.65\ \mu g/m^3\ \ H_2\ S)$$

$$= 1009\ \mu g/m^3\ H_2\ S$$

This concentration should be compared against permissible exposure limits established by OSHA (Occupational Safety and Health Act) and allowable discharges limits set by federal (EPA) and local emissions regulations.

It should be remembered that the concentrations predicted from the equations are merely estimates to be used in guiding the designer. They can be used as a check on the stack height and discharge conditions, and certainly can be used as a basis to justify either a taller stack or higher burning efficiency for the flare. The estimates are not however a guarantee that the design will perform to 100% of the specifications. Some judgement based on the experience of the designer and the plant's operating history should also be applied to the design, and ultimately, ambient air monitoring should be viewed as the only method of accurately determining the atmospheric effects of emissions.

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