MANUAL

PRESSURE RELIEF, EMERGENCY DEPRESSURING, FLARE AND VENT SYSTEMS

DEP 80.45.10.10-Gen.

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DESIGN AND ENGINEERING PRACTICE

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TABLE OF CONTENTS

1.	**INTRODUCTION**	5
1.2	DISTRIBUTION, INTENDED USE AND REGULATORY CONSIDERATIONS	6
1.3	DEFINITIONS	7
1.4	CROSS-REFERENCES	8
2.	**PRESSURE RELIEF**	9
2.1	GENERAL	9
2.2	RELIEF REQUIREMENTS; COMMENTS ON API RP 520/521	10
2.3	THERMAL EXPANSION RELIEF VALVES (TERVs)	12
2.4	THE USE OF INSTRUMENTED PROTECTIVE FUNCTIONS	13
2.5	RELIEF DEVICE SELECTION	15
2.6	RELIEF DEVICE DESIGN CONSIDERATIONS	19
2.7	RELIEF DEVICE SIZING METHODS	21
2.8	RELIEF VALVE LOCATION AND ARRANGEMENT	22
2.9	PREVENTION OF MALFUNCTIONING OF RELIEF VALVES	24
3.	**EMERGENCY DEPRESSURING SYSTEMS**	25
3.1	GENERAL	25
3.2	DETERMINATION OF DEPRESSURING REQUIREMENTS	25
3.3	SECTIONING OF PROCESS SYSTEMS	25
3.4	DEPRESSURING DESIGN CONSIDERATIONS	26
3.5	EVALUATION OF DEPRESSURING RATE	27
3.6	LOW TEMPERATURE EVALUATION	27
3.7	DEPRESSURING DEVICE LOCATION	27
4.	**FLARE AND VENT SYSTEMS**	28
4.1	GENERAL	28
4.2	DESIGN OF PIPING UPSTREAM OF A RELIEF DEVICE	29
4.3	SELECTION OF GAS DISPOSAL SYSTEMS	30
4.4	FLARE/VENT SYSTEM LOAD ANALYSIS	34
4.5	SIZING OF DOWNSTREAM PIPING SYSTEMS	35
4.6	LAYOUT OF DOWNSTREAM PIPING SYSTEMS	36
4.7	BLOCKAGE DUE TO HYDRATE FORMATION IN DOWNSTREAM PIPING SYSTEM	38
4.8	FLOW MEASUREMENT REQUIREMENTS	39
4.9	PIPING DESIGN	40
5.	**KNOCK-OUT DRUMS, WATER SEAL VESSELS AND LIQUID DISPOSAL FACILITIES**	41
5.1	DESIGN OF FLARE AND VENT KNOCK-OUT DRUMS	41
5.2	WATER SEAL VESSELS (SEE APPENDICES 3, 4, AND 5)	44
5.3	LIQUID DISPOSAL FACILITIES	45
6.	**STRUCTURES FOR FLARE AND VENT STACKS AND LIQUID BURNERS**	47
6.1	GENERAL	47
6.2	TYPE OF STRUCTURES	47
6.3	HEAT RADIATION LEVELS	48
6.4	DISPERSION LEVELS	49
6.5	NOISE LIMITS	49
7.	**FLARE AND VENT TIPS**	51
7.1	GENERAL	51
7.2	FLARE TIP DESIGN CONSIDERATIONS	51
8.	**FLARE AND VENT PURGING**	53
8.1	GENERAL	53
8.2	PURGING DESIGN CONSIDERATIONS	53
8.3	PURGE REDUCTION SEALS	54
8.4	FLAME ARRESTORS	54
9.	**VENT SNUFFING**	56
9.1	GENERAL	56
9.2	VENT SNUFFING REQUIREMENTS	56

10. **FLARE PILOTS AND IGNITION**....57
10.1 GENERAL....57
10.2 FLARE PILOT REQUIREMENTS....57
10.3 FLARE IGNITION REQUIREMENTS....58

11. **VENTING ATMOSPHERIC AND LOW PRESSURE STORAGE TANKS**....59

12. **REFERENCES**....60

APPENDICES

APPENDIX 1 VAPOUR RELIEF REQUIREMENTS FOR FIRE EXPOSURE....62

APPENDIX 2 TYPICAL LINE-UPS OF THERMAL EXPANSION RELIEF VALVES....64

APPENDIX 3 HYDROCARBON FLARE SYSTEM AND H2S FLARE SYSTEM....66

APPENDIX 4 WATER SEAL VESSEL DESIGN CHART....68

APPENDIX 5 TYPICAL DESIGN FEATURES OF WATER SEAL VESSEL....69

APPENDIX 6 ARRANGEMENT OF BLOCK VALVE FOR ISOLATING UNIT....70

APPENDIX 7 F-FACTORS USED IN THE API MODEL TO DETERMINE RADIATION 71

APPENDIX 8 ESTIMATE OF STEAM INJECTION REQUIREMENTS FOR FLARING....72

APPENDIX 9 PURGE RATES REQUIRED FOR PIPE FLARES....73

APPENDIX 10 FLARE KNOCK-OUT DRUM DESIGN CONSIDERATIONS....74

1. INTRODUCTION

This DEP specifies requirements and gives recommendations for the design and engineering of pressure relief, emergency depressuring, flare and vent systems.

This is a revision of the DEP of the same number dated August 1988.

For pressure relief facilities of liquefied petroleum gas in bulk storage installations, DEP 30.06.10.12-Gen. shall also apply.

Relieving facilities for vessels in marketing installations of capacities up to and including 135 m^3 shall be assessed in accordance with NFPA 58.

For the rating and adjustment of relief valves on power boilers refer to ASME I.

The relieving facilities for pressure vessels shall be in accordance with the following standards:

- ASME VIII, Division 1 or 2;
- API RP 520, Part I and Part II;
- API RP 521.

This DEP gives clarification of, amendments to and additional requirements to the above standards.

1.2 DISTRIBUTION, INTENDED USE AND REGULATORY CONSIDERATIONS

Unless otherwise authorised by SIOP and SIEP, the distribution of this DEP is confined to companies forming part of or managed by the Royal Dutch/Shell Group, and to Contractors nominated by them (i.e. the distribution code "C" as defined in DEP 00.00.05.05-Gen.).

This DEP is intended for use in oil refineries, gas plants, chemical plants, oil and gas production facilities and supply/marketing installations.

If national and/or local regulations exist in which some of the requirements may be more stringent than in this DEP the Contractor shall determine by careful scrutiny which of the requirements are the more stringent and which combination of requirements will be acceptable as regards safety, environmental, economic and legal aspects. In all cases the Contractor shall inform the Principal of any deviation from the requirements of this DEP which is considered to be necessary in order to comply with national and/or local regulations. The Principal may then negotiate with the Authorities concerned with the object of obtaining agreement to follow this DEP as closely as possible.

1.3 DEFINITIONS

1.3.1 General definitions

The **Contractor** is the party which carries out all or part of the design, engineering, procurement, construction, commissioning or management of a project, or operation or maintenance of a facility. The Principal may undertake all or part of the duties of the Contractor.

The **Manufacturer/Supplier** is the party which manufactures or supplies equipment and services to perform the duties specified by the Contractor.

The **Principal** is the party which initiates the project and ultimately pays for its design and construction. The Principal will generally specify the technical requirements. The Principal may also include an agent or consultant authorised to act for, and on behalf of, the Principal.

The word **shall** indicates a requirement.

The word **should** indicates a recommendation.

1.3.2 Specific definitions

Pressure and **temperature** terms	are defined in DEP 01.00.01.30-Gen.
LODMAT	Lowest One-Day Mean Ambient Temperature.

1.4 CROSS-REFERENCES

Where cross-references to other parts of this DEP are made, the referenced section number is shown in brackets. Other documents referenced in this DEP are listed in (12).

2. PRESSURE RELIEF

2.1 GENERAL

This section covers the selection, design and arrangement of pressure relieving devices.

2.2 RELIEF REQUIREMENTS; COMMENTS ON API RP 520/521

2.2.1 General

A pressure system can be protected by one or more relief valves, provided it is ensured that the system remains open under all conditions. It shall be established that blockage cannot occur due to valve closure, freezing, solidification, fouling, sublimation, damage of internals, etc., which could cause a section of the system to become isolated with no means of overpressure protection.

For each relief device the applicable causes of relief action and the resulting relief requirements shall be determined. The data for all pertinent relief conditions shall be recorded on a relief valve calculation sheet (DEP 31.36.90.94-Gen.) and in a safeguarding memorandum according to DEP 01.00.02.12-Gen. The possible causes are listed in API RP 520, Part 1, Table 1. A more detailed description of the causes of relief action is given in API RP 521, Chapters 2 and 3.

For the case of excess flow from a high pressure source (high pressure - low pressure interface) the largest contribution of either the fully open control valve or the fully open bypass valve shall be taken into account. The calculation shall be based on the worst case situation; typically this will be 100% vapour breakthrough (no liquid) and shall be made for the actually installed valves.

For the highly improbable event that gas breakthrough takes place when both control valve and bypass are fully open, the capacity of the relief valve shall be such that the downstream pressure shall not exceed the hydrotest pressure of the system, taking into account any differences between the hydrotest temperature and the maximum operating temperature.

In addition to the causes stated in API RP 520 and API RP 521 the emergency releases under start-up and shutdown conditions shall also be considered since they can be more severe than at design operating conditions.

In liquid-full systems, where no vapour buffer is present, relief valve chatter (i.e. rapid opening and closing of the valve) can be violent and very destructive due to its high frequency. It can easily result in damage to the relief valves themselves and to piping components. Therefore, the need for liquid relief in the event of blocked outlets should be obviated by matching the design pressure of liquid-full systems to the pressure of the blocked outlet case.

2.2.2 Fire case requirements

Amended per Circular 32/98

Equipment shall be protected against high pressure due to fire under the following conditions:

a) If the equipment is located in an area where a sustained intense fire might occur, and

b) If it is conceivable that the equipment is blocked in without having been emptied first when such a fire occurs.

An exception to both a and b above may be made for small-capacity (< 500 litres) equipment containing liquid which is non-toxic and has an initial boiling point higher than the maximum ambient temperature.

During a fire, all input and output streams to and from the fire-affected equipment and all internal heat sources within the process are assumed to have ceased after fire detection and operator intervention. This may require ESD valves to be located appropriately.

Potential fire areas shall be identified and clearly shown on a plot plan. The fire areas shall be numbered and each number shall be indicated on the relief valve data sheet. For process units a typical fire area of 300 m^2 should be assumed, depending on the drainage design of the plot.

The height of the flame to be considered shall be 8 metres from grade or a platform on which liquid can accumulate (concrete platform). Refer to DEP 30.06.10.12-Gen. for fires in

LPG bulk storage installations.

NOTE: For buildings or offshore facilities the concept of fire areas is replaced by one of fire compartments segregated by fire walls or floors. The area of these compartments is usually determined by fire containment requirements.

If the equipment is already designed to be protected against operational emergencies by a relief valve, the same valve may serve for the fire case, provided it is of adequate size for all emergency cases.

Vapour relief requirements in the event of fire should be based on the heat absorbed by the internally wetted surface area of the vessel, considering the effects of liquid vaporisation and expansion, and vapour density change. In addition to the vessel the surface area of associated piping and equipment should be included. For the purposes of these calculations, the liquid level in the process vessel should be based on the normal liquid volume including any liquid draining down from the internals of the vessel.

Although fire water deluge provides additional protection, no allowance shall be made for active fire protection facilities when calculating the relief rate. However, with the approval of the Principal, allowance may be made for fire protection systems (e.g. fire proof insulation) as stated in API RP 521, Table 5 if it is ensured that the insulation is resistant to fire heat and to the impact of fire water jets and jet fires. For further explanation see also Appendix 1 of this DEP.

Piping within process or utility boundaries does not normally require relief valves sized for fire relief but relief may be required for thermal expansion in liquid-full systems (refer to section 2.3). Where possible the design of a facility should avoid blocked-in sections of pipework.

2.2.3 Requirements for plate and printed circuit heat exchangers

Essentially the considerations given in API RP 521 with respect to shell and tube heat exchanger failure also apply to plate and printed circuit heat exchangers. Although both sides of the actual exchanger will have the same design pressure this may not be the case for other equipment on the low pressure side which may be subjected to overpressure. This shall be taken into account when sizing the relief device.

For a plate and printed circuit heat exchanger the maximum relief rate shall be based on a complete rupture running longitudinal to a corrugation. It is then necessary to consider the particular arrangement of the exchanger in question to determine the maximum number of inlets that are connected to the corrugation. The total cross-sectional area shall be calculated allowing for twice the cross-sectional area of each inlet connected to the corrugation.

2.3 THERMAL EXPANSION RELIEF VALVES (TERVs)

Thermal expansion relief valves are required in liquid-full systems if the system can be blocked in and subjected to heat input from the atmosphere or process. The theoretical pressure rise for most liquid systems lies in the range 5 to 15 bar for each degree centigrade of temperature increase. In practice, the theoretical pressure rise is not attained because systems are rarely totally liquid full and usually have small leakage's through valve seats or gaskets. Calculations of the pressure rise are thus of little use in formulating realistic guidelines for the application of thermal relief valves.

The following factors should be considered before deciding not to fit TERVs:

- Is the linework or equipment continuously in operation and thus unlikely to be isolated without being depressurised and drained? Most process equipment and piping is in this category.
- Is the liquid non-toxic and non-corrosive and does it have an initial boiling point higher than the maximum ambient temperature?
- Is there a weak point in the system such as a flange where expansion could relieve pressure without unacceptable consequences?
- Is it unlikely that the system will be totally liquid-full (i.e. to more than 95%)?
- Has the release a safety or environmental impact?

As a general guide, TERVs are not needed for:

- process plant piping;
- storage or transport piping sections which are not normally shut in for operational or emergency purposes;
- lines in which there is normally two-phase flow;
- systems which are not totally liquid filled, i.e. less than 95% liquid.

TERVs are normally fitted to:

- the cold side of heat exchangers which can be blocked in;
- sections of piping containing more than 500 litres of LPG or toxic liquids which could be blocked in;
- piping in storage areas or transport lines which will be regularly blocked in during normal operation, to protect against pressure rise due to solar heating or heat tracing;
- all sections of cryogenic piping (operating below ambient temperature) which can be blocked in irrespective of whether this is routinely done or not.

In all cases, operating procedures should be such that overpressure due to thermal expansion cannot occur, even when a relief valve is fitted. This can be done by prescribing drainage of a small quantity after a pipe section has been blocked in. In this case, if allowed by the code, e.g. ASME B31.3, the set pressure of the TERVs may be 110% of the maximum allowable working pressure. This at least guarantees that the maximum allowable working pressure is not exceeded by more than 33% (refer ASME B31.3) and that during normal operation the TERV will not operate and cause operational mishaps such as liquid spillage, permanent leakage, valve chatter, etc.

The standard TERV size for piping systems is 25 x 25 mm, flanged, with a minimum orifice area of 0.71 cm^2. Bronze valves, which are allowed for water service only, may have screwed connections.

2.4 THE USE OF INSTRUMENTED PROTECTIVE FUNCTIONS

The conventional approach to overpressure protection of piping and pressure vessels is to utilise two totally independent forms of protection. The first is a pressure control system. The second utilises relief valves sized to relieve the pressure to a safe location if the first system fails and is a complementary requirement dictated by the code and local authority regulations. This implies that normally relying solely on control is not permitted.

In specific circumstances, it may be highly desirable to limit or even to eliminate emergency relief, since the flare relief system will become disproportionately large and expensive, or the environmental impact of such a relief would be unacceptable. In these cases the level of protection afforded by a relief valve may be complemented with an Instrumented Protective Function (see DEP 32.80.10.10-Gen.).

Complete substitution of relief valves by an Instrumented Protective Function can rarely be justified unless RVs are impossible or impractical. It should be recognised that local regulatory authorities may either prohibit Instrumented Protective Functions as a substitute for protection by RVs or demand to approve the design before issuing a certificate of fitness.

Gas transmission pipelines are considered to be special cases and are covered in DEP 31.40.10.14-Gen.

Typical examples of the application of Instrumented Protective Functions are:

a) Equipment connected to a multiple well system

The design pressure of the equipment will normally be lower than the shut-in pressure of the wells, therefore overpressure protection to guard against shut-in conditions is required. However if the relief system has to cope with the full capacity of all production wells the relief system, especially when large quantities of liquid have to be released, becomes disproportionately large. Fast-acting valves triggered by high pressure switches may be implemented. A reliability study shall be conducted to determine on how many wells the relief valve sizing is to be based. This will also ensure that there is no common mode failure mechanism capable of allowing all wells to stay open.

b) Reduction of flare size for common failures

In oil and gas processing facilities a cause of overpressure could be a common failure affecting a number of pressure systems simultaneously. An example of this is the simultaneous loss of overhead condensing capacity on a number of columns due to loss of cooling medium (either loss of cooling water supply or loss of all air cooler fans due to power failure). To reduce the total load on the common relief system, it may be considered to use Instrumented Protective Functions (IPFs) consisting of high pressure initiators on each pressure system to close the heat input and thus prevent the individual relief case. However, it must be assumed that an IPF can fail and the greatest relief load that would result from an IPF failure shall be used for the design of the common relief system. In the unusual event that there are more than five IPFs it must be assumed that two IPFs can fail and the two greatest relief loads shall be used for the design of the common relief system.

In any case each individual pressure system shall still be equipped with sufficient relief capacity for its individual relief case.

c) The use of air-assisted relief valves

With spring-loaded relief valves the recommended margin between Maximum Operating Pressure and Set Pressure (refer DEP 01.00.01.30-Gen.) is 10%. By adding an air-operated actuator on top of the relief valves (implementing an air-assisted relief valve) this margin can be reduced to 5%, reducing the required design pressure of the unit with consequential substantial cost savings (refer 2.5.2.iv). If the instrumented system fails, the relief valve will not open at set pressure. However the supplemental load of the air-operated actuator is such that the relief valve will open at a pressure not higher than 110% of set pressure, which is guaranteed by an additional small relief valve of sufficient capacity. The reliability of the instrumented system is guaranteed by applying a 1 out of 2 voting system.

The above are typical examples. Detailed engineering shall be carried out to guarantee that they are fully correct. Reference should also be made to DEP 32.80.10.10-Gen. If such systems are applied they shall be submitted to the Principal for approval.

2.5 RELIEF DEVICE SELECTION

2.5.1 General criteria

Amended per Circular 32/98

Relief valves shall have established data and shall have been confirmed by the Manufacturer as suitable for the service conditions.

For standardization, the following range of standard sizes in accordance with API 526 shall be used for valves not exceeding 8T10 size:

Inlet/Orifice/Outlet	Orifice Area	
	[sq. in]	[sq. cm]
1D2	0.110	0.71
1E2	0.196	1.26
1-$^1/_2$F2	0.307	1.98
2H3	0.785	5.06
3K4	1.838	11.86
4L6	2.853	18.41
4P6	6.380	41.16
6Q8	11.050	71.23
6R10	16.000	103.23
8T10	26.000	167.75

The risk of unstable valve operation, e.g. in liquid filled systems where oversized valves can cause chattering of the valve, makes it sometimes necessary to make use of sizes not included in the above standard range.

For required orifices larger than 8T10, pilot-operated valves may be used; the process medium should then be dry and clean.

The minimum rating for flanges shall be ASME Class 300 for inlet flanges and ASME Class 150 for outlet flanges, unless the service requires a higher rating. The flanges shall form an integral part of the body.

The set pressure limit according to API 526 shall be not less than 20.7 bar (ga) (ASME class 300) at 232 °C, except for the sizes R and T for which the set pressure limit shall be not less than 15.8 bar (ga) at 232 °C.

