DTA PROCESS DESIGN RULES

SAFETY VALVES

SPECIFICATION

Obligatoire

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Ce document remplace les document(s) suivant(s):

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TABLE DES MATIERES

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1. OBJET

Ce document a pour objectif de déterminer la position d'implantation des soupapes , leur dimensionnement et leur contraintes d'installation (collectage , isolation , etc …) pour les ensembles conçus et réalisés par Air Liquide DTA .

Il participe à la maîtrise des risques sur nos installations : « la sécurité d'abord » , à prendre en compte TOUS les risques sans se limiter au risque de perte de vide , à homogénéiser la philosophie de protection sur les affaires DTA , à être homogène par rapport aux pratiques du groupe Air Liquide .

Pour ce dernier point , il est issu du document de la DI [1] dans lequel ont été introduites les particularités des affaires DTA .

2. MOTS – CLES

PROCESS – DESIGN – RULES – SAFETY VALVES – SPECIFICATION

PROCESS – CONCEPTION – REGLES - SOUPAPES - SPECIFICATION

3. CHAMP D'APPLICATION

Ce document s'applique pour les installations conçues et réalisées par Air Liquide DTA , avec une spécificité particulière pour les affaires CRYG / Unités de Réfrigération et Liquéfaction He .

4. DEFINITIONS

Voir le paragraphe 7 ci-dessous .

5. REFERENCES ET DOCUMENTS ASSOCIES

Les documents listés ci-dessous doivent être utilisés dans leur version la plus récente, sauf avis contraire lorsque sont mentionnées le numéro de l'édition et/ou la date de parution .

5.1. INTERNATIONAL STANDARDS

[1]	API Recommended Practice 520 Sixth edition, March 1993 Fourth edition, December 1994	Sizing, Selection and Installation of Pressure Relieving Devices in Refineries Part I – "Sizing and Selection" . Part II – "Installation"
[2]	API Recommended Practice 521 Fourth edition, March 1994	Guide for Pressure Releving and Depressuring System

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[3]	API standard 526 Third edition February 1984	Flanged Steel Safety Relief Valves
[4]		
[5]	Norme Française NF E 29-410 September 1984	Soupapes de sureté, définition des termes techniques
[6]	Norme Française NF E 29-411 September 1984	Soupapes de sureté, conception générale, calcul du débit, essais, marquage, conditionnement
[7]	CGA S-1.3-2005 Seventh Edition	PRESSURE RELIEF DEVICE STANDARDS PART 3- STATIONNARY STORAGE CONTAINERS FOR COMPRESSED GASES
[8]	Norme européenne NF EN 13648-3 Dec. 2002	Récipients cryogéniques – Dispositifs de protection contre les surpressions Part 3 : Détermination du débit à évacuer - Capacité et dimensionnement

5.2. AIR LIQUIDE DOCUMENTATION

[9]	CP.990.20 30/05/1972	Note de calcul : soupapes de sécurité B. Vandenbussche,
[10]	CP.990.23 17/11/1976	Proposition de règles pour la spécification des soupapes de sécurité d'un oxytonne B. Vandenbussche,
[11]	CP.990.22 20/01/1977	Tuyauterie de liaison équipement-soupape basse pression B. Vandenbussche,
[12]	CP.990.21 09/06/1977	Règles AL pour la définition des soupapes de sécurité Oxytonne simple cycle B. Vandenbussche,
[13]	Standard MC 44850.0 18/02/1980	Dimensionnement des soupapes de sûreté à ressort
[14]	Note DCVM/DI/CP19/RV/bg 1 26/03/1990	Dimensionnement des soupapes des appareils air R. Vancauwenberghe,

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[15]	Note DI/DP/RV 14/10/1991	Compte rendu de la réunion du sous groupe de travail soupapes colonnes BP du 11/10/91 R. Vancauwenberghe,
[16]	Note DI/DP/RV 06/11/1991	Groupe de travail soupapes : compte rendu de la réunion du groupe de travail du 29/10/91 R. Vancauwenberghe,
[17]	CP.990.24-0 12/03/1992	Dimensionnement des principales soupapes R. Vancauwenberghe,
[18]	RG.286.01-a 13/07/1995	Etude de risques, définition des organes de sécurité R. Mathé,
[19]	Note ICI\tsvfire 14/12/1995	Soupapes au feu boite froide E. Michalik,
[20]	RG.257.51-0 10/06/98	Conception des réfrigérants et des circuits d'eau appareils H2CO L. Greter,
[21]	Note DI/CCT/RV 20/08/1998	Dimensionnement soupapes R. Vancauwenberghe,
[22]	RG.104.10-f 20/09/1998	Pressions et températures : définitions et déterminations L. Greter,
[23]	Meeting report 01/03/99	Définition de règles de conception et de standardisation des surpresseurs d'air L. Greter,
[24]	Meeting report 12/03/99	Surpresseurs d'air, turbine booster L. Greter,
[25]	Meeting report 02/04/99	Surpresseurs d'air L. Greter,
[26]	RTS B.02.01-b 08/07/1999	Réservoirs cryogéniques basse pression, protection contre une surpression ou une dépression P. Fournier,
[27]	Standard 331.12.GS-a September 1999	Water coolers for gas under pressure, general specification Ph. Fert,
[28]	Meeting report 24/03/2000	H2CO cold box module safety relief valves J. Leroy/N. Haik,
[29]	ALPC Engineering practice EP 57.01 Rev 0 01/08/1997	Relief device selection T. Kundel/D. Tanguay,
[30]	Memo AB002 19/05/2004	Conventional, balanced and piloted safety valves A. Briglia/N. Schmitt,

[31]	ICEC 7, London, 1978	Safety aspects for Lhe Cryostats and Lhe Transport Containers , W. Lehman and G. Zahn
[32]	DTA/FT 011	Détermination des pressions clés d'un appareil selon DEP 97/23/CE \\fr-s-dta-03\P-InfoPerm$\Etudes\STANDARD\FichesTechniques\Nouveau dossier\FT011.doc
[33]	DI / S05 / A 0030 – (b)	PROCESS DESIGN RULES – SAFETY VALVES – Specification

6. INTRODUCTION

The scope of the present note is to provide the criteria, procedures and formulas for sizing pressure relief devices for our units. Sources of overpressure are described in a general way, after which particular cases for helium refrigeration / liquefaction units are developed.

Mentioned cases are not exhaustive and should be regarded as examples. One should carefully consider particular configurations, being inspired by given cases.

This document is mainly based on API RP 520 and API RP 521 and describes the rules adopted by Air Liquide. Local regulation or Client requirements shall be applied if more constraining.

7. MAIN DEFINITIONS

The following terms are encountered when relating to pressure relief devices. Definitions listed below per API 520. For more detailed devices descriptions refer to API 520 or vendor catalogues. French translation of terms is issued from French Norm NF E 29-410.

7.1. PRESSURE RELIEF DEVICES

7.1.1. PRESSURE RELIEF DEVICE (SOUPAPES)

It is actuated by inlet static pressure and designed to open during an emergency or abnormal conditions to prevent a rise of internal pressure in excess of a specified value. The device also may be designed to prevent excessive vacuum.

7.1.2. SPRING LOADED PRESSURE RELIEF VALVES (SOUPAPES À RESSORT)

It is a pressure relief device designed to automatically reclose and prevent the further flow of fluid.

7.1.3. SAFETY VALVES (SOUPAPES DE SURETE)

A safety valve is a spring loaded pressure relief valve actuated by upstream static pressure and is characterized by rapid opening or pop action. These valves provide full opening with a minimum over-pressure. They are normally used with compressible fluids.

7.1.4. RELIEF VALVE (SOUPAPE DE DECHARGE)

A relief valve is a spring loaded pressure relief valve actuated by inlet static pressure which opens in proportion to the pressure increase over the opening pressure. These valves are primarily selected for liquid (incompressible fluids) service.

7.1.5. SAFETY RELIEF VALVE

A safety relief valve is a spring loaded pressure relief valve that may be used as either a safety or relief valve depending on application.

7.1.6. CONVENTIONAL PRESSURE RELIEF VALVE (SOUPAPE CONVENTIONNELLE)

It is a spring loaded relief valve whose performance characteristics are directly affected by changes in the back-pressure on the valve. They should typically not be used when the build up back pressure is greater than 10 percent of gauge set pressure.

7.1.7. BALANCED PRESSURE OR BALANCED BELLOWS PRESSURE RELIEF VALVE (SOUPAPE ÉQUILIBRÉE OU À SOUFFLET)

It is a spring loaded relief valve that incorporates a means for minimizing the effect of back-pressure on the valve performance characteristics.

7.1.8. PILOT OPERATED RELIEF VALVES (SOUPAPE PILOTÉE)

These are pressure relief valves in which the main relieving device is combined with and controlled by a auxiliary pressure relief valve. The useful nature of the pilot valve lies in the fact that these valves will open fully at the set pressure (not slowly begin to open as the set pressure is approached) allowing tighter tolerances between operating and set pressure conditions.

Pilot valves should be avoided when the process service is dirty or corrosive. It is acceptable to use specialized pilot valves in cryogenic service where required by operating conditions.

7.1.9. RUPTURE DISK DEVICES (DISQUES DE RUPTURE)

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It is a non-reclosing differential pressure relief device actuated by inlet static pressure and designed to function by bursting of the pressure containing disk. A rupture disk is the pressure containing and pressure sensitive element of a rupture disk device. A rupture disk holder is the structure which encloses and clamps the rupture disk in position.

7.2. PRESSURE RELIEF DEVICE DIMENSIONAL TERMINOLOGY

7.2.1. ACTUAL DISCHARGE AREA (SECTION D'ÉCOULEMENT RÉELLE)

This is the measured minimum net area that determines the flow through the valve.

7.2.2. EFFECTIVE DISCHARGE AREA OR EQUIVALENT FLOW AREA (SECTION D'ÉCOULEMENT EFFECTIVE)

A nominal or computed area of a pressure relief valve used in recognized flow formulas to determine the size of the valve. It will be less than the actual discharge area.

7.2.3. HUDDLING CHAMBER

An annular pressure chamber in a pressure relief valve located beyond the seat for the purpose of generating a rapid opening.

7.2.4. INLET SIZE

The nominal pipe size (NPS) of the valve at the inlet connection, unless otherwise designated.

7.2.5. OUTLET SIZE

The nominal pipe size (NPS) of the valve at the discharge connection, unless otherwise designated.

7.2.6. LIFT (LEVÉE)

The actual travel of the disk away from the closed position when the valve is relieving.

7.3. OPERATIONAL TERMINOLOGY

For definition of these terms, refer also to Air Liquide Standard E-EP 14-0-1. One should be careful with definition of these terms which are not consistent between different regulation (ASME, AFNOR, European Regulation…)

7.3.1. MAXIMUM WORKING PRESSURE (PRESSION MAXIMALE DE FONCTIONNEMENT)

This is the maximum pressure expected during system normal operations (including start-up, maintenance…).

7.3.2. MAXIMUM ALLOWABLE WORKING PRESSURE : MAWP (PRESSION MAXIMALE ADMISSIBLE)

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This is the maximum gauge pressure permissible at the top of a complete vessel in its operating position (for a designated temperature). The pressure is based on calculations for each element in a vessel (each head, shell, etc.), using nominal thickness, excluding additional metal thickness allowed for corrosion or loads other than design pressure loads. The maximum allowable working pressure is the basis for the pressure setting of the pressure relief devices intended to protect any given vessel.