Flange face finish shall be in accordance with the piping class of the connected piping, refer to DEP 31.38.01.11-Gen.

Where a number of relief valves are provided in parallel the set pressures may be staggered in accordance with (2.6) to avoid valve chatter.

Relief valves shall be adequately supported to withstand the high reaction forces exerted when they are relieving.

Relief valves for steam service discharging to atmosphere shall be of the exposed-spring type if operating above 250 °C.

Manual lifting devices may be provided only for steam, hot water and air service.

Test rods shall not be provided (they prevent the stem moving and are sometimes used to block the relief valve during hydrotesting of the equipment or relief valve transport).

Where a number of relief valves are provided in parallel the set pressures may be staggered in accordance with (2.6) to avoid valve chatter.

Adequate support of relief valves shall be provided, since they are subjected to high reaction forces when they are relieving.

Relief valves for steam service discharging to atmosphere shall be of the exposed-spring type if operating above 250 °C.

Manual lifting devices may only be provided for steam, hot water and air service.

Test rods shall not be provided (they prevent the stem moving and are sometimes used to block the relief valve during hydrotesting of the equipment or relief valve transport).

2.5.2 Specific selection criteria

Relief valves shall normally be of the spring-loaded type. For special applications pilot-operated relief valves, air-assisted relief valves or bursting disks may be applied.

For a description of the various types of relief devices, reference is made to API RP 520 Part I.

The criteria for the selection of a relief device are as follows:

i) Non-balanced spring-loaded relief valves

These shall be used when the built-up back pressure is not greater than the maximum allowed overpressure.
The built-up back pressure (variable back pressure) is the pressure drop across the piping downstream of the relief valve.
The maximum allowed overpressure is the maximum allowable accumulated pressure minus 1) the set pressure and 2) the pressure drop across the piping between the relief valve and the protected equipment. Determination of the maximum allowable accumulated pressure is given in (2.6).
For cases where a constant back pressure is exerted, the spring of the relief valve should be selected and set for the difference between the set pressure (being the maximum allowable working pressure) and the constant back pressure.
Spring-loaded relief valves connected to a closed discharge system shall always have a closed bonnet.

NOTE: Local regulations may not accept non-balanced relief valves set at equipment design pressure if the actual opening pressure could at times be higher as a result of the back pressure generated by other relief valves blowing at the same time. In that case, the use of balanced valves, or a slightly lower setting of the non-balanced valves, should be considered for those valves which may blow as the result of a common emergency.

ii) Balanced type relief valves

These shall be used when the built-up back pressure is higher than that allowed for a non-balanced valve. However, the back pressure should not exceed 50% of the set pressure, taking into account the maximum allowable operating pressure for the bellows.

Relief valves of the balanced type shall have a vented bonnet. See (6.4) regarding the need to route to a safe area.

A suitably corrosion-resistant material shall be specified for the bellows since, in the event of a bellows failure, fluid could enter the bonnet space and escape through the bonnet vent. If no suitable corrosion-resistant material can be selected or the chance of leakage is present, the use of an auxiliary balancing piston should be considered.

Balanced-bellows valves, with a bonnet vent open to atmosphere, shall not be used on streams which cause the valve to freeze up when it leaks. If this happens, the bonnet and balancing bellow will fill up with ice due to condensation and subsequent freezing of the moisture in the air, making the valve inoperative. Instead of balanced-bellows relief valves, the use of an auxiliairy balancing piston should be considered. In clean service, pilot-operated relief valves, of which the pilot stays relatively warm, may be selected. See also paragraph 4.2.

iii) Pilot-operated relief valves

Pilot-operated relief valves may be used in a clean service where proper operation of the pilot valve is always guaranteed, if permission has been obtained from local authorities. The pilot valve shall be of the non-flowing type. With this type, the flow through the pilot lines and pilot valve ceases immediately when the main valve opens, thus minimising the likelihood of foreign matter being drawn into the small bore sections of the valve.

Situations in which the selection of pilot-operated relief valves may be viable are:

- where the pressure loss between the protected equipment and the inlet flange of the relief valve exceeds 3% of the set pressure of the relief valve (refer 2.6), since the pilot sensor tapping could be located close to the source of overpressure;
- where leakage is a problem, since they have superior sealing qualities due to their soft seats;
- where high back-pressures are possible;
- in large capacity and high set pressure service due to their smaller size and weight;
- if the margin between the maximum operating pressure (MOP) of the protected system and the relief valve set pressure is less than 10% of the relief valve set pressure. This is due to the high positive closing force which increases with process pressure up to the relief valve set pressure.

It shall be ensured before selecting a pilot-operated relief valve that there is no possibility of blockage of the pilot valve or sensing line due to hydrates, ice, wax or solids. There shall be no low points in the sensing line or its take off, and all fine bore elements exposed to process fluids shall be heat-traced and insulated if non-blockage cannot be guaranteed. Filters shall not be used in the sensing line to the pilot valve because they can increase the risk of blockage.

A field device shall be provided for checking the set pressure of the valve.

iv) Air-assisted relief valves

Air-assisted relief valves are non-balanced spring-loaded relief valves equipped with a pneumatic actuator. The function of the actuator is to keep the relief valve closed when operating close to set pressure. In this way the operating margin between set pressure and maximum operating pressure (MOP) may be smaller, and instead of 10% a margin of 5% may be used for design. To obtain this goal the following options are possible:

1) The actuator assists to open the relief valve exactly at set pressure. In this case the spring setting of the relief valve is 105% of design pressure. This means that the relief valve will open at 105% of design pressure regardless of the instrumentation.
2) The actuator assists to keep the relief valve closed up to the latter's set point (supplemental load). The spring setting of the relief valve is 100% of design pressure. Opening of the relief valve also relies on the operation of the instrumentation. However in the event of an instrumentation malfunction the relief valve shall in any case open at 110% of set pressure.
3) The action of the actuator is a combination of 1) and 2). But since a supplemental load is applied, the spring setting of the relief valve may be 100% of design pressure.

The correct operation of the relief valve depends on the reliability of the instrumentation, but the above options cover the eventuality of instrumentation malfunction.

It should be noted that air-assisted relief valves are not described in most pressure vessel codes and their use may require the approval of local regulatory authorities. The same approach as outlined for instrumented pressure protection systems (see section 2.4) shall also be followed for air-assisted relief valves.

v) Bursting disks

As a general rule bursting disks should not normally be installed since they cannot be tested. They have to be replaced after rupture, and once they have ruptured flow continues until all the pressure is relieved.

Bursting disks, preferably of the reverse buckling type, may be used:

- to accomplish a fast response time, which cannot be achieved with a relief valve. This could be required to cope with a sudden gas breakthrough due to a heat exchanger tube burst or malfunctioning of a level control valve into a liquid filled system;
- to prevent RVs in vacuum service from drawing gas or air back into the process;
- to protect relief valves from being in continuous contact with a corrosive or solidifying process fluid;
- to prevent leakage of toxic substances through the relief valve.

If a bursting disk is installed upstream of a relief valve, small pin-holes in the disk could create an equal pressure upstream and downstream of the disk, which will prevent the disk from collapsing at bursting pressure (relief valve set pressure). Therefore, in pressure service the compartment between the disk and the relief valve should have an open outlet through a restriction orifice to either the atmosphere (safe location) or into the RV discharge piping. Additionally, to safeguard the integrity of the system, a pressure gauge (in vacuum service) or a pressure alarm (in pressure service) shall be provided between the disk and the relief valve. This will give an indication when the bursting disk is ruptured.

Two bursting disks shall not be installed in series to provide protection against fatigue and creep. It is important to select the correct disk for the particular application. The desirability of replacing the bursting disk after a specified period of operation should also be considered.

In order to avoid premature failure due to the effects of creep and fatigue, a substantial margin (at least 30% of the nominal bursting pressure for tension-loaded disks, and at least 10% of the minimum bursting pressure for reverse-buckling disks) shall be allowed between the maximum operating pressure of the vessel and the design bursting pressure of the disk. It should be noted that the bursting pressure of bursting disks is substantially influenced by temperature; care should be taken when specifying a bursting disk to ensure that it protects the equipment under all possible working conditions.

If a relief valve is installed downstream of a bursting disk, a flow correction factor (derating factor) of 0.9 shall be used in sizing the relief valve unless a flow test of the combination is carried out which shows that a more closely defined derating factor shall be used (Refer ASME Section VIII, Division 1, UG-127).

2.6 RELIEF DEVICE DESIGN CONSIDERATIONS

The size of a relief device shall be determined for the most severe individual relief condition.

Set pressures (SP) and maximum relief pressures, expressed in relation to the design pressure (DP) of the protected equipment, all expressed in gauge pressures, shall not exceed the values given in the table below:

NOTE: See DEP 01.00.01.30-Gen. In the table below, MAWP may be substituted for DP.

	Set pressure (SP)		Maximum allowable accumulated pressure ***	
	Non-fire conditions	Fire conditions	Non-fire conditions	Fire conditions
Single valve	100% of DP	100% of DP	110% of DP	121% of DP
Multiple valves	At least one valve 100% of DP maximum settings at 105% of DP**	110% of DP*	116% of DP	121% of DP

The above is strictly related to ASME VIII, Division 1. If equipment is built in accordance with another code, that code should be applied. BS 5500 for instance does not allow an accumulation of more than 10% of the design pressure in any situation, including fire.

* Relief valves for fire protection may only be set at 110% of DP if they are installed in addition to adequate relief protection of the process equipment against non-fire situations.

** For set pressures below 10 bar, staggering of set pressure becomes impracticable because the difference between the set pressure tolerance of 3% (according to ASME VIII, Division 1, UG 134) and the value of 5% of the DP becomes too small.

*** Maximum allowable accumulated pressure is the sum of the design pressure and the maximum allowable accumulation.
The overpressure used for the calculation of the relief valve can be derived from the maximum allowable accumulated pressure. This overpressure is the maximum allowable accumulated pressure minus 1) the set pressure and 2) the pressure drop of the piping connecting the relief valve and the vessel. The pressure drop shall not exceed 3% of the set pressure. This shall be based on the relief valve capacity.

NOTE: Selecting higher design pressures for equipment than prescribed by operating conditions according to DEP 01.00.01.30-Gen. will result in lower volumetric relief rates and consequently smaller relief valves and discharge piping. Therefore, the use of higher design pressures should be considered if it results in an overall cost saving, (e.g. if the wall thickness of the pressure vessel is not determined by internal pressure, but by external loads, wind, transport, handling, or design for full vacuum).

It shall also be ensured that the blowdown pressure (reseating pressure) is above the maximum operating pressure. However, if the blowdown pressure is set too close to the set pressure the relief valve may open and close rapidly causing damage to the valve. For most services the blowdown pressure will usually be 5-7% below the valve set pressure.

For equipment protected by a pressure relieving device, either on the equipment itself or elsewhere within the pressure system containing the equipment, there shall be an adequate margin between the set pressure and the Maximum Operating Pressure (MOP). DEP 01.00.01.30-Gen. more specifically defines which margins should be maintained.

Valve materials shall be suitable for the inlet and outlet temperatures that result from extreme operating and emergency conditions. The possibility that the valve may leak under otherwise normal operating conditions shall be considered. For the choice of valve spring materials, only the extremes of non-fire conditions shall be considered.

Material selection should be in accordance with API Std 526. The springs of relief valves shall be protected with a suitable coating preventing the occurrence of general corrosion and/or sulphide stress corrosion cracking. Coatings of cadmium or zinc are not allowed due to the risk of liquid metal embrittlement during service and/or hydrogen embrittlement during the galvanising process. Suitable aluminium coatings are permitted. For services not indicated in API Std 526 the material shall be selected in consultation with the Principal.

2.7 RELIEF DEVICE SIZING METHODS

2.7.1 Sizing of relief valves

For each relief valve, the relevant relief conditions will be established (see Section 2.2.1.). The size of the relief valve and the inlet and outlet piping shall be determined using the largest relief load.

For reporting the calculations of the relief valve size, the standard calculation sheet (DEP 31.36.90.94-Gen.) shall be used. The formulae in API RP 520 shall be applied.

For liquid relief valves the sizing method for certified relief valves (refer API RP 520) should be used. This ensures a more accurate size determination, avoiding oversizing and thereby minimising the probability of chatter.

For two-phase flow relieving conditions (which also exists if liquid is released above bubble point pressure) the method specified in API RP 520 should not be used. This method determines the required orifice area assuming the fluid passes through the orifice as two separate fluids (gas and liquid). Based on the standard equations for gas and liquid the individual orifices are calculated, which are added together to give the required orifice area. However, since the two-phase fluid chokes at a much lower velocity than the pure gas phase whereas the liquid's velocity is not restricted (does not choke), the API RP 520 method results in too small a calculated orifice area. This has been recognised by DIERS (Design Institute for Emergency Relief Systems). They have proposed a physically more sound solution, based on the classical homogeneous equilibrium model assuming no slip between the liquid phase and the vapour phase.

Thermal equilibrium between the two phases is also assumed. This assumption is considered valid for cases where mass, momentum and heat transfer are exchanged very rapidly between the phases, as is the case when relief valves operate.

The flashing or two-phase mass flux through a relief valve can be determined from the following expression.

$$G = \frac{W}{A} = \frac{[2(h_0 - h)]^{1/2}}{\nu} = \frac{[2\int_{Po}^{P} -\nu.dP]^{1/2}}{\nu}$$

where:

G = mass flux (kg/m^2 s)

W = discharge rate (kg/s)

A = relief area (m^2)

h = fluid specific enthalpy (J/kg) at P

ν = fluid specific volume (m^3/kg) at P

P = (decreasing) downstream pressure (N/m^2)

subscript $_o$ indicates upstream condition

The integral in the above expression is evaluated by determining the fluid specific volume, ν, as the downstream pressure, P, is decreased. This can be achieved by successive flashes of the fluid starting at the relief valve upstream pressure, Po, and continuing to successively lower pressures. The critical mass flux is when the value of G becomes maximum and the choking pressure for the fluid occurs at this point. The thermodynamic properties can be obtained from PROII which makes use of Shell's proprietary thermodynamic basic data and calculation methods. For this reason a calculation module to be used in PROII has been prepared to carry out the above described calculation routine.

To determine the minimum required area the calculated area shall be divided by a Kd (effective coefficient of discharge) of 0.90.

2.8 RELIEF VALVE LOCATION AND ARRANGEMENT

2.8.1 Relief valve location

To ensure protection of the whole system, the relief assembly should be located, where practical, in the upstream part, i.e. where the highest pressure occurs, and as close as possible to the source of overpressure.

Relief valves shall be connected to the protected equipment in the vapour space above any contained liquid or to piping connected to the vapour space. This means that the relief valve is preferably connected to the highest point of the vessel. An exception can be made if the vessel is fitted with a demister mat. In which case the relief connection shall be upstream of the mat, unless the relieving capacity is of the same order of magnitude as the normal operating flow through the demister mat.

The pressure drop of the piping between protected equipment and its relief valve shall not exceed 3% of the set pressure.

Spring-loaded, pilot-operated or air-assisted relief valves, and thermal expansion relief valves (TERVs) shall always be installed in the upright position. The inlet and outlet piping shall be installed without pockets to ensure that liquid does not accumulate at the relief valve outlet or inlet. An exception to the latter may be made for TERVs, since a position close to the protected equipment is preferred. However, it should be ensured that the discharge pipe will not be plugged by freezing.

Relief valves discharging to atmosphere should be located at the maximum practical elevation to keep discharge piping (to safe location) as short as possible. To keep this discharge pipe free from liquid a small (8 mm dia) weephole shall be drilled at the lowest point. In the case of multiple relief valves (including one spare), each relief valve shall have an individual discharge pipe.

Relief valves connected to a closed relief system shall be located above the relief header. Relief valve outlet lines should be connected to the top of the header, or at least so that the header cannot drain back into outlet lines even with the header full of liquid.

If the valves cannot be put above the header, they shall be lined up to discharge into a local drain vessel. This line shall be locked open and adequately sized (50 mm for headers up to 200 mm, 100 mm for headers up to 400 mm and 150 mm for headers larger than 400 mm). The local drain vessels shall be equipped with a high level alarm, set as low as possible to provide a maximum hold up.

Alternatively, if the problem of elevation is confined to a few valves, and if the Principal agrees, outlet lines to the header shall be heat-traced from the relief valve to the highest point of the line. Such an arrangement is not permitted for relief valves which discharge a medium which can leave a deposit. The heat-tracing may be omitted if the relief valve and connecting header only handle products which vaporise completely at the lowest ambient temperature.

Relief valve systems require periodic inspection and maintenance and hence they should be easily accessible.

For liquids with a high pour point, insulation and heat tracing of piping upstream and downstream of the relief valves (or bursting disks) should be applied. Upstream, and sometimes also downstream, of the relief valve may be flushed with a low viscosity hot process fluid. If the downstream lines are flushed, a local knock-out drum shall be provided. This knock-out drum will prevent introduction of any liquid into the main flare system. The flushing rate shall be such that only a small layer of liquid will be present in the relief header.

2.8.2 Relief valve arrangement

Different relief valve arrangements may be used depending on the on-stream factor required and the testing interval for the relief valve if different from the inspection interval of the protected equipment.

Where possible, the approach should be to use a relief valve arrangement which does not utilise any isolation valves. This approach eliminates the possibility of a relief valve being isolated in error. However, in this case inspection and maintenance of the relief valve, vent or flare system, and any other equipment connected to it, requires the complete shutdown of the whole system. If a complete shutdown is not practicable, then the use of separate flare or vent systems for each part of the plant which can be shut down independently could be considered. However this could lead to a costly design.

In the light of the above, it may be necessary to use isolation valves either to isolate the individual relief valve or to isolate a complete plant section. If isolation valves are used to isolate relief valves, there is a basic difference between the need for an inlet valve or for an outlet valve. An inlet valve is needed if the process cannot be shut down, whereas an outlet valve is needed if the relief header cannot be taken out of service. Thus a single relief valve (without a spare) connected to a relief header which cannot be shut down will have only an outlet isolation valve. A multiple relief valve arrangement (including a spare) will have an inlet isolation valve and outlet isolation valve. The spare relief valve could be substituted by an open spool piece (in fact a dummy RV with the same flange geometry)

It is mandatory that a pressure system is never without relief valve protection, unless fully emptied and depressured. Therefore a key operated locking system shall be used to ensure the correct mode (open or shut) of these isolation valves. The locking system on block valves used on the inlet and outlet of relief valves, their spares and their open spool pieces shall comply with DEP 80.46.30.11-Gen.

To indicate the proper operation of the upstream block valve, a vent connection shall be provided between the upstream block valve and the relief valve.

For pipeline TERVs (which may be removed while the system is in operation) a single relief valve with upstream and downstream isolation valves should be provided. Strict procedural controls should ensure that the line is not shut-in while the relief valve is out of service. The block valves may be sealed open rather than locked open, in a manner to be agreed by the Principal.

2.9 PREVENTION OF MALFUNCTIONING OF RELIEF VALVES

2.9.1 Relief valves affected by hydrates and freezing

The Joule-Thomson effect, occurring across the relief valve when relieving, may lower the temperature to within the hydrate or ice formation region. Due to the high velocities, there will be no problem of relief valve blockage at relieving conditions. For correct piping, see (4.7).

Valve blockages could occur due to small leaks across the relief valve seat. To prevent this blockage, heat tracing shall be provided around the relief valve. The integrity of the heat tracing shall be secured by proper instrumentation. In addition, the use of inhibitor injection may be considered.