7.3.3. DESIGN PRESSURE (PRESSION DE CALCUL [1])

This refers to at least the most severe coincident pressure and temperature conditions expected during operation (plus a margin depending of applicable standard). This pressure may be used in place of the maximum allowable working pressure (MAWP) in all cases where the MAWP has not been established. For equipment installed in vacuum chambers, refer to doc **Erreur ! Source du renvoi introuvable.** to calculate MAWP from design pressure (see example in Appendix 1)

7.3.4. ACCUMULATION (ACCUMULATION)

This is the pressure increase over the maximum allowable working pressure of the vessel during discharge through a pressure relief device. Accumulation is expressed in terms of pressure units or a percentage value of an established pressure value. Maximum allowable accumulations are established by applicable codes for operating and fire contingencies. Generally, following values issued from API can be applied [2] :

Table 1 : Set pressure and accumulation limits for pressure relief valves (in percentage of MAWP)

	Single-Valve Installation		Multiple-Valve Installation	
Contingency	Set pressure (percent)	Maximum accumulated Pressure (percent)	Set pressure (percent)	Maximum accumulated Pressure (percent)
Non fire only				
First valve	100	110	100	116
Additional valve(s)	_	_	105	116
Fire only				
First valve	100	121	100	121
Additional valve(s)	_	_	105	121

7.3.5. OVER-PRESSURE (SURPRESSION)

This is the pressure increase over the set pressure of the relieving device, expressed in pressure units or as a percentage of an established pressure value. The Over-pressure value is the same as the

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[1] The term “pression de calcul” does not appear in the new European Regulation
[2] Local regulations (ASME, DEP…) do not mention applicable overpressure for fire contingency.

accumulation value when the set pressure is equal to the maximum allowable working pressure (MAWP) of a vessel.

7.3.6. SCENARIO

A process upset condition which causes a relief valve to lift (or rupture disk to burst).

7.3.7. CIRCUIT

Pieces of equipment linked together by piping which have the same design pressure.

7.3.8. SYSTEM

A group of equipment protected by one relief device.

7.4. DEVICE TERMINOLOGY

7.4.1. SET PRESSURE (PRESSION DE DEBUT D'OUVERTURE [3])

This is the inlet gauge pressure at which the pressure relief device is set to open (burst pressure for rupture disks) under service conditions.

7.4.2. COLD DIFFERENTIAL TEST PRESSURE (PRESSION DE RÉGLAGE)

This is the pressure at which the pressure relief valve is adjusted to open on the test stand. The cold differential test pressure includes corrections for the service conditions of back pressure, temperature, or both.

7.4.3. BACK PRESSURE (CONTRE PRESSION)

This is the pressure that exists at the outlet of a pressure relief device as a result of the pressure in the discharge system. This is the sum of the superimposed and built-up back pressures (see next).

7.4.4. SUPERIMPOSED BACK PRESSURE (CONTRE PRESSION INITIALE)

This is the static pressure that exists at the outlet of a pressure relief device at the time the device is required to operate. It is a result of pressure in the discharge system coming from other sources in the same system. This value may be a constant or it may vary.

7.4.5. BUILT-UP BACK PRESSURE (CONTRE PRESSION ENGENDRÉE)

This is the increase in pressure in the discharge header that develops as a result of flow after the relief device opens.

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[3] the former term "pression de tarage" should not be used any more

7.4.6. BLOWDOWN (CHUTE DE PRESSION A LA REFERMETURE)

This is the difference between the set pressure and the closing pressure of a pressure relief valve, expressed as a percentage of the set pressure, or in pressure units.

7.4.7. OPENING PRESSURE (PRESSION D'OUVERTURE)

This is the value of increasing static pressure at which there is a measurable lift of the disk, or at which discharge of the fluid becomes continuous.

7.4.8. CLOSING PRESSURE (PRESSION DE REFERMETURE)

This is the value of decreasing inlet static pressures at which the valve disk re-establishes contact with the seat or at which lift becomes zero.

7.4.9. SIMMER (FUITE)

This is the audible or visible escape of compressible fluid between the seat and disk at inlet static pressure below the set pressure (no measurable capacity)[4].

7.4.10. LEAK-TEST PRESSURE (PRESSION D'ÉTANCHÉITÉ)

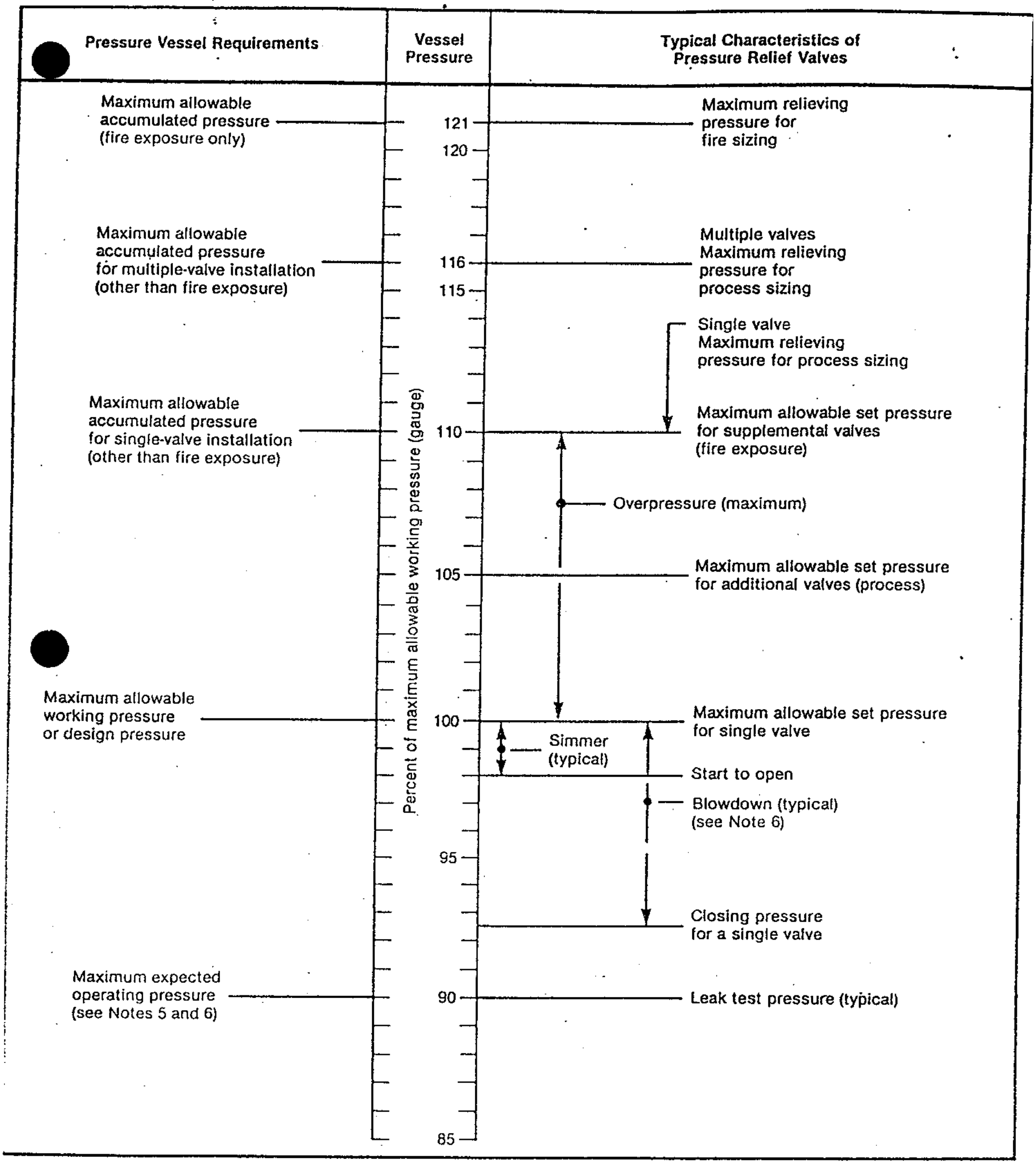
This is the specified inlet static pressure at which a seat leak test is performed.

7.4.11. RELIEVING CONDITIONS (CONDITIONS DE DECHARGE)

This term is used to indicate the inlet pressure and temperature of a pressure relief device at a specific over-pressure. The relieving pressure is equal to the valve set pressure (or rupture disk burst pressure) plus the over-pressure. The temperature of the flowing fluid at relieving conditions may be higher or lower than the operating temperature.

[4] Fabrication tolerance is equal to 10% of set pressure for conventional PSV, 5% for pilot operated PSV. Tolerance is higher for liquid PSV

Example of pressures relationships, valid for gaseous PSV with above definitions :



lotes:

. This figure conforms with the requirements of Section VIII of the \SME Boiler and Pressure Vessel Code.

. T[illegible]essure conditions shown are for pressure relief valves in-talle[illegible] a pressure vessel.

. Allowable set-pressure tolerances will be in accordance with the pplicable codes.

4. The maximum allowable working pressure is equal to or greater than the design pressure for a coincident design temperature.

5. The operating pressure may be higher or lower than 90.

6. Section VIII, Division 1, Appendix M, of the ASME Code should be referred to for guidance on blowdown and pressure differentials.

Figure 1—Pressure-Level Relationships for Pressure Relief Valves

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8. SELECTION PROCEDURE

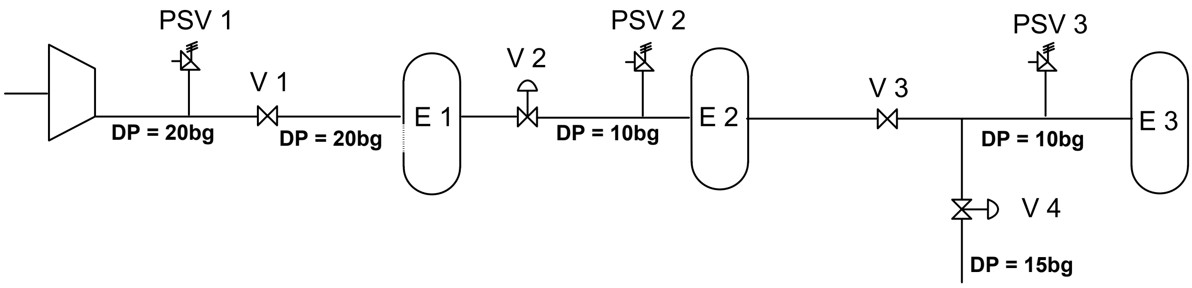
To complete a successful over-pressure protection of the plant, the following procedure shall be followed. All the steps are not necessary in the scope of the Process Engineer but it is a good practice that he/she checks the different points.

8.1. IDENTIFICATION OF PRESSURE CIRCUITS AND SET PRESSURE

The process begins with identification of pressure sources. Start with a pressure source such as main compressor and follow the lines downstream until a sudden pressure change across a piece of equipment (typically a turbine or a control valve) is encountered. Check the design pressure of the equipment downstream of each block valve encountered. If these are similar to the upstream design pressure then downstream circuit belongs to the same pressure circuit.