2.9.2 Relief valves affected by corrosive and/or toxic process fluids

Corrosive fluid, e.g. sulfolane, HF, HCL, may attack the internals of the relief valve. Proper material selection is required to guarantee the suitability of the relief valve. However, the corrosivity of the fluid could be such that a proper material is virtually unobtainable. Furthermore, a relief valve cannot be assumed to be fully leak-tight and corrosive and/or toxic fluid may then enter the flare relief system. To prevent the fluid from corroding and/or passing through the relief valve, one of the following preventive actions shall be taken:

- For corrosive duty: purging or flushing of the inlet piping with a clean fluid which can be accommodated by the process.
- For corrosive and/or toxic duty: installation of a bursting disk upstream of the relief valve. For installation and more details on bursting disks see (2.5.2 v).

2.9.3 Relief valves affected by solidifying process fluids

Although relief valves are installed at the highest point of the vessel, which normally contains vapour, it shall be assumed that liquid may reach the relief valve. This could be by condensation or by liquid entrainment during emergency relief action. This liquid could solidify (e.g. high pour point liquid) and affect the operation of the relief valve and downstream piping. In this case one of the following preventive actions shall be carried out:

- The provision of heat tracing at the relief valve and along its inlet and outlet piping.
- Purging or flushing of the inlet and outlet piping with a clean fluid which can be accommodated by the receiving system.

3. EMERGENCY DEPRESSURING SYSTEMS

3.1 GENERAL

Unless special provisions are made, a relief valve cannot depressurise a vessel or system; it can only limit the pressure rise to the set point during upset or emergency conditions. Since a fire can heat equipment or piping walls to temperatures higher than design, the pressure must be reduced in order to lower the stress. The pressure shall be reduced by using a high rate emergency vapour depressuring system actuated by operator action. Remote actuation shall be possible from a safe distance, usually the control room.

This section covers the selection, design and arrangement of these emergency systems. It is not concerned with the depressuring of equipment and piping for shutdown and maintenance purposes.

3.2 DETERMINATION OF DEPRESSURING REQUIREMENTS

Emergency depressuring systems are principally required to reduce the risk of loss of equipment integrity during a fire or to reduce a local loss of containment arising from a leak when such an occurrence could create an unacceptable safety hazard. Depressuring during a fire must minimise the risk of the pressure within the system exceeding the rupture pressure preventing the occurrence of a BLEVE (Boiling Liquid Expanding Vapour Explosion). Consideration shall be given to the reduction of yield strength with increasing temperature.

In assessing whether or not depressuring facilities are required, particular attention should be paid to equipment location with respect to other equipment, buildings and personnel, and the contents of the equipment in terms of quantity and composition. All process equipment containing at least 4 m^3 of butane or an even more volatile liquid under normal operating conditions shall be provided with remotely operated vapour depressuring valves.

In addition, the following shall be provided with depressuring facilities:

- high pressure sections of hydroprocessing units, e.g. platformers, hydrotreaters, desulphurisers, hydrocrackers, and residue hydroconversion units;
- all manned offshore facilities where high pressure process facilities are present.

For LPG bulk storage, depressuring facilities need not be applied, since liquid pool fires are highly unlikely and such storage facilities shall be fitted with a firewater sprinkler system, see DEP 30.06.10.12-Gen.

3.3 SECTIONING OF PROCESS SYSTEMS

In order to reduce the design emergency depressuring flow rate, process sectionalisation may be considered. Process sectionalisation is a philosophy applied to split an installation into a number of smaller fire zones (see 2.2.2). Each zone shall be isolated by emergency shutdown (ESD) valves. Each zone shall be provided with its own depressuring facilities, such that each zone can be depressurised sequentially, thereby reducing the design peak depressuring rate.

It should, however, be ensured that common mode failure (i.e. loss of instrument air or electrical power) cannot cause all the depressuring valves to open simultaneously, otherwise the flare system shall be sized accordingly. Separate air and power supply systems are usually warranted for each section.

After initiating depressurisation of the first zone, that of other zones may be initiated as soon as the depressuring rate of the first zone has decayed such that a second zone may be initiated without exceeding the design capacity of the flare/vent system. This strategy requires equipment in an adjacent fire zone to be adequately protected by a combination of appropriate layout, fire walls, fire proof insulation, drainage and deluge, such that the risk of a loss of integrity of equipment in a fire zone adjacent to the first affected zone is insignificant. This approach is not usually practical on an integrated offshore production platform.

Within a fire zone all depressuring valves shall open simultaneously; sequenced depressuring using time delays shall not be used.

3.4 DEPRESSURING DESIGN CONSIDERATIONS

3.4.1 Valve requirements

Depressuring valves shall be fitted in addition to a system's required relief valves; relief valves shall not be used for depressuring. The depressuring valves shall normally be actuated by an operator from a remote location. However, in unattended operation they shall be actuated automatically by a signal from the emergency shutdown system, initiated by fire or gas detection. In the case of a gas leak this is both a precautionary measure and a means of reducing the size and duration of the release. For facilities with a control room, operation of any emergency depressuring valve shall be possible from the control room.

For the instrumentation and configuration of depressuring systems reference shall be made to DEP 32.45.10.10-Gen.

3.4.2 Valve sizing considerations

In performing the depressuring analysis it shall be ensured that throughout depressuring the system pressure never exceeds the load bearing capacity of the equipment. Account shall therefore be taken of the reduction of strength with increasing temperature.

The depressuring system shall reduce the pressure of the equipment within a fire zone to 50% of the design pressure within 15 minutes. This does not imply that the depressuring stops after 15 minutes.

The calculation shall take into account the following:

i) vaporisation of the liquid due to the reduction in pressure;

ii) the change in density of the vapour in the equipment due to the pressure reduction and temperature increase;

iii) vaporisation due to heat input from the external fire.

The depressuring of systems containing rotating equipment may also require special consideration. Depressuring may be required in much less than 15 minutes due to the loss of seal oil pressure. The designer shall ensure that rotating equipment manufacturers state their maximum time for depressuring. Larger seal oil tanks may be necessary to achieve this.

In the case of equipment which warms up slowly and operates at less than 50% of the DP, the depressuring valve shall be sized so that the pressure in the equipment will not reach 50% of the DP within 30 minutes of fire exposure.

Sizing of depressuring valves shall be based on the assumption that during a fire all input and output streams to and from the system are stopped and all internal heat sources within the process have ceased. It shall also be assumed when calculating the vapour load generated that fire is in progress throughout the depressuring period. The moderating effect of fire resistant material may be taken into account as in section (2.2.2). See Appendix 1.

For designing the depressuring system capacity it shall be assumed that the initial pressure in the equipment under consideration is the relief valve set pressure or the pressure which will be reached after 15 minutes of fire exposure, whichever is the lower.

To determine the vapour depressuring flowrates it is necessary to establish a liquid inventory and the vapour volume of the system. This shall include all facilities located in the fire area and all equipment outside the fire area which, under normal operating conditions, are in open connection with the facilities located within the fire area.

The effective vapour-generating surface area of a piece of equipment exposed to fire is the area wetted by the internal liquid contents of the equipment.

Subject to corrections made necessary by the final plant design, the following assumptions may be made:

- the liquid inventory of fractionating columns can be estimated as the normal column bottom and draw-off tray capacity plus the normal tray liquid hold-up (based on the pressure drop over the column);

- the liquid inventory of accumulators, flash drums, separators, knock-out drums etc. may be based on normal operating levels;
- for shell and tube heat exchangers, it may be assumed that one third of the total shell volume is occupied by the tube bundle.

The presence of active fire protection systems (e.g. deluge water) to reduce heat input to equipment from a fire is standard practice but their operation cannot be guaranteed and hence they cannot be considered in depressuring calculations.

3.5 EVALUATION OF DEPRESSURING RATE

The method used to evaluate the depressuring rate shall be as given in API RP 521. The effort required may be greatly reduced by the use of a suitable computer program, e.g. PROII. When calculating the maximum relief rate for emergency depressuring, normally no allowance shall be made for the fact that the initial relief rate at the emergency depressuring valve is more than the relief rate at the flare tip due to the "building-up" of back pressure within the flare system. This phenomenon (often referred to as line packing) can be noticeable in systems with high flowrates and large volumes. Provided proper calculations are carried out advantage can be taken of this phenomenon.

3.6 LOW TEMPERATURE EVALUATION

The lower design temperature shall take into account the low temperatures which can be experienced by equipment and piping both upstream and downstream of a depressuring device when depressuring. The material to be selected is highly dependent on whether it is possible to repressurise the system shortly after depressuring (i.e. creating simultaneous high pressures and low temperatures). Material selection shall comply with DEP 30.10.02.31-Gen.

3.7 DEPRESSURING DEVICE LOCATION

The location of depressuring valves shall be governed by the same considerations as relief valves (2.8.1) and they may discharge into the same disposal system as the relief valves on the equipment under consideration.

Particular attention should also be paid to the position of non-return valves when locating depressuring valves, to ensure that equipment downstream of the non-return valve cannot be isolated from the depressuring valve.

Depressuring devices require periodic testing and hence the depressuring device should be located to allow easy access.

4. FLARE AND VENT SYSTEMS

4.1 GENERAL

In order to ensure safe disposal of flared and vented streams certain factors shall be taken into consideration when designing the pipework upstream and downstream of the relief device. These are covered in this section, together with certain design methods which shall be applied.

Wherever possible, the need for disposal should be avoided (by process changes or by raising the design pressure).

4.2 DESIGN OF PIPING UPSTREAM OF A RELIEF DEVICE

Piping upstream of a relief device should be designed with as few restrictions to flow as possible and should not be pocketed.

The flow area through all pipe and fittings between a pressure vessel and its relief valve shall be at least the same as that of the valve inlet (e.g. isolation valves shall be full bore). Depending on the actual relief valve capacity, the pressure drop of the inlet piping and fittings shall not exceed 3% of the valve set pressure (this is to avoid chatter, which will result in significant seat damage and loss of capacity). Exceptions to this requirement are only allowed in the case of a pilot-operated valve with a suitably arranged remote pilot connection close to the source of overpressure.

The above is especially applicable to relief valves handling gas or vapour. Relief valves in pure liquid service require special attention, since in this case chatter may also be caused by the acceleration of the (non expandable) liquid in the inlet piping: a change in pressure amounting to more than 3% of the set pressure will readily occur and cause valve chatter. In this case the likelihood of chatter can be limited by installing a relief valve with a special liquid trim (linear flow characteristic) thereby avoiding the need to take the relief valve capacity to determine the pressure drop of the inlet piping.

When two or more relief valves (spares not counted) are fitted on one connection, the cross-sectional area of this connection shall be at least equal to the combined inlet areas of the valves, and the above pressure drop requirement shall apply for the combined flow of the valves.

Relief valves on cold process streams shall have an uninsulated inlet line of sufficient length to prevent icing of the relief valve, in particular the disk and spring. Alternatively, heat tracing may be required. Special attention shall be paid in this respect to valves which discharge into the atmosphere, i.e. in those having open outlets which may become blocked with ice.

To avoid the need for special high temperature materials, relief valves on hot process streams may be installed using an uninsulated length of inlet line, creating a cold dead ended leg between the process stream and the relief valve. However, consideration should be given to vapour condensation, deposit formation and solidification, which would affect operation of the relief valve.

4.3 SELECTION OF GAS DISPOSAL SYSTEMS

Streams requiring disposal are:

- relief vapour and/or liquids;
- depressuring vapours;
- any operational waste streams that do not have a more suitable outlet.

In selecting a means of disposal for these streams it is important to find a solution in which all streams are handled with the smallest number and diversity of systems and individual outlets.

The discharge pipe of a relief device shall be sized according to the most severe individual relief condition.

4.3.1 Flaring versus venting

Wherever possible disposal streams shall be collected in a closed system and directed to a flare or vent, except when they can be sent back into the process or stored. In this context, the use of a gas recovery system should be considered.

Considerations to be made in deciding whether to vent or flare the disposal streams are:

- the impact on the environment;
- the safety and integrity of the disposal system, taking into account that disposal streams could contain products which are not combustible;
- local regulations;
- economic evaluations.

Considerations indicating whether venting is allowed are:

- If the release occurs only in an emergency situation.
- If the vapours are lighter than air. Gases shall be considered to be lighter than air if the actual density of the gas after release, taking into account the cooling associated with expansion, is less than 0.9 times the density of the air in the area at 15 °C.
- If the vapours are heavier than air because of low temperature and/or high molecular weight but are in locations where the installation of a flare is impracticable (e.g. product storage areas, marketing depots) or where potential ignition sources are remote (see also DEP 30.06.10.12-Gen.).
- If concentrations of toxic and/or corrosive components in the dispersed vapour cloud do not reach harmful or irritating levels on nearby work levels (platforms) and outside property limits (see 6.4). Calculations of effluent emissions shall be submitted for the approval of the Principal.
- If the risks and consequences of accidental plume ignition (e.g. generation of shock waves) are acceptable.
- If the vapours are such that the condensation of inflammable or corrosive substances cannot occur. This shall be calculated as outlined in API Division of Refining, Volume 43, III. The LODMAT value shall be used in this calculation.
- If the stream does not contain any liquids.
- If the (hot) vented stream cannot self-ignite.

In addition to the above, streams which are not foreign to the atmosphere may be vented without environmental reservations. However, safety near the point of discharge shall be considered, i.e. factors such as temperature, noise, local concentrations of carbon dioxide and nitrogen, etc.

A common vent system may be provided if it is economically more viable or if the requirements for safe venting of relief and depressuring streams cannot be met with the provision of a limited number of individual vent outlets because of the magnitude of the streams or the need for knocking out liquids. In this case it should be assured that the disposal streams do not contain products which, when mixed with other relief streams, may endanger the operation of the vent system through exothermic reactions or the formation of deposits.

4.3.2 Segregated flare systems

Multiple flare system arrangements may offer significant advantages or prove mandatory on analysis of the streams that require disposal.

Segregated flare systems may be required in order to:

i) segregate sources of release into high and low pressure systems. This may be required to accommodate the differing back pressure limitations of individual relief/depressuring devices, or to enable a high pressure low radiation tip to be used with a consequent saving on flare structural requirements. This may also mean that only the low pressure gas requires assistance in order to burn cleanly;

ii) segregate sources with widely differing potentials for liquid release;

iii) segregate sources of cold, dry gas from significant quantities of warm, moist gas and thereby avoid the possibility of freezing and hydrate formation. A relief header after passing a cold stream will be cold. If a warm, moist gas then passes, hydrates could be formed and block the relief header;

iv) segregate corrosive or potentially corrosive fluids (e.g. CO_2 and H_2S) from non-corrosive or moist fluids;

v) meet requirements dictated by the plant geometry or layout and/or economics;

vi) segregate disposal streams which contain products which with other relief streams may endanger the operation of the flare system through exothermic reactions or the formation of deposits.

The selected design should use the minimum practicable number of separate systems but remain operable and safe under all foreseeable conditions. The systems installed may be totally independent, or may share common facilities such as flare knock-out drums and flare tips in certain circumstances.

Streams containing highly corrosive or toxic vapours (e.g. hydrogen fluoride or hydrogen chloride) shall be neutralised before being discharged into the flare system.

When considering the requirement for a high and low pressure disposal system it is necessary to consider the relief valve set pressures present in the system. If there are a large number of high pressure sources with large gas volumes and a relatively few low pressure sources, then generally it would be more economical to install one high pressure relief header and one low pressure relief header. An economical analysis is usually required to ascertain the optimum number of flare systems, and to which system each relief device should discharge.

4.3.3 Disposal back into process or storage

Consideration shall be given to the lining up of relief valves for discharge into an unrestricted, lower-pressure part of the same process system, or into a suitable receiving (storage) vessel.

This line-up is to be favoured for streams which cause problems when flared or vented or, particularly in the case of liquids, where their recovery is of value. It can also be used to temporarily take away the initial high load on a flare relief system during a emergency depressuring situation. The pressure in the receiving system into which the relief valve under consideration discharges generally varies between certain values. The maximum value of this pressure will be taken as the back pressure for the determination of:

- capacities of both balanced and non-balanced relief valves;
- maximum allowed spring setting of non-balanced relief valves.

The maximum value of the pressure in the receiving system shall generally be taken to be equal to the set pressure of the relief valves protecting this receiving system.

However, the pressure in the receiving system can be assumed to be the maximum operating pressure therein if it can be shown that:

- none of the contingencies resulting in operation of the relief valve under consideration would also overpressure the lower-pressure equipment;
- the load imposed by the higher-pressure relief valve would not result in a pressure rise that will exceed the maximum operating pressure in the equipment under lower pressure.

If the discharge of the higher-pressure relief valve is handled by the relief valves of the lower pressure system, it shall be checked that these are adequate for the additional load.

If the relief valve is of the non-balanced type, it shall be checked that the lowest pressure of the process system into which the relief is discharged does not cause inadvertent discharge of the relief valve with its spring setting determined on the basis of the highest back pressure. For this check, the lowest pressure of the process system from all foreseen process conditions (including start-up) shall be used.

4.3.4 Disposal of TERV discharge

TERVs should discharge back into the process, the storage system, or a plant disposal system (for more details see Appendix 2). However, if the discharged liquid cannot be accepted in any of these outlets, TERVs may discharge into an open drainage system, subject to local regulations and to any possible impact on the environment. Light hydrocarbons should be discharged only to a location where vapours may safely disperse, and only with the agreement of the Principal.

Systems containing hydrogen sulphide or other toxic fluids shall never be discharged into an open drainage system.

Liquids which contain components which impair gravity separation, emulsifying agents or spent chemicals which tend to flocculate upon dilution shall not be discharged into the oily open drainage. For open drainage systems leading to a biological waste treatment unit, liquids containing components which will impair biological activity shall be avoided.

4.3.5 Disposal of hydrogen sulphide gas

Streams which are rich in hydrogen sulphide (see the levels below) shall not be discharged into a common HC flare or vent system unless it has been designed for this purpose. This prevents the spreading of sour gas throughout the entire main flare system and also avoids corrosion attack by hydrogen sulphide and the subsequent accumulation of deposits of (pyrophoric) ferrous sulphide.

These streams shall have a separate line-up, preferably a separate flare stack equipped with a tip of the air pre-mix ("Bunsen") type.

Alternatively, the gas may be lined up to the bottom (downstream of the water seal) of the hydrocarbon flare stack, but this should only be done if the hydrogen sulphide rich flow constitutes a minor additional load.

The installation of a separate sour gas flare relief system implies additional capital expenditure. From this point of view it is always better to exclude such a system. The following factors should be considered before deciding that a separate H_2S flare relief system need not be installed, in which case the sour gas release can be tied into the HC flare system:

1) continuous HC release with an H_2S content < 2% by volume;
2) intermittent HC release (only during startup and shutdown) with an H_2S content < 20% by volume, provided this stream is less than 10% by volume of the total continuous HC release rate;
3) emergency HC release (e.g. PZV, emergency depressuring) with an H_2S content < 50% by volume.

The combustion efficiency of a large diameter open pipe flare is poor when gas at low exit velocities has to be flared. Good combustion can only be guaranteed if a minimum exit velocity of 0.5 m/s is maintained. When hydrogen sulphide rich gas has to be flared, incomplete combustion can cause a hydrogen sulphide smell resulting in complaints by people in the vicinity. At a low exit velocity back burning will occur, causing sulphide stress corrosion, especially below the refractory. This means that when H_2S rich gas has to be released into the HC flare system more purge (sweep) gas has to be injected as well on account of the larger size of the flare, which could offset the saving on capital expenditure.

If a hydrogen sulphide flare relief system is used, this shall be heat-traced up to 4 m below the top of the stack. Header materials shall be carbon steel, except for the top 4 metres of the hydrogen sulphide stack, which shall be of AISI 310 S or equivalent.

The knock-out drum for the hydrogen sulphide flare system shall conform to the requirements of section 5.1. Since no water seal vessel has to be installed, the design pressure of the knock-out drum shall be 7 bar (ga). To prevent flashback and consequential detonation purge gas shall be used.

4.3.6 Discharge of oxygen-containing gas

Streams which contain oxygen (or air) shall not be discharged into a common flare or vent system on a continuous basis, unless this system is free from H_2S and no flammable mixtures are created. These streams typically originate from plant sections that operate under sub-atmospheric pressure or vacuum, where some air will inevitably be drawn in.

Even small amounts of oxygen can convert H_2S into elemental sulphur by partial oxidation (Claus reaction). The generation of elemental sulphur in the common flare or vent system could give rise to serious local blockages over a prolonged period. Such blockage will become apparent only during a major relief load, and could remain otherwise undetected.

Flammable mixtures with gas and oxygen (air) shall not be permitted in a common flare or vent system due to the risk of a detonation blast wave inside the header piping. The ignition source in the header could be pyrophoric deposits or discharges of static electricity. This type of explosion could lead to pressures in excess of the header's design pressure and severe loss of integrity.

4.4 FLARE/VENT SYSTEM LOAD ANALYSIS

In order to size the main headers the maximum load that can be expected at any one time shall be ascertained. This requires careful consideration of potential occurrences that could affect several vessels or systems and cause them to relieve simultaneously. The maximum load is not necessarily the largest mass flowrate at any time but rather it is the flow that will impose the highest pressure drop in the system. Thus the temperature and molecular weight of the vapours must be known.

Since the simultaneous occurrence of two or more unrelated contingencies is unlikely, unrelated contingencies should not be used as a basis for determining the maximum system load. Therefore, while lines from individual relief valves should be sized for the maximum calculated design flow, sections of a main header or sub-header should be sized for a specific maximum contingency. Care should be taken to ensure that one contingency cannot remain undetected for a long period of time; otherwise a coincidental second contingency shall be considered.

Each facility (e.g. process unit) will have a maximum relief load that is unique to that facility. Normally, the maximum relief load includes one or more of the following items:

i) relief rates from a total utility or power failure;

ii) effects of a partial utility failure;

iii) maximum flow rate in the event of a single item failure;

iv) maximum flow rate in the event of emergency depressurisation;

v) maximum flow rate in the event of fire.

The impact on the common discharge system design produced by relief flows from connected process units as the result of a utility or power failure shall be calculated by taking 100% of the quantities established under (2.2) for the relief flow of each unit in turn, together with 50% thereof for the other units. The case of low rate depressuring should be included in the assessment. The rationale for choosing 50% of the other units rather than a greater value is because the actions of operators and instrumentation control during failure will tend to limit the peak flows, and stagger the times of occurrence. Moreover not all relief valves will blow at the same time, due to the dynamics of the process. It is hereby assumed that the actual maximum flare relief load is always smaller than that determined by this rational approach. If it is obvious that two units will blow simultaneously, the relief loads should be combined as coming from one unit. The individual relief valve back pressure shall be taken as the highest value in any one of the above combinations.

An alternative to the above is to make a dynamic analysis of the process units which could generate relief loads, taking into account the classification of Instrumented Protective Functions (IPFs), operating conditions, and possible operator interaction. However, especially with a large refinery this could be very cumbersome.

Relief flows and depressuring flows for fire conditions shall be calculated by assuming a fire in only one of the established potential fire areas, thus taking each fire area in turn. Relief and depressuring conditions shall be considered separately, unless it can be shown that the substitution of a relief flow by the depressuring flow from the same process item will greatly increase the back pressure for the other relief valves belonging to equipment in the same fire area and assumed to blow simultaneously.

4.5 SIZING OF DOWNSTREAM PIPING SYSTEMS

Once the maximum design load on each header, sub-header, and lateral has been ascertained it is possible to size the downstream piping system. By starting from the tip of the flare or vent stack where the pressure is atmospheric or critical, and adding each calculated pressure drop, the built-up back pressure downstream of each relief or depressuring device can be determined. For this SIOP's backpressure calculation programme BBGL06 shall be used, in which a roughness factor of 40 µm (assuming the pipe is clean) may be applied. Adjustments in the assumed line sizes may then be made in order to ensure that the operation of the relief or depressuring device is not hindered. If the required piping becomes excessively large, particularly in systems where low back pressures are allowed, it may be preferable to replace non-balanced spring-loaded relief valves with balanced bellows types, thus increasing the maximum allowable back pressure and so meeting the following relief valve selection criteria:

i) Variable back pressure < 10% of set pressure; use non balanced spring loaded relief valves;

ii) Variable back pressure < 21% of set pressure for fire cases and applying equipment following ASME VIII; use non balanced spring loaded relief valves;

iii) Variable back pressure < 50% of set pressure; use balanced-bellows spring-loaded relief valves;

iv) Variable back pressure < 70% of set pressure; use pilot-operated relief valves meeting the criteria as defined in (2.5.2 iii).

The design shall also ensure that if two or more depressuring valves in any process system are opened simultaneously, flow from the high pressure system will not back up into the low pressure system sufficiently to overpressure it or hinder its operation.

To cope with future expansion (e.g. revamps), design velocities in the main relief header should not exceed a Mach number of 0.5. Velocities in subheaders may be higher, up to Mach 1.0. However, since choked flow due to the piping configuration (e.g. elbows, tees, or other discontinuities) could occur at Mach numbers lower than 1, the above margin between back pressure and set pressure should be 10% more than recommended above.

The provision of small branches and instrument connections on flare relief systems shall be avoided, because they are vulnerable to acoustically induced vibration.

4.6 LAYOUT OF DOWNSTREAM PIPING SYSTEMS

4.6.1 Common discharge systems

It is usually simpler and more economic to combine discharges from a number of facilities into a common discharge system served by a central vent or flare.

In the normal configuration of a common discharge system designed for venting or flaring gas at an elevated height, a knock-out drum situated close to the stack is required. The relief valves or depressuring valves installed according to the requirements specified in section (2.8) and (3.7) respectively will discharge via plant subheaders with connections into a main header running outside the battery limits. If this is not possible the flare/vent piping should at least be routed through areas where there is little possibility of a dangerous situation due to local failure of the flare/vent piping (i.e. where possible all piping should be welded).

In the following cases additional knock-out facilities shall be installed within the units:

a) presence of cold flashing liquids in the relief streams, e.g. liquefied gas which may cause blockages in the main knock-out drum or water seal vessel (if installed);

b) presence of liquids at a very high temperature which may cause high stresses due to thermal expansion in the main flare system;

c) presence of liquids with a high pour point which will solidify in the main header or with a high concentration of solids, e.g. catalyst, polymers;

d) to recover liquid relief streams which are expensive or toxic and streams which need to be neutralised before entering the main system;

e) to prevent liquids entering the main flare relief system, since this liquid may be picked up by vapour emergency reliefs from other plants, generating liquid slugs and resulting in high forces at elbows and tees;

f) to overcome problems with header elevations.

Where small quantities of liquids are expected a small drain pot or drip leg shall be installed. Regular inspection of these shall be carried out to prevent blockage of the header. The use of traps or other devices with operating mechanisms should be avoided since they become plugged very easily and have a tendency to freeze.

The disposal piping shall be self-draining towards the knock-out drum. The minimum slope shall be 1:200 for sub-headers and 1:500 for main headers .

If possible, connecting subheaders shall be connected to the top of the header; in any case, they shall drain into the headers. The subheaders shall be connected in such a way that there are no welds in the lower one third of the circumference of the header.

The relief system headers shall leave a free passage to allow access for cranes and other maintenance equipment (see DEP 31.38.01.11-Gen.).

The main flare header should be installed with a pressure device giving an indication in the main control room.
It shall be considered whether thermal expansion loops are required. For new designs, expansion bellows shall not be used.

4.6.2 Sub systems

Where several units are connected to one common disposal system, isolating block valves may be fitted in the subheader from the units, if permitted by local regulations. To assure that these valves are open during operation, they shall be locked open. There shall be a provision for blinding off the line upstream of the block valve. Also a drain and vent with a flange and a purge connection shall be provided to facilitate draining and venting of the isolated branch (see Appendix 6).

For piping sizes up to 400 mm, gate or full bore ball valves according to the piping class can be considered. Gate valves shall be provided with a flush connection. For larger sizes

(≥ 600 mm), butterfly valves may be considered, if there is an economic advantage.

The isolating block valves should be mounted in a horizontal position. By doing this the negative effect of dirt collecting in the valve will be kept to a minimum. Furthermore, in the case of stem fracture (although only a remote possibility), the gate valve will stay open.

4.6.3 Individual vent outlets

Where it is impractical to route discharges to a common centralised vent or flare, consideration may be given to individual local vents.

The location of the vent outlet should be chosen such that:

- the concentration of any toxic products is diluted to a safe level at any area in the vicinity where personnel are likely to be present (see 6.4);
- in the event of accidental ignition of the vent, flames will not impinge upon adjacent equipment and the heat radiation to equipment or personnel will be within the limits of (6.3.2);
- flammable vapours emanating from the vent outlet will be sufficiently diluted (see 6.4);
- the noise requirements are met (see 6.5).

All such vent outlets should be clearly marked on the Hazardous Area classification drawing and the appropriate Hazardous Zones calculated. For vents associated with flammable materials, the end of the discharge pipe shall be cut off squarely and rounded off to minimise the risk of ignition by static electricity, and the pipe earthed. Such vent outlets should be provided with a weep hole (dia = 8 mm) at the lowest point.

The vent velocity should be as high as practicable; whenever feasible not less than 150 m/s at the required relief capacity. For a common vent outlet the diameter shall not be smaller than the outlet of the largest connected relief valve, and is otherwise only governed by back pressure considerations.

4.7 BLOCKAGE DUE TO HYDRATE FORMATION IN DOWNSTREAM PIPING SYSTEM

The blockage of discharge piping downstream of a relief or emergency depressuring valve is not a problem under relieving or depressuring conditions if the discharge is correctly designed.

The correct design of the discharge system should include:

- sufficiently large diameter pipework (velocity < 0.8 Mach);
- short length tail pipes;
- the avoidance of restrictions.

To prevent hydrate or ice formation due to small leaks across the valve or low ambient temperatures, heat tracing shall be installed.

4.8 FLOW MEASUREMENT REQUIREMENTS

Consideration should be given to installing flow measuring devices in each of the main flare headers. The instruments selected should be capable of:

i) identifying significant changes in flowrate in order to assist the operator in recognising the occurrence of upset conditions;

ii) measuring normal operating mass flowrates to an accuracy of ±5% in order assist the operator in monitoring flare and vent losses;

iii) measuring low flowrates in order to quantify purge requirements;

iv) maintainability and possible removal while the flare relief system remains in operation

In selecting the flow measuring device it shall be ensured that the flare header cannot be blocked by the device being installed and that no low points are created within the device or surrounding pipework.

In addition manufacturer's recommendations should be sought with regard to the correct installation of the device particularly concerning upstream and downstream straight run length requirements.

The application of ultrasonic flow meters should be considered, since these meters have a high turn down ratio and low pressure drop.

4.9 PIPING DESIGN

4.9.1 General

All inlet and outlet piping to and from relief valves and depressuring valves shall comply with DEP 31.38.01.11-Gen.

Relief system piping shall be designed and executed to meet the requirements of the piping class that has been specified.

The relief system piping inside plot shall be submitted to hydrostatic testing. For outside plot relief system headers, because of their large diameters and single routing to the flare or vent stack, special test methods may be adopted, e.g. pneumatic testing or 100% X-ray, subject to local regulations.

As a minimum requirement the header and support shall be designed for the presence of liquid. The assumption for the amount of liquid present for a given diameter shall be in accordance with the table below:

Diameters (mm)	Assumption
≤ 250	Full:
300 - 400	1/2 full
450 - 900	1/3 full
≥ 950	1/4 full

4.9.2 Loads due to two-phase flow

Special care shall be taken with respect to the occurrence of slug flow during a two-phase flow discharge.

The possible flow effects of all streams which can enter the flare system shall be considered to identify the maximum forces on the supports. For this a total inventory of all relief streams (e.g. operational, emergency) shall be made and the simultaneous occurrence of liquid and gas releases shall be evaluated. In this connection it should be noted that liquid will stay in the relief header for a certain time period due to the limited slope of the relief header. A vapour relief occurring shortly after a liquid relief will pick up this liquid and it is possible that a slug will be formed, generating high excitation forces. These situations shall be designed out by proper instrumentation or the installation of onplot drainpots or knock-out drums or separate gas and liquid relief headers.

If during the early design stages the possible flow effects of all streams cannot yet be established, one can assume 50 000 N for a 900 mm header and 20 000 N for a 600 mm header as the maximum lateral force as a result of two-phase flow at sonic velocity, provided the two-phase flow rate does not exceed 500 kg/sec. in a 900 mm header and 200 kg/sec. in a 600 mm header. This assumes that no slug flow occurs and the fluid is a homogeneous mixture.

5. KNOCK-OUT DRUMS, WATER SEAL VESSELS AND LIQUID DISPOSAL FACILITIES

5.1 DESIGN OF FLARE AND VENT KNOCK-OUT DRUMS

The objectives of a flare and vent knock-out drum are:

1. to separate liquid from the gas before it is relieved to the atmosphere;
2. to hold the maximum amount of liquid which can be relieved during an emergency situation.

With the above it is important to realise that the maximum gas relief case need not coincide with the maximum liquid relief case. This means that the size of the knock-out drum shall be determined by both the maximum gas relief case as well as the relief case at which a maximum amount of liquid is relieved.

Furthermore, a distinction shall be made as to whether the gas is to be flared or vented. In general a flare stack can handle small amounts of finely dispersed liquids, while a vent stack can only handle dry gas. Therefore the knock-out drum for a vent stack shall follow more stringent criteria and instead of a lambda (λ) of 0.1 m/s recommended for flare knock-out drums, a λ of 0.07 shall be used.

The selection between a horizontal or vertical knock-out drum shall be based on economic considerations taking into account the required slope of the flare header and the maximum amount of liquid which has to be contained.

5.1.1 Gas/liquid separation

The gas/liquid separation efficiency of a knock-out drum is determined by its gas load factor, λ, and calculated from the following equation:

$$\lambda = \frac{Q}{Ag}\sqrt{\frac{Sg}{Sl - Sg}}$$

where:

λ = Gas load factor (m/s)

Q = Gas flowrate (m^3/s) at operating conditions

Ag = Area available for gas flow (m^2)
for vertical vessels, Ag is the cross sectional area of the vessel
for horizontal vessels, Ag is the cross sectional area of the gas cap available above the LA (HH) level (see also 5.1.2).

Sg = Gas density (kg/m^3) at operating conditions

Sl = Liquid density (kg/m^3) at operating conditions

Appendix 10 indicates how the gas load factor relates to the smallest droplet size still separated.

For the purposes of gas/liquid separation, and taking into account the various situations, the designer shall determine which of the following requirements are applicable and the most definitive.

1. **Absolutely dry gas**
If the relief flow is an absolutely dry gas, the possibility could be considered of relieving this gas through a separate (absolutely dry) gas relief line connected directly to the flare or vent stack, bypassing the flare or vent knock-out drum.

2. **Essentially dry gas**

Amended per Circular 32/98

If the relief flow (e.g. depressuring) is essentially dry or if an inside battery limit knock-out drum is used for those units or plants that are expected to produce significant two-phase flow quantities, a λ value of 0.25 m/s can be applied. Greater λ values could lead to entrainment of resident liquid in the knock-out drum.

Alternative designs should be considered for the approval of the Principal if the flare knock-out drum would otherwise become disproportionately large (e.g. diameter larger than 6 m). In this case, if λ becomes larger than 0.3 the design shall be subjected to computational fluid dynamic analysis.

3. Two phase flow

If significant liquid quantities are expected during the major relief case and no inside battery limit knock-out drum is used, a λ value not greater than 0.1 m/s shall be taken. If a schoepentoeter is installed a λ value of 0.15 m/s can be used. However, the schoepentoeter shall be of a sturdy design to be able to cope with high loads. Reference is also made to DEP 31.22.05.11-Gen.

Since λ depends on the conditions of the gas (e.g. gas density, liquid density, etc.) it is important to determine which two phase flow condition is the most definitive.

As indicated in the above equation, the λ value depends on the area available for gas flow (Ag). For vertical knock-out drums this will be the horizontal cross sectional area. For horizontal knock-out drums this will be the vertical cross sectional area of the gas gap above the normal maximum operation level (LA (HH)) and not above the maximum level the liquid will reach during a maximum liquid relief, unless it is obvious that a major gas relief coincides with the major liquid relief or occurs shortly thereafter. In this case the available cross sectional area for the gas cap for proper vapour/liquid disengagement shall be above the liquid level reached after the major liquid relief.

If the horizontal knock-out drum is provided with two inlets, the flow (Q) defining the required cross sectional area shall be 0.6 times the maximum relief flow. This takes into account any maldistribution of the flow entering the knock-out drum. In this case the vessel shall be of sufficient length (L > 5 D).

Since the gas velocity in the knock-out drum declines with increasing pressure the size of the knock-out drum can be reduced by maintaining a higher operating pressure. This can be accomplished by allowing choked (sonic) flow in the knock-out drum discharge piping or at the vent or flare tip. This shall take into account plant noise levels and maximum allowable back pressures of the relief valves.

5.1.2 Liquid hold up and pump out capabilities

Amended per Circular 32/98

The liquid space in the knock-out drum depends on the maximum amount of liquid the knock-out drum has to contain during an emergency situation. Bear in mind that during normal operation the vessel may already contain liquid to facilitate proper control and pump operation and the liquid relieved during an emergency has to be accommodated on top of this. This means that the area above the LA (HH) level (5.1.2.5) is available. Furthermore the type of liquid in terms of temperature, viscosity, solidification point etc., shall be taken into account. Attention shall therefore be paid to the following:

1. Liquid space on top of LA (HH) liquid level

This space shall be designed to contain the maximum emergency liquid relief rate for a period of fifteen minutes. Where liquid relief is likely to continue for longer than fifteen minutes, for example with unattended operation, the hold up time should be increased accordingly. The period of fifteen minutes has been selected on the basis that corrective measures to control the relief will be taken within this time. If a horizontal knock-out drum is used, it may be considered to have a liquid boot at the bottom. The LA (HH) level could then be set in the top part of the boot, in which case the full vertical cross sectional area of the knock-out drum is available for gas/liquid separation.

2. Maximum liquid level

After the maximum emergency liquid relief:

a) in vertical knock-out drums, the liquid level shall remain one inlet pipe diameter or 0.3 m, whichever is greater, below the bottom of the inlet pipe;

b) in horizontal knock-out drums, the liquid level shall remain below the horizontal centre line of the vessel.

3. Pump out capacity

For refinery service the knock-out drum shall be equipped with electrically driven pumps (one operating, one spare) capable of emptying the drum in two hours. Since the hold up volume needs to be sufficient to contain the maximum amount of liquid generated during any emergency situation, a secured power supply for the pumps is not necessary.

Appendix 3 shows the pumps operating on “on/off”, to control the normal level. This is the recommended mode of operation for operator convenience, but not essential since a period of two hours is acceptable for pump-out.

If it is justified not to install a pump (instead relying on auto-evaporation and/or occasional pump-out using a gully sucker and/or drainage to a safe location), the level required for pump control need not be taken into account and the required control level may be less than that determined in (5.1.2.5).