Each identified system shall be protected by a relieving device, located upstream the first block valve.

The example below shows the reasoning used to determine whether a safety valve is required on each circuit. It doesn't consider the risk of overpressure induced by fire or thermal expansion.



Compressor discharge is protected by PSV 1.

- Downstream block valve V 1, design pressure remains the same as upstream. If V 1 is open, equipment E 1 is protected through PSV 1. If V 1 is closed, no source can overpressure E 1. No additional safety valve is required on E 1.

- Downstream control valve V 2, design pressure is lower than upstream. PSV 2 is required to protect E 2 against overpressure from upstream circuit.

- Downstream block valve V 3, design pressure remains the same as upstream. No overpressure can come from E 2 circuit. Overpressure can however come from circuit upstream control valve V 4. PSV 3 is therefore required to protect E 3.

If fire is to be considered, a safety valve has to be added on each piece of equipment that can be blocked in and possibly submitted to fire (see § 9.9).

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If thermal expansion is to be considered, a safety valve has to be added on each circuit portion that can be blocked in, including small pipe sections (see § 10.1).

Thermal expansion has always to be taken into account in the following cases:

- in systems operating at temperatures below ambient (ex: cold boxes)
- when temperature difference between night and day is significant.

Set pressure of each system is generally taken as design pressure excepted :

- in case of high differential pressure developed between the source of pressure and the relieving device ,
- in case of static head due to liquid height (for example the set pressure of a relief device on top of a distillation column must be corrected for pressure drop through packing and maximum liquid height in column sump),
- in case of an equipment installed in a vacuum shell. In that case MAWP(barg) = DP(bar) -1.013 – ρgH

8.2. IDENTIFICATION OF PRESSURE RELIEVING SCENARIO

Each of the contingencies described in § 9 should be considered when analyzing the relieving requirements of any given system. Some may be eliminated as they may not apply to specific case. The contingency which requires the largest pressure relieving area (usually the highest relieving flow but it is not always the case) shall be used as the design case for the system relieving requirements.

8.3. SELECTION OF RELIEF DEVICE

8.3.1. TYPE SELECTION[5]

- Safety valve with discharge to atmosphere : a conventional safety valve has normally to be chosen. In the case where it is not possible to design outlet line to have a build up back pressure lower than 10% of set pressure (see § 8.4.2) then a balanced PSV can be chosen.
- Safety valve connected to a flare network : a balanced safety valve has to be selected if

set pressure - (SBP_{max} - SBP_{min}) < 1.1 maximum working pressure

[5] all pressure in this section are gauge.

otherwise, a conventional safety valve can be chosen, excepted if it is not possible to have a build up back pressure less than 10% of set pressure (see § 8.4.2)

SBP_{max} = Superimposed Maximum Back Pressure. It corresponds to the maximum value due to the release of other sources.

SBP_{min} = Superimposed Minimum Back Pressure. It corresponds to the minimum value generally due to the absence of release sources.

- Piloted safety valves can be used where set pressure cannot be higher than 110% of maximum working pressure

8.3.2. SIZE SELECTION

The size of the required safety valve can be calculated using following formulas issued from API 520 (calculation per ASME VIII div 1 or ISO 4126 uses same formulas with different coefficients). The purpose of this process is not to tell the vendor the specific orifice size required but rather to check the independent calculations generated by the vendors. If the two sizes match, potential mistakes may be averted.

- For a gas with $r \le \left(\frac{2}{k+1}\right)^{\frac{k}{k-1}}$ (critical flow) or for a balanced-bellows valve :

Eq. 1 $$A = 1.316 \frac{W}{C K_d P_1 K_b} \sqrt{\frac{TZ}{M}}$$

- For a gas with $r > \left(\frac{2}{k+1}\right)^{\frac{k}{k-1}}$ (subcritical flow). To be used only for conventional and pilot operated relief valves :

Eq. 2 $$A = \frac{W}{558.4 F_2 K_d} \sqrt{\frac{ZT}{M P_1 (P_1 - P_2)}}$$

- For steam :

Eq. 3 $$A = \frac{W}{52.5 P_1 K_d K_N K_{SH}}$$

- For a liquid :

Eq. 4 $$A = \frac{W}{161.1 K_d K_w K_v}\sqrt{\frac{1}{\rho(P_1 - P_2)}}$$

- For a two phase flow (pressure relief valve handling a liquid at equilibrium or mixed phase fluid will produce flashing). The amount of flashing liquid should be determined by isenthalpic expansion from relieving conditions to the back pressure or critical downstream pressure for the flashed vapor, whichever is greater. The two individual areas calculated for the flashed vapor and for the unflashed liquid are to be added[6].

Where :

-P_1 = upstream relieving pressure in bar abs. This is the set pressure plus the allowable overpressure (see § 7.3.5) plus atmospheric pressure.

-P_2 = back pressure in bar abs.

-r = P_2/P_1.

-k = C_p/C_v, ratio of specific heats at relieving conditions.

-A = required effective discharge area of the valve in cm^2.

-W = required flow through the valve in kg/h.

-C = coefficient determined from ratio of specific heats at standard conditions.