If pumps are applied it shall be ensured that liquid back-flow cannot occur from the disposal system back into the flare liquid knock-out vessels, either through gravity flow from storage or from pressurised disposal systems or back through the pump in the standby operation mode. This will avoid pumps and/or alarms being actuated unnecessarily. It should thus be verified whether single check valves are sufficiently reliable or whether heat tracing is required to guarantee their proper operation, or whether further redundancy may be required.

In cryogenic service (LNG) the use of blowcases instead of pumps should be considered.

4. Heating coil / heat tracing

If the liquid could solidify (i.e if the pour point is above LODMAT) then electrical or steam heating shall be provided, with or without temperature control. In this case external heat tracing shall also be provided to the piping upstream and downstream of the pumps.

If the liquid is volatile (e.g. liquid propane) heating shall be provided to vaporise the liquid. The heating may be provided by a steam coil, but the possibility of (steam) condensate freeze-up shall be eliminated and an adequate steam trap shall be installed.

5. Instrumentation

The instrument requirements are specified on the basis of DEP 31.22.05.11-Gen. taking into account the specific requirements inherent to the operation of a flare knock-out drum.

In the text below, LS/LSA are mentioned since it is recommended that the pump-out system is automatic, with pump start on LS (H) and pump stop on LS (L). It is essential that the specified distances are applied irrespective of pump-out mode, since it will ensure the design of a proper volume in the vessel for normal level control.

a. LSA (LL) is 0.15 m above BTL (vertical vessel) or 0.15 m above vessel bottom (horizontal vessel).

b. LS (L) is 0.20 m above LSA (LL) or located such that there is sufficient liquid hold-up time between LSA (LL) and LS (L) to prevent any nuisance alarm or pump trip; alternatively the liquid volume between LS (L) and LSA (LL) should be 60 seconds pump out capacity, whichever requirement is the more stringent.

c. LS (H) is 0.20 m above LS (L) or the liquid volume between the two levels shall be at least 5 minutes pump out capacity, whichever requirement is the more stringent.

d. LA (HH) is 0.15 m above LS (H) or 60 seconds pump out capacity, whichever is the more stringent. The alarm is required to give the operator a timely alarm that the pump out facilities are not operating properly or that a major liquid relief is entering the knock-out drum.
   This alarm is critical and shall have proper safety integrity and availability in compliance with an IPF classification in accordance with DEP 32.80.10.10-Gen., which will determine the correct hardware and testing frequency. The alarm should also have a very high priority within the framework of the site’s alarm management system.

e. TSA (L). The purpose of the low temperature switch is to prevent a pump from being started if low temperature liquid has collected in the flare knock-out drum. This liquid could be volatile (e.g. propane or butane) and cannot be transferred directly to the slops tank, since the contents of the slops tank could be too hot, generating too much vapour which cannot be handled by the off gas system of the slops tank.

f. TSA (H). The purpose of the high temperature switch is to prevent a pump from being started if high temperature liquid has collected in the flare knock-out drum. The high temperature liquid could generate too much vapour when pumped into the slops tank. As an alternative passing the liquid through a cooler to the slop oil tank may be considered.

5.1.3 Other requirements

Nozzle sizes
The momentum criterion (mix density * two-phase flow velocity squared) for nozzles provided with a half open pipe shall not exceed 5 000 N/m^2. With a schoepentoeter inlet device this value can be increased to 10 000 N/m^2.

Design pressure
The knock-out drum shall be designed as a pressure vessel with a design pressure of at least 3.5 bar (ga). If no seal vessel is used, the design pressure shall be at least 7 bar (ga).

5.1.4 Example drawings

Several rules are specified above to arrive at a suitable design of the vent or flare knock-out drum. Appendix 10, Figures 1, 2 and 3 give typical layouts of knock-out drums to assist in fulfilling these requirements.

5.2 WATER SEAL VESSELS (SEE APPENDICES 3, 4, AND 5)

The purpose of the water seal vessel is:

1) to prevent any flashback, initiated from the vent or flare tip and propagated further upstream of the water seal vessel;

2) to prevent air ingress due to a sudden temperature change of the flare and relief system;

3) to maintain a slight overpressure in the flare system to ensure that air will not enter the system and also, this may be necessary if a flare gas recovery system is in use.

The design of the water seal vessel, see Appendix 5, shall be based on the maximum vapour quantity to be released. Appendix 4 gives rules for its sizing.

If two or more flare stacks operate in parallel, each flare stack should have its own dedicated water seal vessel. At a low gas relief rate, one stack shall burn preferentially and to achieve this the dip legs of the various seal vessels should be set staggered.

The seal vessel shall be equipped with a skimmer (intermittent operation); see Appendix 3.

For LNG plants no water seal vessels shall be used, since in the event of a cold release this will form an obstruction in the flare relief system. If two or more flare stacks operate in parallel, anti-flashback devices should be installed at the bottom of the flare stacks. If one flare is spare to another flare, an anti-flashback device is not necessary. However, operating procedures shall prevent flashback (e.g. higher purge rate) when the operational flare is switched to the spare flare.

5.3 LIQUID DISPOSAL FACILITIES

Depending on required capacity, nature of the liquid (e.g. viscosity, pour point), economic value, expected frequency and duration of disposal, different systems as described below may be selected to dispose of the liquid. These systems should be evaluated on their suitability and economic merits.

5.3.1 Process feed vessel

If the emergency liquid release is related to a certain process, it may be beneficial to release it back to its feed vessel or other suitable vessel. It shall be ensured that the receiving vessel will not be overfilled or over pressured.

5.3.2 Release in hydrocarbon relief system

If liquid is released into the hydrocarbon relief system, the following shall be considered:

Even a small quantity of liquid present in the hydrocarbon relief system could form slug flow at a high gas release, generating high dynamic forces on elbows, tees, reducers, etc. Therefore, the standing presence of liquid in a flare relief system shall be prevented. For this reason the correct slope of the flare header (1:200 in plant; 1:500 off plot) is essential.

The coincidence of a liquid and a major gas release should be evaluated to assure the suitability of the relief piping. For this, a complete inventory (matrix method) of all potential liquid and gas streams feeding the flare header shall be made. Since liquid is not immediately drained away through the long flare relief header, a gas release occuring even a considerable time after the liquid release could generate slug flow. Depending on the flare header diameter, just a few centimetres of liquid could be enough to cause this.

If excessive forces may occur, a separate gas relief and liquid relief header up to the knock-out drum, installed either at the battery limit or within the flare area, shall be considered.

5.3.3 Slop Oil storage

Amended per Circular 32/98

Small quantities of liquids, occasionally released and collected in the flare knock-out drum, should be sent to slop oil storage. In this case it is important to bear in mind the following:

1) Sufficient storage shall be available.
2) The liquid shall not be too volatile, because this could generate too much vapour in the receiving slop oil storage vessel.
3) The liquid shall not be at too high a temperature (> 70 °C), because this could evaporate the liquid present in storage. If a temperature above 70 °C cannot be avoided, the use of an intermediate cooler should be considered.
4) The pour point of the liquid shall not be too high, since the liquid could solidify in the transfer line to the slop oil storage tank, plugging the disposal system.
5) Toxic gases (e.g. hydrogen sulphide) entrained with the liquid should be disposed of safely via the slop oil tank vents, ensuring that this will not have adverse environmental consequences.

5.3.4 Evaporators

Liquids which are volatile may be evaporated. This is achieved by either installing a heating coil in the flare knock-out drum or using an external heat exchanger. In this case the liquid is disposed as vapour together with the other gases released to the flare.

5.3.5 Liquid Disposal Burners

Liquid burners may be used if large streams of liquids (e.g. off-specification products) have to be disposed of over prolonged periods and there is no economic incentive to recover this liquid for other purposes.

The same criteria (e.g. radiation levels, noise levels) as outlined for flare stacks and flare tips (see sections 6 and 7) are applicable. To obtain a high combustion efficiency and improve the turn down capability of the liquid burner, assist gas shall be employed.

If required, adequate purge provision shall be provided to prevent flashback.

5.3.6 Burn pits

Burn pits produce considerable smoke and therefore their use shall be considered in exceptional cases only, such as when liquids are to be disposed of occasionally.

A burn pit has a storage capacity, and hence drain lines can be routed directly to the pit. It shall have its own sterile area of at least 100 m radius, not to be combined with the sterile area of the main flare.

All piping to burn pits or liquid disposal burners shall be protected against fire by either burying them or by putting them in a trench covered by concrete slabs.

If required, adequate purge provision shall be provided to prevent flashback.

6. STRUCTURES FOR FLARE AND VENT STACKS AND LIQUID BURNERS

6.1 GENERAL

The type and height of the structures supporting flare or vent stacks or liquid burners depend on the following operational and environmental aspects:

i) required availability of the flare and relief system;
ii) acceptable heat radiation levels;
iii) acceptable dispersion levels;
iv) acceptable noise levels.

6.2 TYPE OF STRUCTURES

The various types of structures to support a flare stack (or vent stack) are:

1) free standing stack;

2) guyed stack;

3) derrick structure;

4) boom structure installed at an angle (especially on offshore platforms).

The structures shall comply with DEP 34.00.01.30-Gen., DEP 34.24.26.31-Gen. and DEP 34.28.00.31-Gen.

The type selection is based on economical and operational grounds.

If only one stack is required, any of the four types mentioned above may be selected.

However, it shall be recognised that the flare tip has a limited service life and therefore inspection and dismantling of the tip shall be feasible. This can be done by a crane (when locally available) or a davit. With the latter it shall be taken into account that a davit installed near the tip will be exposed to the flames of the flare, impairing its availability and safe operation. Therefore it shall be possible to retract the davit underneath a heat shield to protect it against excessive heat radiation. The radiation level shall be less than 15.8 kW/m^2.

For refineries, a 100% availability of the flare system may be required. In this case at least two flare stacks should be installed (one operating and one spare). These stacks should be installed in one derrick structure and it shall be possible to retract one flare stack while the other remains in operation, without personnel having to work above the riser removal/replacement platform. A typical arrangement is displayed in Standard Drawing S 28.028. The 100% availability makes it necessary to design a flare structure, above the riser removal/replacement platform, requiring no maintenance whatsoever during its entire service life (e.g. > 30 years). Material and selection of the coating system shall be such as to meet this requirement.

A derrick structure shall be supplied with the following access equipment:

a) a stairway from grade to the riser removal/replacement platform;

b) a ladder from the riser removal/replacement platform to the top platform;

c) step-off platforms at intervals of 9 m to the top platform at the flare tip;

d) platforms with ladder access for all manways and handholes;

e) ladders from the riser removal/replacement platform to all eyes attached to the derrick structure that are used to raise, lower and tilt the riser sections.

Riser removal/replacement platforms may comprise more than one level and shall be of a retractable type.

The exact elevation of the riser removal/replacement platform will depend upon the number of riser sections and the elevation of the hoisting blocks and associated eyes that are attached to the support structure to raise, lower and tilt the riser sections.

The distance between the burner tips and the top of the structure shall be laid down by the flare system supplier. The heat radiation from the burner tips shall not affect the structure itself, nor the conservation system or personnel. Suitable heat radiation screens may be considered to achieve this.

Surface preparation and surface protection shall be in accordance with DEP 30.48.00.31-Gen.

Detailed calculations shall be carried out by the Manufacturer to verify the integrity of the system. If the whole (refinery) complex is planned to be shut down every 3 to 5 years, no spare flare is necessary.

For aviation warning lights requirements see DEP 34.24.26.31-Gen.

If a retractable flare stack system is selected, the aviation warning lights shall be retractable without shutting down the flare relief system.

6.3 HEAT RADIATION LEVELS

6.3.1 Calculation Method

Amended per Circular 32/98

The calculation method to determine the heat radiation levels of burning flares is given in API RP 521. The basic equation is as specified below:

$$K = \frac{\tau F Q}{4 \pi D^2}$$

where:

K = Heat radiation level (kW/m^2)

τ = Fraction of heat intensity transmitted through the atmosphere

F = Fraction of heat radiated

Q = Heat released related to Low heating Value (kW)

D = Distance from midpoint of flame to the object considered (m). To determine this point a wind speed of 10 m/s shall be assumed.

The results of the calculations greatly depend on the factor F. Parameters influencing this factor include the composition of the gas, the exit velocity of the gas and the geometry of the burner.

A limited number of experiments have been carried out to determine this factor. For a first approach, the F-factors given in Appendix 7 should be used in the above equation.

Since manufacturers have their own calculation methods and the F factor also depends on the geometry of the tip, manufacturers shall confirm that the specified radiation levels are met at the given height and other conditions specified in the requisition.

If the effect of the stack is critical (e.g. if there are very high relief loads or if the radiation level close to the flame has to be determined) the Shell hazard consequence analysis package ("FRED") should be applied (see Report OP 97-47088).

6.3.2 Acceptable heat radiation levels

The acceptability of heat radiation levels is dependent on:

1) the effect on humans, and;

2) the effect on equipment.

Normally, the only equipment allowed in the flare's sterile area shall be that directly related to its operation, such as knock-out drums, pumps, valves, etc. Special attention shall be paid to construction materials (for example, either eliminate or ensure suitability of aluminium or plastic), heat sensitive streams (no open oil sewers generating flammable vapours). Electrical equipment and instrumentation shall be able to withstand the heat radiation in the sterile area.

Taking into account topographical and meteorological conditions, the height of the flare stack shall be selected to meet the following conditions:

1) The sterile area radius should be 60 m.

2) At the boundary of the sterile area the heat radiation level shall be 6.3 kW/m^2 maximum (excluding the effect of solar radiation).

3) At the property limit the heat radiation level shall be 3.15 kW/m^2 maximum (excluding the effect of solar radiation).

To determine the above maxima, the exposure times needed to reach the pain threshold as outlined in API RP 521 have been taken as a basis with the exception that the effect of solar radiation can be excluded, since its spectrum is better accepted by the skin and a 100% addition is considered not to be realistic.

In the maximum emergency release case, heat radiation levels will exceed the 6.3 kW/m^2 limit within the sterile area. Since flare-related equipment is normally located within the sterile area and sometimes requires maintenance, personnel could be exposed to radiation levels higher than 6.3 kW/m^2. Such a situation could also occur when a flare stack is being taken down. Personnel will be present on the first platform to unbolt the flange connections of the different stack sections. Consequently, a proper heat shield at the first platform and temporary shelters at grade shall be provided within the sterile area to protect personnel.

The same guidelines as applied to flares shall also be applied to (ignited) vents, with the exception that a sterile area is not required. Shelters shall be provided if personnel could be exposed to radiation levels higher than 6.3 kW/m^2. Equipment installed in the vicinity may be used to serve this purpose. To limit the level of heat radiation, advantage shall be taken of the fact that at high exit velocities the F-factor is lower. Taking into account noise criteria (refer section 6.5) and back pressure requirements (refer section 4.5), the use of sonic tips shall be considered in order to achieve a cost effective design.

For liquid burners, the same guidelines as specified for flare stacks are applicable, except that the height of the structure may be less. To determine the heat radiation an F-factor of 0.3 shall be taken.

6.4 DISPERSION LEVELS

The (emergency) release of toxic and/or flammable streams shall be at a safe location. This means that they must be dispersed in such a way that personnel present at nearby work levels (e.g. platforms, tank roofs, etc.) are not exposed to a hazardous situation.

The following shall be observed in order to meet the above criterion:

1. Dispersion shall be such that within the hazardous contour (the area within which either an ignition source or personnel could be present):

   a) the concentration of flammable components is less than the lower flammability limit. and

   b) the concentration of toxic components is less than the TLV-TWA (Threshold Limit Value - Time Weighted Average).

2. No noticeable stench or irritation levels shall be caused outside the property limits. Proper operation of the flare (combustion efficiency > 98%) may be assumed.

6.5 NOISE LIMITS

Amended per Circular 32/98

DEP 31.10.00.31-Gen. shall apply.

For emergency conditions:

- The noise level at the base of the stack shall not exceed 115 dB (A). If the stack is provided with a derrick structure, including a platform for coupling/uncoupling segments of the retractable stack, the noise limit applies to this platform.

For normal operation (including starting-up and shutting-down):

- Noise levels at the perimeter of the sterile area shall not exceed 85 dB (A) at flow rates up to 15% of maximum flaring capacity or at the maximum relief rate that may occur during normal operation (including starting-up and shutting-down), whichever is the higher.
- If there are limits on the allowable noise levels outside the plant then the sound power level generated during normal operation shall be taken into account when assigning sound power levels to noise sources.

7. FLARE AND VENT TIPS

7.1 GENERAL

This section is concerned with the general design of flare and vent tips. Specific proprietary flare tips are not covered in this DEP. To obtain specific requirements for a particular type of tip it will be necessary to contact the manufacturer concerned.

7.2 FLARE TIP DESIGN CONSIDERATIONS

A flare tip should be selected with the aim of:

(i) Improving combustion to reduce the proportion of unburnt gases released to the atmosphere.

(ii) Reducing radiation levels. If more than one flare stack is required the tip centre-to-centre dimensions shall be selected such that the heat radiated from an operating tip has no detrimental effect on adjacent tips.

(iii) Reducing or eliminating smoke formation.

(iv) Ensuring that the flared gases burn with a stable flame over the whole operating range.

(v) Reducing to a minimum the maintenance required over its operating life. For maintenance purposes, provisions for lowering either the entire flare stack or the flare tip alone shall be agreed upon with the Principal.

(vi) Meeting permissible noise levels, which are specified in (6.5).

In order to improve combustion, reduce smoke formation and reduce heat radiation, high speed flare tips (sonic tips) should be used wherever possible. Manufacturers should be contacted for more detailed data on particular types of tip.

The suppression of smoke formation is achieved by good pre-mixing with an excess of air so reducing the release of elemental carbon through the flare. If there is a deficit of entrained air due to a low exit velocity (e.g. pipe flare) then air can be forced into the gas using fans or steam.

If one of the above is considered necessary in order to achieve a smokeless flame over the whole operating range of the flare an economic analysis of the above methods shall be performed to find the most suitable. As a minimum the flare system shall be designed to produce a non-luminous and smokeless flame which meets Ringelmann No. 1 criteria (BS 2742) for the maximum continuous flare flowrate or to cover an acceptable range of emergency flaring (which is 15% of the maximum flaring capacity).

Where steam injection is selected medium pressure steam (approximately 17 bar (ga)) shall be used. The injection of excess steam wastes energy, can create a very noisy flame and may even cause the flame to be extinguished. It is therefore essential to ensure that the correct quantity of steam is injected.

As a first estimate the table given in Appendix 8 may be used to obtain the required steam flowrates. A typical flare gas composition should be used in performing the calculation. Fine adjustment of the steam flowrate may be performed during start-up. Steam supply shall be controlled either manually from the control room, or automatically on ratio from a flow measurement of the gases to the flare. It shall also be possible to supply steam remotely controlled from the control room with zero gas flow.

Damage is most commonly caused to a flare tip when operating at low flow rates due to flame impingement on the inside and outside of the tip which causes thermally induced stresses. External burning of the tip is generally influenced by the wind force itself and by low pressure zones caused by the wind. All tips shall be fitted with flame stabilisers and wind deflectors and in addition the following points shall be considered in order to increase service life and thus reduce maintenance costs and loss of production:

(i) The use of refractory linings, provided with adequate supports, on the inside of flare tips (see DEP 64.24.32.30-Gen.).

(ii) The use of windshields or deflectors which breakup the low pressure zones created by the wind and physically stop the flame from coming into contact with the tip.

(iii) The use of multipoint flares which reduce or eliminate low pressure zones around the tip, due to the smaller diameter of the individual burners.

(iv) The use of the available flare tips in increments related to the amount of gas to be flared. This can be obtained by proper setting of the dip legs in the water seal vessels or by applying a staged control of the number of burners of a multi-tip flare.

(v) Pushing the flame away from the tip by the use of a supplementary flow, such as compressed air, to overcome the forces exerted by the wind.

(vi) Upgrading the material specification for the flare tip beyond the following requirements.

The material of the flare stack tip shall be sufficiently heat and corrosion resistant, e.g. AISI 310S, Incoloy 800H or equivalent. For offshore applications a material highly resistant to chloride corrosion (such as Inconel 625 or equivalent) should be used.

All ancillaries connected to the tip of the stack shall be of the same material.

In view of the high cost of Incoloy 800 H or Inconel 625 material, the inlet flange of the tip may be made of carbon steel if approved by the Principal.

NOTE: This is justified since the hottest area is at the discharge end of the tip and heat generation lower in the tip due to back burning is prevented by applying a proper purge rate.

Particularly high levels of heat emission from the flame will occur only under high gas flows and consequently the cooling effect of the gas rising through the stack together with the induced higher air flow around the tip will then be greatest, thus keeping the inlet flange sufficiently cool.

The windstrakes fitted to the tip shall be of the same material as the main body of the flare. Any carbon steel materials should be coated with thermal sprayed metallic aluminium and the bolts for mounting the tip to the rest of the flare shall be of a high nickel alloy. To cope with thermal expansion the windstrakes shall be attached to the tip at one fixed point. Other connection points shall be of the sliding type.

8. FLARE AND VENT PURGING

8.1 GENERAL

This section defines the criteria applicable to the purge rates required in order to prevent oxygen ingress and the possibility of detonation within the flare and vent system. The passive aids which may be utilised are also considered.

8.2 PURGING DESIGN CONSIDERATIONS

Purging of a flare or vent system shall be considered, taking into account the following:

1. Oxygen ingress can lead to the formation of flammable air/fuel mixtures in the stack, which when ignited will cause a flashback. This is most likely to limit itself to a deflagration but under certain conditions could result in a detonation.
2. Oxygen ingress can lead to the formation of deposits (partial oxidation of sulphur compounds) causing flare blockage.
3. If too little purge gas is used it can lead to back burning or a licking flame, reducing the service life of the flare tip.
4. If too little purge gas is used an unstable flame could result in inefficient burning causing an obnoxious impact (stench) on the surroundings.
5. If fuel gas is used as purge gas, this will have an impact on the environment.
6. If the relief system has to handle corrosive gas or gas prone to condensing or solidifying then, as well as purge gas, sweep gas shall be injected (either continuously or intermittently) at strategic locations in the flare relief system. For an H_2S rich disposal system, see also (4.3.5).

Air ingress into the stack may occur as follows:

a) diffusion of air down into the stack;
b) wind action across the tip at low flowrates, resulting in a differential pressure at the top of the stack;
c) relief gas with a lower density than air; this will create a problem especially when two or more stacks are operating in parallel. The relief gas could tend to leave through one stack only, while the heavier air will enter through the other stack and mix with the relief gas, creating an explosive mixture;
d) condensation and/or shrinkage of the contents of the relief system resulting in an underpressure within the relief system. This may be caused by an increase in heat removal as a result of a hot release or a rain shower on the header. This shrinkage could be considerable after a major hot, heavy gas relief;
e) during a plant (relief system) shutdown, during which some connections are open and the purge system is inoperative for a prolonged period.

To prevent air ingress due to a) and b) the recommended minimum purge rates at different vent (flare) stack diameters and purge gas molecule weights are as presented in Appendix 9.

Detailed calculations show that considerable purge gas rates are required to prevent air ingress due to phenomena c) and d) and purge gas rates as specified in Appendix 9 are then inadequate. If these phenomena can occur a water seal vessel shall be installed upstream of each flare stack and the purge gas should be injected downstream of the water seal vessel. The design of water seal vessels is covered in (5.2).

If cold gas (e.g. in NGL plants) can be relieved, water freezing can occur and the water shall be replaced by glycol. If very cold gas (e.g. LNG plants) can be relieved a seal vessel, even if filled with glycol, is not recommended. In this case, if two stacks have to operate in parallel, and an anti-flashback device at the inlet of the stack should be installed. With single stack operation no anti-flashback device is required, since it is realistic to assume that phenomenon d) will not occur in LNG plants only handling cold gases at sub ambient temperatures.

The above recommendations only have a direct effect on item 1) and indirect effects on items 2), 3), 4), 5) and 6). Recommendations for dealing with the effects of items 2), 3), 4), 5) and 6) are difficult to specify in this DEP and depend on the actual situation. Factors of influence are:

- The type of purge gas available; hydrogen rich gas has a higher tendency to back burning.
- The economic value of purge gas.
- The direct impact on the environment; residential areas require greater attention; purging with an inert gas has less environmental impact but does not support incineration and is not compatible with fuel gas.
- The composition of the gas to be flared; H_2S rich gas has a higher tendency to form deposits; inefficient combustion will easily cause an obnoxious odour.
- The layout of the vent and flare system.
- The presence of a flare gas recovery system.
- The type of flare tips used; tips with refractory lining are more susceptible to back burning.
- The flaring philosophy; for a non flaring refinery the use of an inert gas as purge gas may be attractive. To control the amount of (inert) purge gas the use of oxygen monitors may be considered.

These factors shall be taken into account to arrive at an economic and environmentally acceptable solution.

8.3 PURGE REDUCTION SEALS

Several types of device have been developed to reduce, but not eliminate, the overall purge gas requirement.

Gas seals of the labyrinth type (e.g. molecular seal) shall not be used as they are easily blocked, very heavy and easily damaged by shock loads, such as those occurring at the start of emergency depressuring.

Only gas seals that prevent the infiltration of air along the wall of the stack by returning air to the unrestricted central zone of the stack shall be used. Unlike the labyrinth type of seal, this type of seal offers no protection against air ingress in the event of interrupted purge flow. Manufacturers shall be contacted in order to ascertain purge rate requirements for a particular type of seal.

Seals shall be located close to the flare tip base flange for ease of inspection and maintenance.

8.4 FLAME ARRESTORS

Flame arrestors shall not be used as an alternative to a continuous purge for flash back protection because they:

- are susceptible to blockage;
- are susceptible to undetected mechanical damage;
- provide an obstruction to flow;
- become ineffective within a few minutes of ignition due to heat build up if located near the vent tip.

Flame arrestors should only be used if the use of purge gas is not feasible and the discharge fluid is clean. If used, flame arrestors shall be placed near to the tip of the vent, but still be accessible for maintenance and inspection. This is to prevent explosions occurring in the vent pipe.

The manufacturers should be given full details of the intended duty and location of a flame arrestor so that a suitable selection can be made. This is another reason for siting arrestors at or near the end of vent pipes since long pipes would otherwise allow the flame front to accelerate.

Regular inspection shall be carried out on all installed flame arrestors, particular attention being paid to arrestors which do not have a constant vent stream through them, those which are used on an inbreathing service and those used on any service where the risk of blockage is relatively high.

Consideration may be given to providing a constant nitrogen purge stream through normally non-flowing flame arrestors so that blockage may be detected early.

A flame arrestor is not required directly downstream of:

- relief valves releasing to atmosphere;
- P/V valves on tanks,

since these have a flame arresting action and should not have restrictions placed downstream.

9. VENT SNUFFING

9.1 GENERAL

The fitting of a remote controlled snuffing system on all vent stacks should be considered in order to avoid continuous burning in the event of accidental ignition of the vented gases.

9.2 VENT SNUFFING REQUIREMENTS

The snuffing medium may be carbon dioxide or steam (if available); Halon or other CFCs shall not be used due to their adverse effects on the environment.

The snuffing system shall be operated from a manual station. Once the flame is extinguished, the control system shall ensure that metal temperatures at the tip of the vent drop sufficiently to prevent spontaneous ignition of gas and the danger of flashback.

The snuffing facilities shall be sized to extinguish the stack at least three times in succession when it is burning and discharging at a rate corresponding to one percent of the maximum vent rate.

10. FLARE PILOTS AND IGNITION

10.1 GENERAL

All flare systems shall be provided with continuous pilot burners to ignite the flare gas as it leaves the tip. The pilots shall each be provided with an ignition system in case they are extinguished.

10.2 FLARE PILOT REQUIREMENTS

The pilots provided at the flare tip shall be capable of sustaining stable combustion under all flaring and meteorological conditions. It is therefore recommended that all pilots are checked using a flame stability model.

For pipe flare tips up to 400 mm diameter, at least two pilot burners shall be provided. Three pilot burners shall be provided for flare tips larger than 400 mm. Each of these burners shall be ignited by means of an individual ignition line. For proprietary flare tips the manufacturer's proposal should be followed.

The location of the pilot burners should be such that they are not engulfed by the flame of the main flare even in strong winds, whilst ignition of the main flare is guaranteed under these conditions.

The materials and design of the pilots and their method of support should be such that they require minimum maintenance and are suitable for at least a five year service period. The ignition lines to the pilot burners shall be stainless steel (AISI 321) to prevent the possibility of internal corrosion, with the top 4 metres being AISI 310S, Incoloy 800H or equivalent. Emphasis should be placed on the provision for differential expansion between the flare stack and the pilot lines, with particular attention being paid to the support brackets.

The use of advanced pilot systems is recommended. These systems are equipped with a venturi device in the pilot nozzle which allows a reduction in fuel gas consumption and increases pilot gas exit velocity and so enhances air entrainment and improves flame stability. Manufacturer's recommendations should be followed concerning the quantity of fuel gas required for proprietary pilot light systems. The pilot flame can be further stabilised by the installation of wind shields around the pilot nozzles.

The fuel gas and the air used to supply the pilot lights and ignition system shall be dried and filtered to prevent blockage of the lines. The hydrocarbon and water dew points of the fuel gas and the water dew point of the air shall be such that condensation is not possible whatever the mode of operation. The filter shall be installed between the carbon steel fuel gas supply line to the flare and the stainless steel supply lines to the pilots and ignition system. In addition the gas should be of a constant composition as a change in the Wobbe Index of the gas will affect flame stability.

The pilot gas supply lines shall be arranged without pockets and any build-up of condensation shall be alarmed at the control room before blockage can occur.

One stainless steel fuel gas supply line along the stack is required for each pilot burner. A filter shall be installed between the carbon steel fuel gas supply line to the flare and stainless steel fuel gas supply lines to the pilots.

One 'K' type thermocouple shall be provided for each pilot. Failure of a pilot shall be indicated in the control room by an alarm. It shall be ensured that the thermocouple is not damaged by direct heat of the flare or pilot flame. Therefore its design shall be robust and adequate insulation shall be provided.

Regular maintenance and inspection of the pilot light system is essential.

A back-up bottled gas supply shall be provided for start-up if there is no other reliable source available.

10.3 FLARE IGNITION REQUIREMENTS

Considerable reliability problems have been encountered with ignition systems; it is therefore essential that the whole of the pilot/ignition system is correctly designed, operated and maintained.

It is recommended that the ignition system be of the flame front generator type, designed to ignite the pilot burners at the design wind conditions. Separate ignition lines shall be provided for each pilot.

Re-ignition of the pilot shall be performed manually from a safe location at grade from where the flare tip is visible.

It is essential to create a (near) stoichiometric gas/air mixture, e.g by carefully designed restriction orifices and carefully controlled gas and air supply pressures.

For the pilot light supply lines, particular care should be taken when routing the ignition lines to ensure that pocketing is not possible.

The fuel gas and air shall be dried and filtered as for the pilot gas.

The use of an ignition system of the flame front generating type is recommended because of its reliability and because if it fails it can be repaired while the flare relief system remains in service.

11. VENTING ATMOSPHERIC AND LOW PRESSURE STORAGE TANKS

For pressure relief and venting of low-pressure storage tanks, including refrigerated storage tanks, reference should be made to API Std 2000 and DEP 34.51.01.31-Gen. Due to the very low design pressures of the tanks it is not normally permissible to connect them to the flare headers, since the backpressure in this system is generally too high.

12. REFERENCES

In this DEP, reference is made to the following publications:

NOTE: Unless specifically designated by date, the latest edition of each publication shall be used, together with any amendments/supplements/revisions thereto.

Amended per Circular 32/98

SHELL STANDARDS

Index to DEP publications and standard specifications	DEP 00.00.05.05-Gen.
Definition and determination of temperature and pressure levels	DEP 01.00.01.30-Gen.
Preparation of safeguarding memoranda and process safeguarding flow schemes	DEP 01.00.02.12-Gen.
LPG bulk storage installations	DEP 30.06.10.12-Gen.
Metallic materials - prevention of brittle fracture	DEP 30.10.02.31-Gen.
Painting and coating of new equipment	DEP 30.48.00.31-Gen.
Noise control	DEP 31.10.00.31-Gen.
Gas/liquid separators - Type selection and design rules	DEP 31.22.05.11-Gen.
Safety/relief valve calculation sheet (in Requisitioning binder DEP 30.10.01.10-Gen.)	DEP 31.36.90.94-Gen.
Piping - General requirements	DEP 31.38.01.11-Gen.
Pipeline overpressure protection	DEP 31.40.10.14-Gen.
Instrumentation of depressuring systems	DEP 32.45.10.10-Gen.
Classification and implementation of instrumented protective functions	DEP 32.80.10.10-Gen.
Minimum requirements for structural design and engineering	DEP 34.00.01.30-Gen.
Steel stacks (amendments/supplements to CICIND Model Code)	DEP 34.24.26.31-Gen.
Steel structures	DEP 34.28.00.31-Gen.
Standard vertical tanks - selection, design and fabrication	DEP 34.51.01.31-Gen.
Insulating and dense refractory concrete linings	DEP 64.24.32.30-Gen.
Interlocking systems for safety/relief valves	DEP 80.46.30.11-Gen.
“FRED” (hazard consequence analysis package)	OP 97-47088
Standard Drawing: Flare structure - Erection Procedure	S 28.028

AMERICAN STANDARDS

Sizing, Selection and Installation of Pressure - relieving Devices in Refineries	API RP 520
Part I - Sizing and Selection	6th Edition, March 1993
Part II - Installation	4th Edition, Dec. 1994
Guide for pressure-relieving and depressuring systems	API RP 521 4th Edition, 1997
Flanged Steel Safety Relief Valves	API Std 526

Venting Atmospheric and Low Pressure Storage Tanks (Non-refrigerated and Refrigerated)	API Std 2000 4th Edition, Sept. 1992
Requirements for Safe Discharge of Hydrocarbons to Atmosphere (presented at the API Division of Refining, 28th midyear meeting, session on pressure-relieving systems, May 1963)	API Division of Refining Vol 43, III

Issued by:
American Petroleum Institute
Publications and Distribution Section
1220 L Street Northwest
Washington DC. 20005
USA.

Chemical plant and petroleum refinery piping	ASME B31.3
ASME Boiler and Pressure Code	
Section I - Power Boilers	ASME I
ASME Boiler and pressure vessel code: Rules for construction of pressure vessels	ASME VIII, Division 1
Alternative rules for construction of pressure vessels	ASME VIII, Division 2

Issued by:
American Society of Mechanical Engineers
345 East 47th Street
New York, NY 10017, USA.

Standard for the storage and handling of liquefied petroleum gases	NFPA 58

Issued by:
National Fire Protection Association
470 Atlantic Avenue, Boston
Massachusetts, 02210, USA.

BRITISH STANDARDS

Notes on the use of the Ringelmann and miniature smoke charts	BS 2742
Specification for Unfired Fusion-welded Pressure Vessels	BS 5500

Issued by:
British Standards Institution
389 Chiswick High Road
London W4 4AL
United Kingdom.

APPENDIX 1 VAPOUR RELIEF REQUIREMENTS FOR FIRE EXPOSURE

For fire conditions the following calculation procedure shall be used:

1) The boiling temperature of the liquid at the relief pressure is determined and subsequently the latent heat of vaporisation at these conditions.

2) The internally wetted surface of the vessel is calculated. Take the wetted surface up to a height of 8 m above grade or, in the case of spheres or spheroids, at least the elevation of the maximum horizontal diameter or a height of 8 m, whichever is the greater. The term "grade" usually refers to ground level, but may be at any level at which a sizeable pool fire could be sustained (e.g. concrete platform).

   If a variable liquid level lies within the above specified height the calculation of the wetted surface shall be based on following:

   - Storage vessels are assumed to contain their maximum working volume.
   - Process vessels are assumed to contain their normal working volume.
   - Columns (packed and tray) are assumed to contain a liquid volume corresponding to the height from the bottom to the highest level controller connection and the liquid on the trays which can be calculated on the basis of normal pressure drop over the columns.

   Cases not dealt with in the above should be considered individually.

3) If all floors are self draining and any liquid is led away from the equipment which has to be protected, the heat input shall be calculated as follows :

   Q = 43.2 $FA^{0.82}$ (based on formula of API RP 521, Q =21 000 $FA^{0.82}$ Btu/h)

   Q = Total heat input in kW

   A = Wetted surface in m^2

   F = Environmental factor; depends on the presence of an insulation jacket or a protective layer.

4) The following values for F shall be used:

	F
No insulation	1.0
Insulation thickness 25 mm and greater *	0.3
Heating or cooling jacket	0.6
Mounded storage (excl. fire exposed part)	0.03
Underground storage	0.0

* The following fire resistant insulation requirements and alternatives apply:

a) The sheeting shall be made of galvanised steel or stainless steel and be fastened with stainless steel or galvanised self-tapping screws, not with aluminium blind rivets. The combination of sheeting and insulation shall be resistant against prolonged fire exposure and dislodgement by fire hose streams (see API RP 521).

b) Suitable insulation, in descending order of excellence, is:

   1) Shaped foam glass panels. These are heat-resistant, do not soak up oil and will not burn as a wick. They are, however, expensive.

   2) Ceramic fibre blanket at least 20 mm thick (e.g. Fibrefax or Kaowool) with outer sheeting over the regular insulation. This system will soak up oil but will not burn as a wick provided the sheeting remains intact.

   3) Mineral wool insulation throughout, at least 100 mm thick with outer sheeting. The wool, or at least the outer layers, should have a heat-resistant binder. Because the mineral wool cannot withstand the high temperatures that a fire can attain (1000

°C) it will gradually deteriorate to the 25 mm required by the code, if the outer sheeting stays intact. This system will soak up oil but will not burn as a wick provided the sheeting remains intact.

This system is the most common form of "fire-resistant" insulation for vessels.

5) If drainage cannot be provided in the area under the vessel, the heat input shall be calculated as follows:

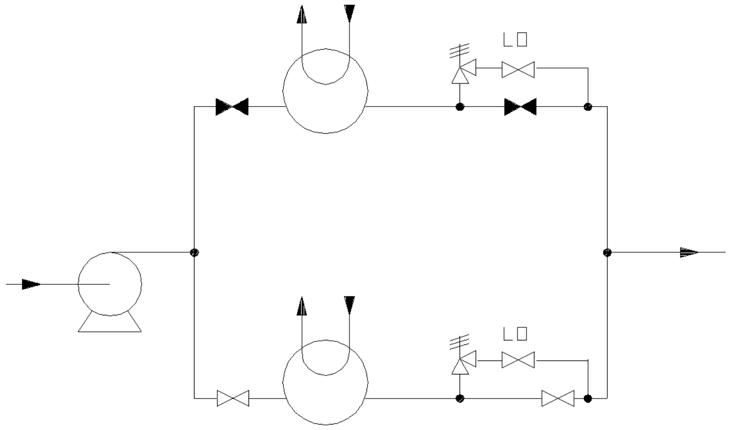
$Q = 71\ FA^{0.82}$ (based on formula of API RP 521, $Q = 34\ 500\ FA^{0.82}$ Btu/h)

6) The rate of discharge in kg/s in the event of fire is:

Total heat input rate (kW) / Latent heat of vapourisation (kJ/kg)

APPENDIX 2 TYPICAL LINE-UPS OF THERMAL EXPANSION RELIEF VALVES

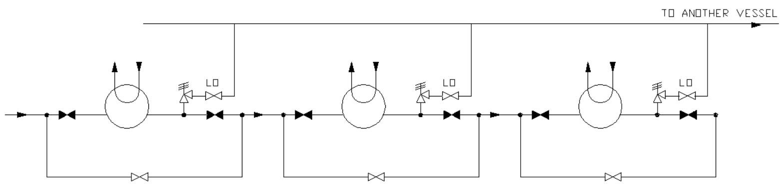
EXAMPLE 1



The thermal expansion relief valves (TERVs) in the above situation can be lined up to discharge in the downstream direction, via the downstream isolation valve.