Eq. 5 $$C = 520\sqrt{k_{st}\left(\frac{2}{k_{st}+1}\right)^{\frac{k_{st}+1}{k_{st}-1}}}$$ where k_{st} is the ratio of specific heats at standard conditions

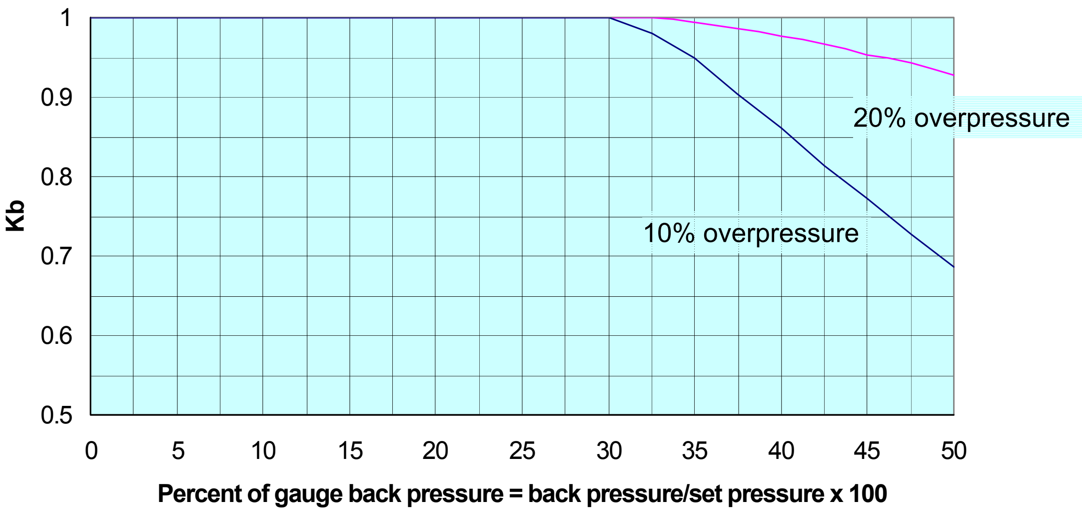
-K_d = effective coefficient of discharge. Use 0.975 (0.875 if calculation per ASME or ISO) for gases and steam. For liquid, should be obtained from manufacturer (0.62 can be used for estimation).

-K_b = capacity correction factor due to back pressure. For conventional safety valves, K_b = 1. For balanced-bellows valve, this can be estimated from

Figure 1(valid only for critical flow and for set pressure above 3.4 barg) or obtained from manufacturer.

[6] since gas is not present at inlet, area for flashed vapor is calculated using Z and M at discharge conditions.

Figure 1 : Back pressure sizing factor K_b for balanced-bellows pressure relief valves



-T = relieving temperature of the inlet gas or vapor in °K.

-Z = compressibility factor for the deviation of the actual gas from a perfect gas, evaluated at relieving conditions. If Z is unknown, assume Z=1.

-M = molecular weight of the gas or vapor in g/mol.

-F_2 = coefficient of subcritical flow.

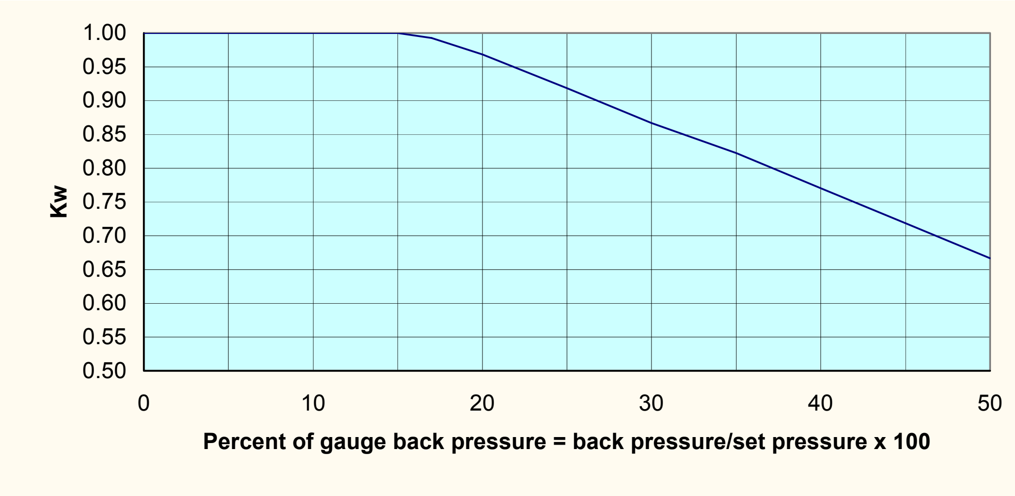
Eq. 6
$$F_2 = \sqrt{\left(\frac{k}{k-1}\right) r^{\frac{2}{k}} \left[\frac{1-r^{\frac{(k-1)}{k}}}{1-r}\right]}$$

-K_N = correction factor for Napier equation. $K_N = 1$ for $P_1 \leq 104.5$ bar abs. For $104.5 < P_1 \leq 222$ bar abs then $K_N = (2.7637P_1\text{-}1000)/(3.3234P_1\text{-}1061)$.

-K_{SH} = superheat steam correction factor. $K_{SH} = 1$ for saturated steam. For superheated steam refer to table 10 in API 520 or ask vendor.