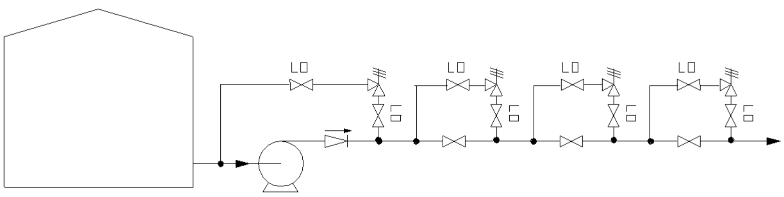
The spring setting (differential set pressure) of the TERVs shall not exceed the maximum allowable operating pressure (design pressure) of the protected equipment minus the maximum operating pressure at the point of discharge of the TERVs. The spring setting of a TERV so lined up can generally be quite low (1 bar) without causing troublesome inadvertent relief, because there is also an upstream isolation valve which is closed when the equipment is isolated. The spring setting of the TERV shall also be higher than the pressure drop between its inlet and outlet tie-in points with the isolation valves open. This pressure drop should normally be quite low, unless the downstream valve is a control valve.

EXAMPLE 2



If the system as proposed in example 1 is not possible the TERVs may relieve into a header, transferring the liquid to another vessel. Depending on the operating conditions (e.g. maximum operating pressure, maximum allowable working pressure and constant (maximum) back pressure) a suitable spring setting of the TERVs shall be selected.

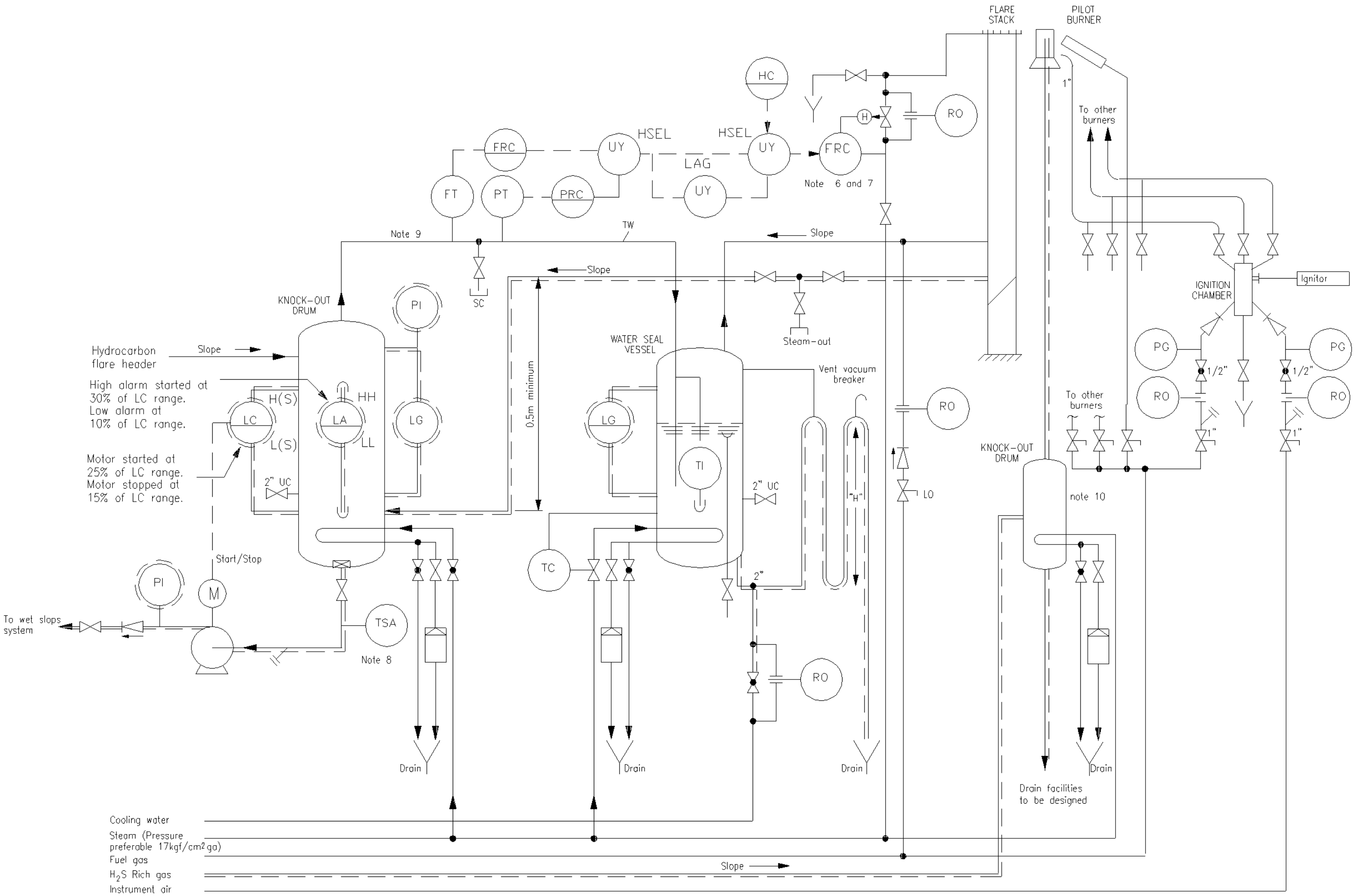
EXAMPLE 3



The above represents a pipeline transfer system of which the various pipeline sections can be blocked in. In this case the TERV nearest to the transfer pump relieves the liquid back to storage. If operation procedures are such that after the system is blocked in the section nearest the pump is partly drained, the relief valve on this section may have a set pressure of 110% of the maximum allowable working pressure. This takes advantage of the fact that a pipeline system can occasionally be overpressured up to 133% or 110% of the maximum allowable working pressure (refer ASME B31.3 and B31.4). This will prevent liquid being recycled when the system is operating close to its maximum allowable working pressure.

The spring setting of the other TERVs may be 1 bar, similar to the settings proposed for example 1.

APPENDIX 3 HYDROCARBON FLARE SYSTEM AND H_2S FLARE SYSTEM



NOTES:

1. The need for steam tracing depends on climatic conditions.

2. The pump capacity shall be such that the liquid hold-up of the knock-out drum can be disposed of within two hours.

3. For LNG plants no water seal vessel and no steam injection in the flare tip is applied.

4. Water seal column height, "H" shall be at least 2 m or P/(gJ), whichever is greater
   where
   P = maximum gauge pressure in the water seal vessel (Pa);
   g = acceleration due to gravity (m/s^2);
   J = density of liquid (kg/m^3).

5. Steam flow depicted for electronic transmission of signals.

6. Operator set maximum steam flow.

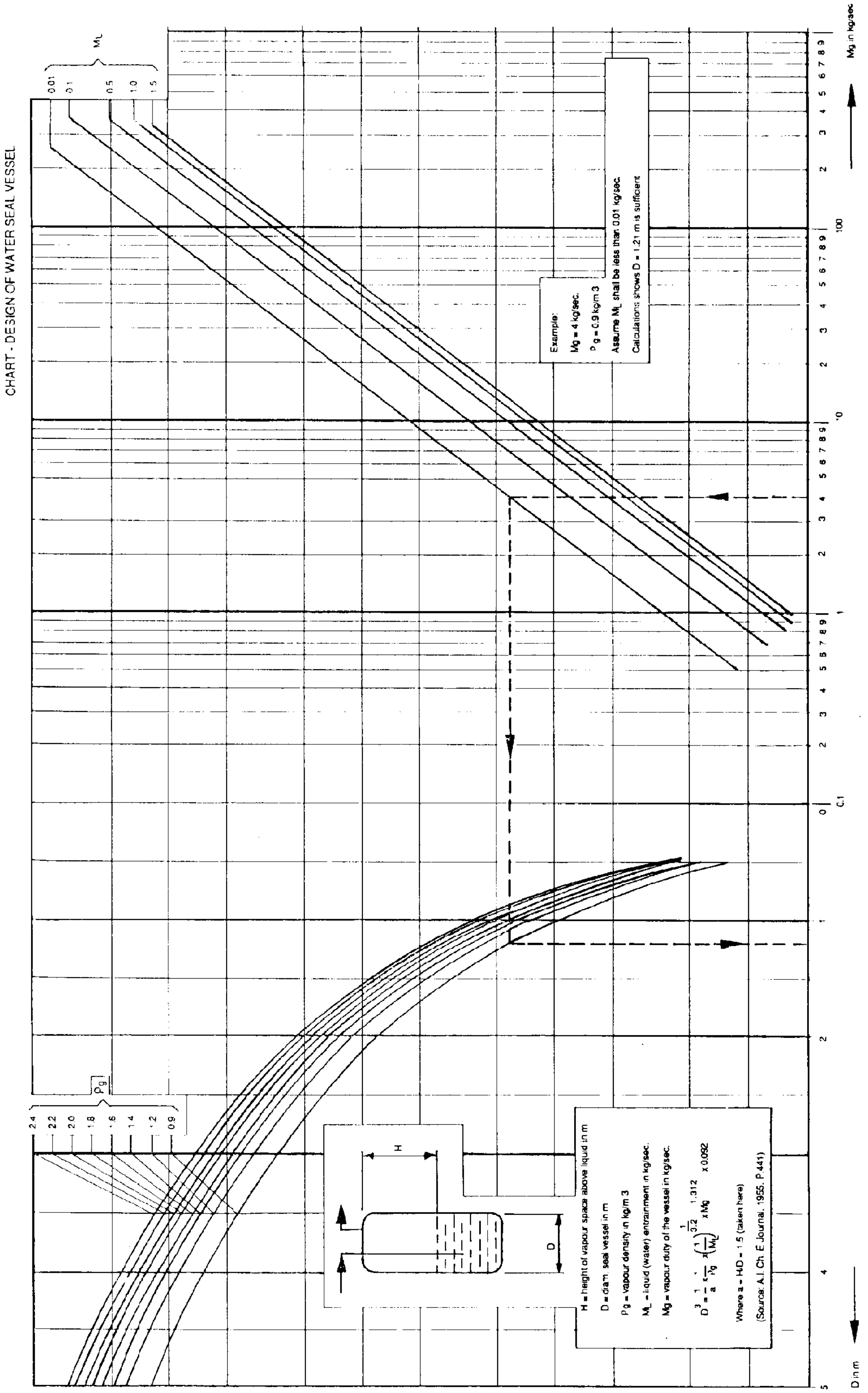
7. Range of required steam flow may necessitate more than one transmitter (auto range selection).

8. TIC is optional but shall be applied when liquid which is too hot or too cold (e.g. LPG) is pumped to slops.

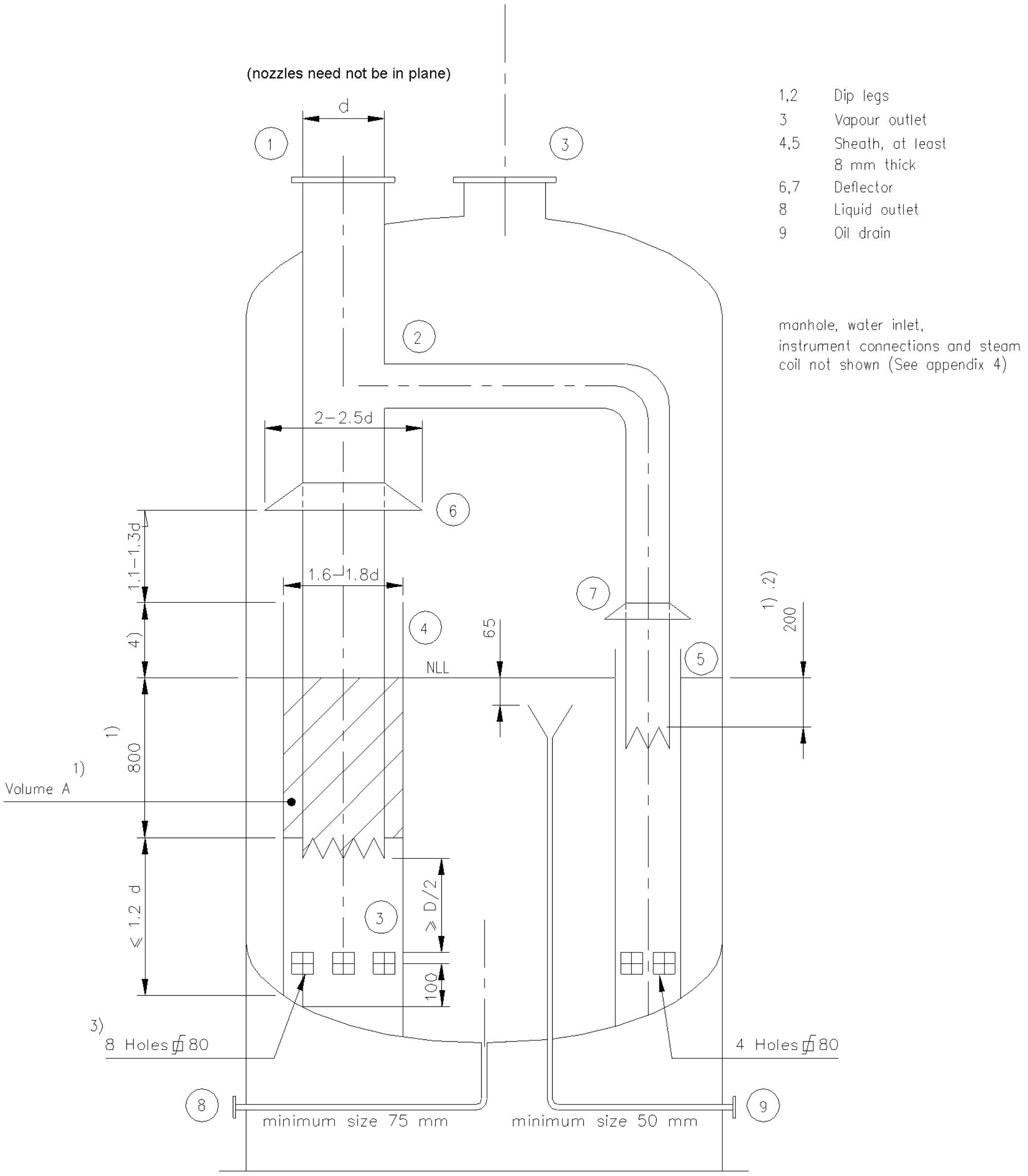
9. Ultrasonic flow meter.

10. A level alarm shall be provided if large quantities of liquid are expected.

APPENDIX 4 WATER SEAL VESSEL DESIGN CHART



APPENDIX 5 TYPICAL DESIGN FEATURES OF WATER SEAL VESSEL



NOTES:

1. Immersion depth may vary for different applications but the small dip leg shall remain immersed when, by operation of the swan neck level control during flaring, volume "A" is drained away and the water level subsequently drops after flaring.

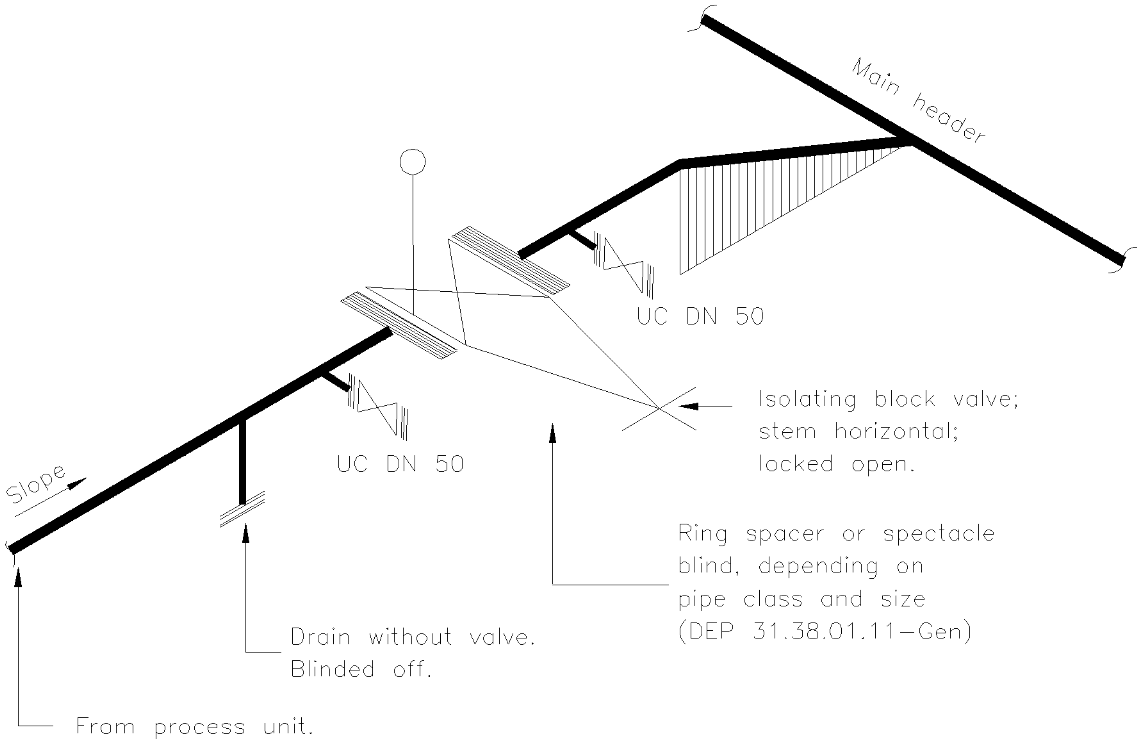
2. Submersion at least such that in the event of reverse pressure, water in vessel can be pushed 3m up the dip legs.

3. For a single operated flare stack/water seal vessel. For parallel operated flare stacks / water seal vessels the total area of inlet holes to be approximately equal to the area of annulus between skirt and dip leg.

4. Height such that annular space can contain at least submerged volume of dip leg.

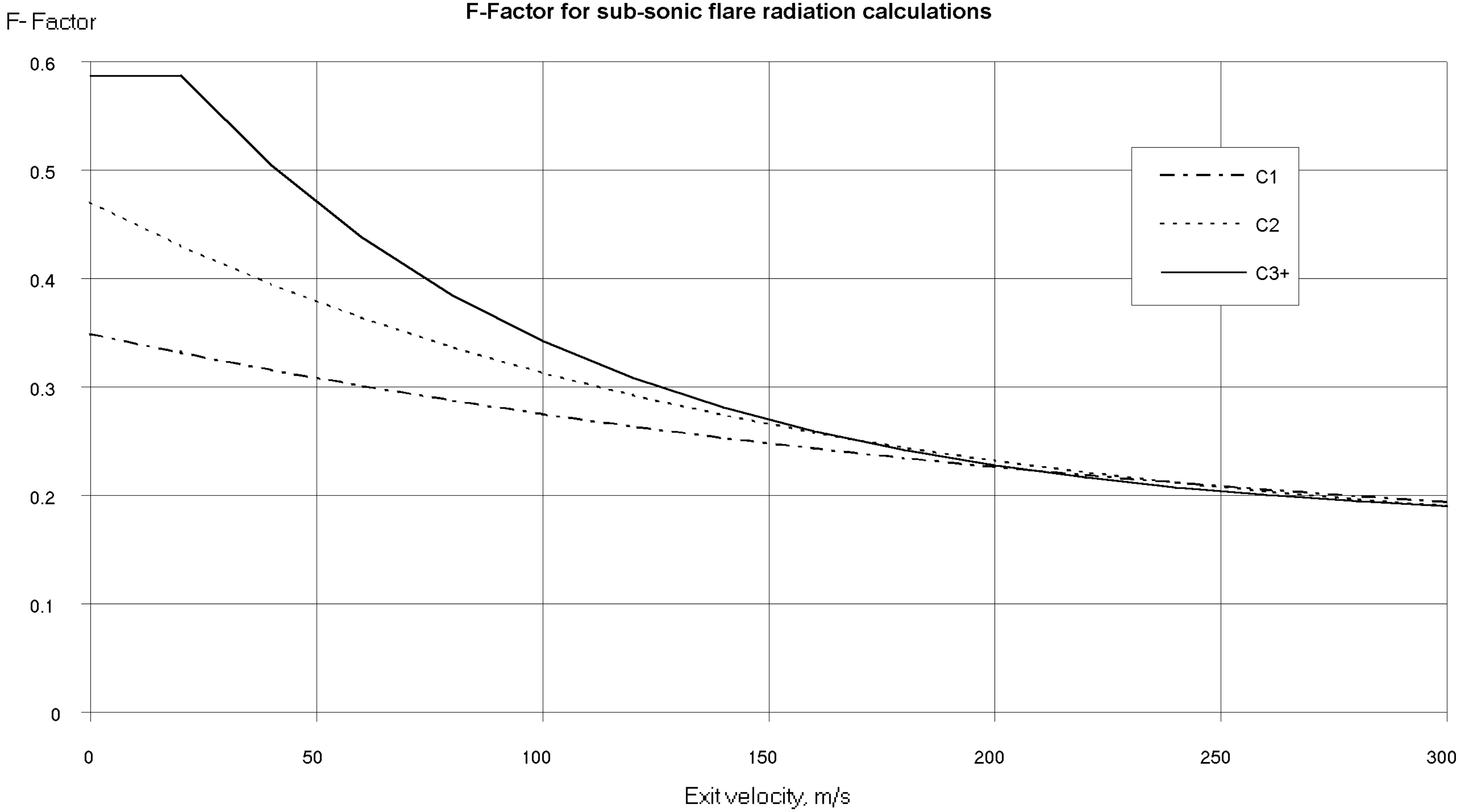
5. Each stack requires its own seal vessel; staggered setting of dip legs of different seal vessels is advised.

APPENDIX 6 ARRANGEMENT OF BLOCK VALVE FOR ISOLATING UNIT



APPENDIX 7 F-FACTORS USED IN THE API MODEL TO DETERMINE RADIATION

F-Factor for sub-sonic flare radiation calculations



APPENDIX 8 ESTIMATE OF STEAM INJECTION REQUIREMENTS FOR FLARING

Waste gas		Formula	Steam/gas weight ratio
Paraffins	Ethane	$C_2 H_6$	0.15
	Propane	$C_3 H_8$	0.25
	Butane	$C_4 H_{10}$	0.30
	Pentane	$C_5 H_{12}$	0.35
	Hexane	$C_6 H_{14}$	0.38
Olefins	Ethylene	$C_2 H_4$	0.40
	Propylene	$C_3 H_6$	0.50
	Butene	$C_4 H_8$	0.58
	Pentene	$C_5 H_{10}$	0.65
Diolefins	Propadiene	$C_3 H_4$	0.70
	Butadiene	$C_4 H_6$	0.90
	Pentadiene	$C_5 H_8$	1.05
Acetylenes	Acetylene	$C_2 H_2$	0.55
Aromatics	Benzene	$C_6 H_6$	0.80
	Toluene	$C_7 H_8$	0.85
	Xylene	$C_8 H_{10}$	0.90

For other components the following equations should be used to calculate the mass flowrate of steam required.

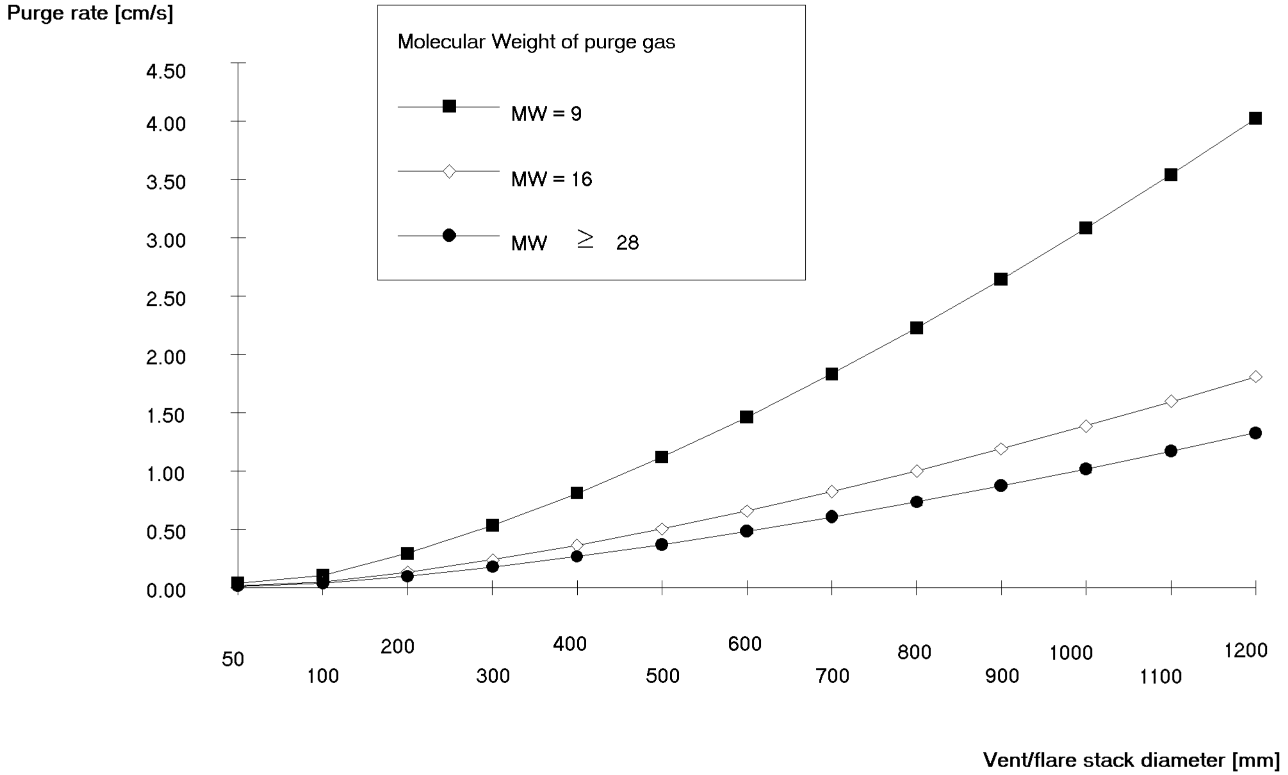
Paraffins

W_{steam} = $W_{paraffin}$ (0.49 - 10.8/MW), where MW = Molecular Weight

Olefins

W_{steam} = W_{olefin} (0.79 - 10.8/MW), where MW = Molecular Weight

APPENDIX 9 PURGE RATES REQUIRED FOR PIPE FLARES



APPENDIX 10 FLARE KNOCK-OUT DRUM DESIGN CONSIDERATIONS

1. LOAD FACTOR VERSUS DROPLET SIZE REMOVAL

In the case of a vertical knock-out drum there is a direct relationship between the gas load factor and the diameter of the smallest droplet which still can be separated from the gas stream, provided there is no maldistribution in the vapour upflow.

This relationship is given in Table II for the two representative flow scenarios presented in Table I.

The physical parameters in bold print determine the relationship between the gas load parameter and droplet diameter.

Table I Representative flow scenarios for flare knock-out drums

	flow scenario 1	flow scenario 2
temperature, °C	244	0
pressure, bara	1.35	1.2
vapour mol. weight	87.2	44
vapour flow rate, kg/s	44.46	114.14
vapour density, kg/m^3	2.78	2.36
vapour viscosity, Pa.s	1.E-5	1.E-5
liquid flow rate, kg/s	1	1
liquid density, kg/m^3	800	600
liquid viscosity, Pa.s	0.001	0.0003

Table II Relationship between gas load factor and diameter of smallest liquid droplet which will still be separated

gas load factor (m/s)	diameter (µm) of smallest droplet which can still be separated	
	flow scenario 1	flow scenario 2

0.02	118	129
0.04	219	237
0.06	324	350
0.07	**379**	**408**
0.08	435	468
0.10	**550**	**590**
0.12	668	716
0.14	790	846
0.16	917	980
0.18	1047	1117

It is seen that both flow scenarios lead to roughly the same relationship.

The droplet diameters associated with a gas load factor of 0.07 and 0.10 m/s are printed in bold.

A similar gas load factor/droplet diameter relationship cannot be given for a horizontal knock-out drum, since in a horizontal vessel the settling process is also determined by the length of the vessel.

In API 521 the vessel sizing method is based on the settling of droplets, assuming a horizontal and uniform vapour flow.

However, in reality the vapour flow is far from uniform, especially if a small feed nozzle with no feed inlet device is used.

Also, in reality, the vapour flow contains a vertical component, in particular in the vicinity of the vapour outlet. Due to these two effects, the API 521 method results in vessels that are too small.

Therefore it is advised to follow the more conservative design method, taking the gas load factor as the criterion. This is also the approach followed in DEP 31.22.05.11-Gen.

However, in the case of flare knock-out drums the gas/liquid separation is not a critical issue, and the sizing rules for these type of separators have been relaxed, allowing high load factors.

Only in special cases where only limited space is available for the knock-out drum, a dedicated Computational Fluid Dynamics (CFD) study may be carried out to arrive at the smallest possible separator, taking vapour flow maldistribution into account. The Principal should be contacted for such a CFD study.

2. TYPICAL LAYOUTS OF FLARE KNOCK-OUT DRUMS

Figure 1 Horizontal knock-out drum with two inlets

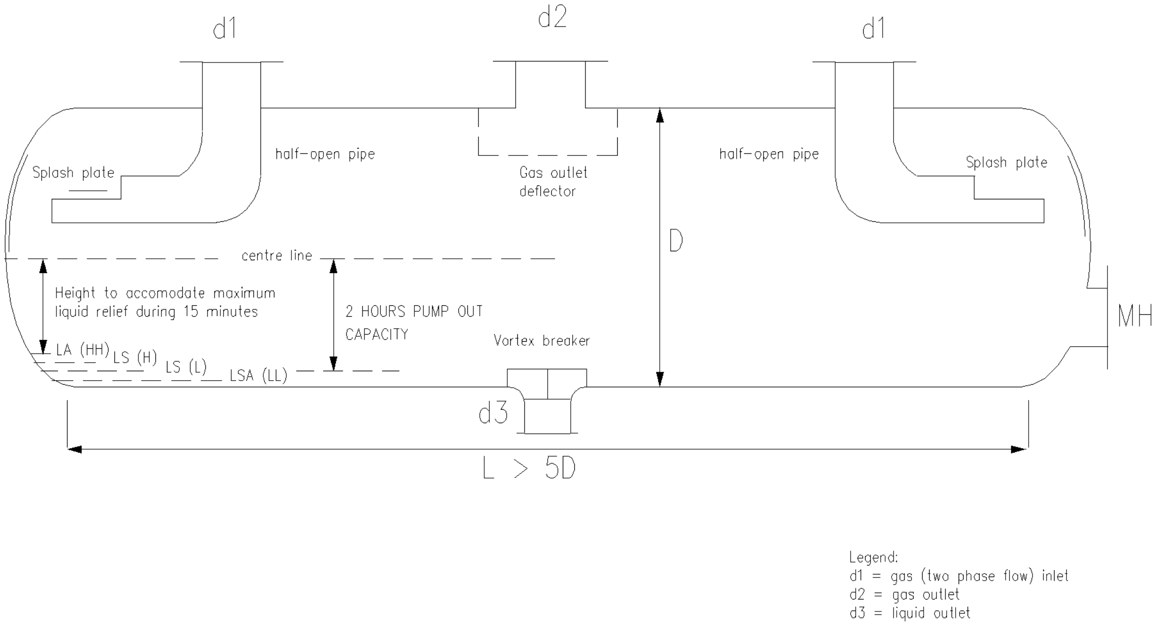


Figure 2 Vertical knock-out drum

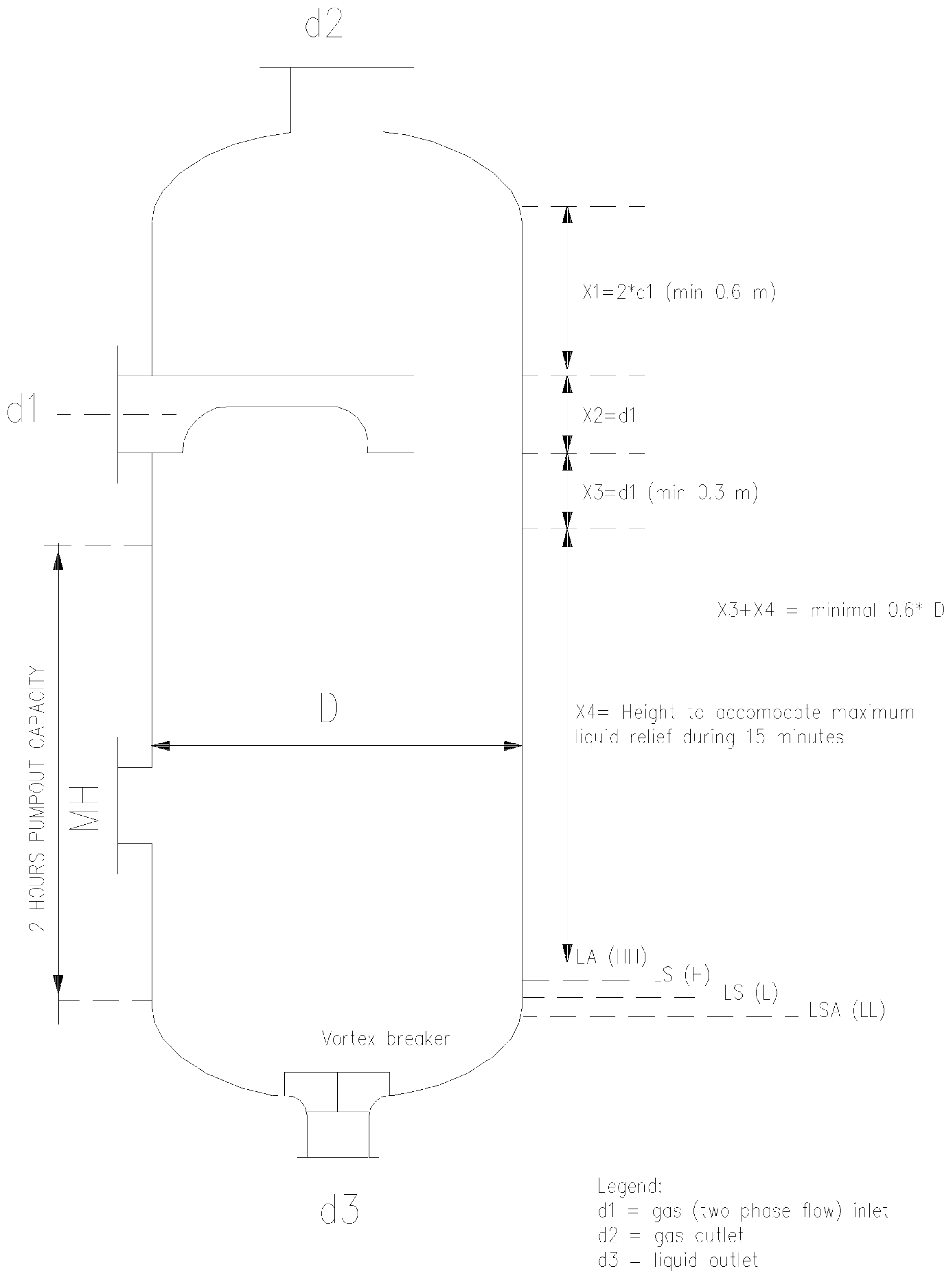
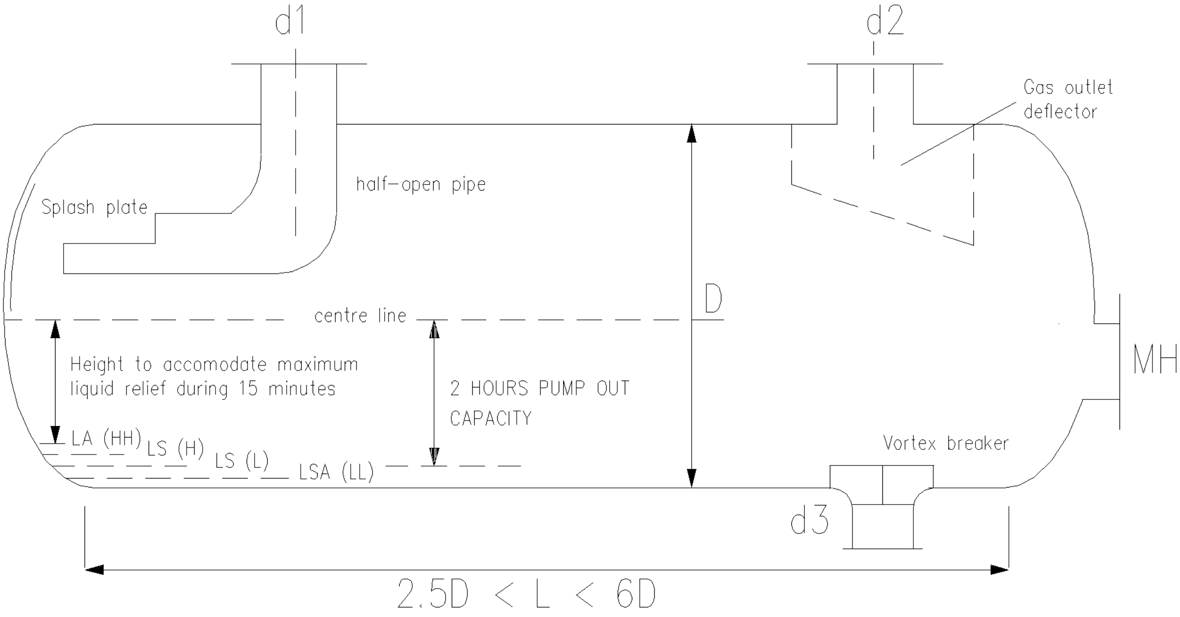

d2
X1=2*d1 (min 0.6 m)
d1
X2=d1
X3=d1 (min 0.3 m)
X3+X4 = minimal 0.6* D
2 HOURS PUMPOUT CAPACITY
D
X4= Height to accomodate maximum
liquid relief during 15 minutes
MH
LA (HH)
LS (H)
LS (L)
LSA (LL)
Vortex breaker
d3
Legend:
d1 = gas (two phase flow) inlet
d2 = gas outlet
d3 = liquid outlet

Figure 3 Horizontal knock-out drum



Legend:
d1 = gas (two phase flow) inlet
d2 = gas outlet
d3 = liquid outlet