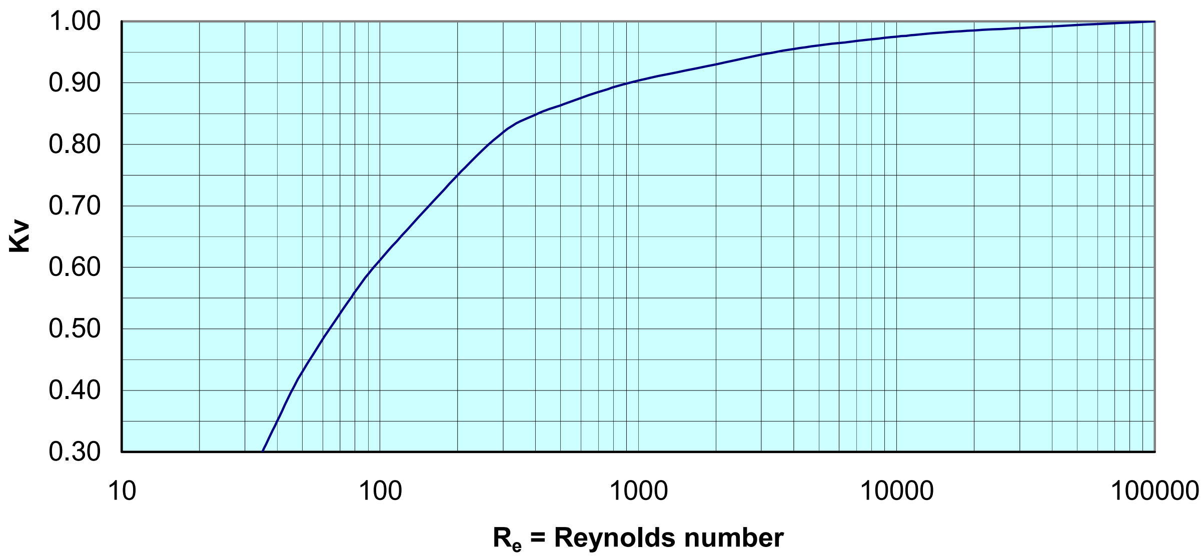
-K_w = correction factor due to back pressure. If back pressure is atmospheric, K_w=1. Balanced bellows valves in back pressure service will require the correction factor determined in. Conventional valves require no correction.

Figure 2 : Correction factor K_w due to back pressure on balanced pressure relief valves in liquid service



-K_v = correction factor due to viscosity. When a relief valve is sized for viscous liquid, it should be first sized as it was for non viscous-type application so that a preliminary required discharge area can be obtained. From manufacturers' standard orifice sizes, the next larger orifice should be used in determining the Reynolds number, R_e, which is used to determine K_v from Figure 3. K_v is then applied to correct the preliminary required area. If the corrected area exceeds the chosen standard size, the above calculation should be repeated using the next larger standard orifice size.

Figure 3 : capacity correction factor K_v due to viscosity



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The selected safety valve will be the next largest nominal orifice size to the orifice calculated (can slightly change following manufacturers catalogue). However, each size is submitted to pressure and temperature limitations. It is sometimes necessary to install two or more smaller sizes rather than a big one (see manufacturer catalogues).

8.4. SELECTION OF INLET AND OUTLET LINES

The safety valve end connections depends on selected size and on inlet/outlet pressure. They can be found on vendor catalogues. Inlet and outlet lines shall not be smaller than flange connections end.

8.4.1. INLET LINE

Excessive pressure loss at the inlet of a pressure relief valve will cause extremely rapid opening and closing of the valve, which is known as “chattering”. Chattering may result in lowered capacity and damage to the seating surfaces.

Design criteria for PSV inlet line are :

- The pressure drop between the protected equipment and the pressure relief valve should not exceed 3% of the valve gauge set pressure. (not applicable if set pressure is less than 1 bar g).
- Mach number[7] should not be higher than 0.3
- In case of Oxygen service, rules for maximum velocity are applicable (refer to Process Design rule A0017)

Velocity and pressure drop should be calculated using the maximum rated flow through the pressure relief valve[8]

8.4.2. OUTLET LINE AND FLARE HEADER

Excessive back-pressure may also reduce the rated capacity of a relief valve. When the relief valve discharge to atmosphere, it should be checked that the pressure drop from the relief valve discharge to the atmospheric conditions does not exceed 10% of the valve gauge set pressure. If developed back pressure exceeds 10%, the discharge area should be corrected for back pressure.

The situation becomes more complicated when the relief discharges are collected and routed to a common vent or flare header. The likelihood of multiple simultaneous relief valve discharges to the common header must be considered and back-pressure to each relief device calculated. Determine then the back-pressure effect on the relieving capacity of each valve. Discharge lines and flare header size are determined in order to meet acceptable developed back pressure for any pressure relief valve in the system.

Design criteria for PSV outlet line are :

- The build up back pressure should be less than
  - 10 % of the valve gauge set pressure for a conventional safety valve

[7] for definition of Mach Number refer to Process Design rule A0042
[8] given by vendor or calculated using actual safety valve area

  - 50 % of the valve gauge set pressure for a balanced safety valve

- Mach number[9] should be lower than[10] :
  - 0.9 in PSV outlet line
  - 0.8 in flare sub header
  - 0.5 in flare main header
- In case of Oxygen service, rules for maximum velocity are applicable (refer to Process Design rule A0017)

Velocity and back-pressure should be calculated using :

- the maximum rated flow through the pressure relief valves[11] for PSV outlet line.
- design required capacity for common headers

[9] for definition of Mach Number refer to Process Design rule A0042
[10] absolute limits not to be exceeded in any case
[11] given by vendor or calculated using actual safety valve area

9. DETERMINATION OF INDIVIDUAL RELIEVING RATE

9.1. GENERALITIES

9.1.1. OVERPRESSURE CRITERIA

Generally the simultaneous occurrence of two or more conditions that could result in overpressure will not be postulated if the causes are unrelated. The causes of overpressure, including external fire, are considered to be unrelated if no process or mechanical or electrical linkages exists among them, or if the length of time that elapses between possible successive occurrence of these causes is sufficient to make their classification unrelated. For example, if one assumes loss of instrument air, he/she cannot add the loss of power without demonstrating a cause-effect relationship between the two utility failures. It is safe to assume, however, that if power is lost, and there is no back-up system in place (UPS), and the cooling water pump motors are electric, then a power failure would also lead to the loss of cooling water.

However, in some cases it is necessary to consider the simultaneous occurrence of two independent causes, for example :

- Two independent defaults one of whom is dormant (open blocked check valve …)
- Two independent defaults but with each a high probability (default on valve while a pump is stopped…)
- Two operator errors following a same logic

Sound engineering judgment must be used when considering the contingency effects.

9.1.2. SOURCES OF OVERPRESSURE

The liquid or vapor rates used to establish relief requirements are developed by the net energy input. The two most common forms of energy are heat input, which is indirect pressure input through vaporization or thermal expansion and direct pressure input from higher pressure source.

9.1.3. EFFECT OF PRESSURE, TEMPERATURE AND COMPOSITION

Pressure and temperature should be considered to determine individual relieving rates, since they affect the volumetric and compositional behavior of liquids and vapors. Vapor is generated when heat is added to a liquid. The rate at which vapor is generated changes with equilibrium conditions. If the liquid is a mixture of components with different boiling points, composition of generated vapor changes. During pressure relieving, the changes in vapor rates and molecular weights at various time intervals should be investigated to determine the peak relieving rate and the composition of the vapor.

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Relieving pressure may sometimes exceed the critical pressure of the components in the system. In such cases, reference must be made to compressibility correlations to compute the density-temperature-enthalpy relationships for the system fluid.

9.1.4. EFFECT OF OPERATOR RESPONSE

Particular care should be exercised in assessing operator intervention. A commonly accepted time range for the response is between 10 and 30 minutes depending of plant complexity. The effectiveness of this response depends on the process dynamics. Moreover, it is rarely sure that the problem and the appropriate action will be clear to operator. Stress situation or believed good intention could lead to an error or incorrect action. For all these reasons <u>it is not recommended to take credit for operator intervention.</u>

9.1.5. OVERPRESSURE PROTECTION BY INSTRUMENTS

<u>Fail-safe devices, trip systems and automatic start-up equipment should not replace pressure relieving devices as protection for individual process equipment</u>. They should be regarded as additional safeguarding, installed to reduce the likelihood of a relief case occurring.

However, in some particular cases where protection by pressure safety valve is unpractical or impossible it is accepted to protect by instrumentation. Protection system are then case by case defined in order to meet probabilistic and deterministic requirements. Some of these particular cases are described in following chapters.

9.1.6. AUTOMATIC CONTROL

<u>No credit should be taken for any favorable instrument response</u>. Normal valve position is the expected position of the valve prior to the upset incident. Credit may be taken for the normal flow of these valves corrected to relieving conditions (provided that the downstream system is capable of handling any increased flow) but only to the extent permitted by the operating position at normal flow, although controllers actuated by variables other than the system pressure may try to open their valves fully. On the other hand, any unfavorable action shall be taken into account at the relief requirement evaluation.

Each outlet control valve should be considered both in the fully opened and fully closed positions, regardless of the control valve failure position. No reduction in relief capacity should be considered when fail stationary valves are used.

9.2. BLOCKED OUTLET

The inadvertent closure of a block or control valve on the vapor or liquid outlet of a pressure vessel while plant is on stream may expose the vessel to a pressure exceeding the MAWP. If closure of an outlet valve can result in overpressure, a pressure relief device is required. Every control valve should be considered as being subject to inadvertent operation.

A blocked outlet could also be caused by a fixed bed reactor blockage or blockage by equipment internals (baffles, trays, demister mats, vortex breaker…). These elements are generally not considered as potential overpressure causes in our processes. On the other hand, we consider the possible blockage by a liquid height in a pipe or vessel bottom.

It is not recommended to prevent overpressure by installing a locking system on a block valve. In some exceptional cases this can be accepted but a strict administrative procedure, fully documented, should then be put in place.

9.2.1. RELIEF CAPACITY CALCULATION:

For system capacity design, it may be assumed that control valves which are normally open and functioning at the time of failure and that are not affected by the primary cause of failure will remain in operation at their normal operating position.

The quantity of material to be relieved should be determined at relieving conditions instead of at normal operating conditions. In particular temperature increase due to pressure increase is to be taken into account.

- for applications involving single outlet with control device that fails in the closed position, the required relief capacity is equal to the maximum expected inlet flow at relieving conditions.
- for applications involving more than one outlet with control device on one individual outlet that fails in the closed position, the required relief capacity is the difference between the maximum expected inlet flow and the design flow through the remaining outlets, assuming that the other valves in the system remain in their normal operating conditions, at relieving conditions.
- for applications involving more than one outlet, each with control device that fails in the closed position because of the same fail, the required relief capacity is equal to the maximum expected inlet flow at relieving conditions.
- **liquid relief** : capacity of pressure relief valve for liquid relief should be at least equal to the liquid pump-in rate at normal conditions. Any reduction in pump capacity at the set pressure of the PSV is normally not taken into account. Liquid relief of distilling columns (overfilling) is not normally considered as the volume capacity below the relief valve usually provides ample time for operator intervention in case of blocked liquid outlet.
- **reciprocating pumps and reciprocating compressors**: relief facilities are required at the discharge of reciprocating machines to prevent overpressure and equipment damage in case of a blocked outlet. The relief capacity is equal to the design capacity of the machine with suction pressure equal to upstream PSV set pressure[12], and suction temperature taken at minimum possible at compressor inlet.

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[12] In some cases, it is unpractical to use suction set pressure. A particular rule has been defined, described in particular cases.

- **Oil lubricated screw compressors:** relief facilities are required at the discharge of screw compressors to prevent overpressure and equipment damage in case of a blocked outlet. The relief capacity is equal to the design capacity of the machine with suction pressure equal to normal operating pressure and discharge pressure equal to set pressure, and suction temperature taken at minimum possible at compressor inlet.

- **centrifugal pumps** : discharge system of a centrifugal pump is generally designed for the stalling pressure[13] in order to prevent a blocked outlet relief case rather than to install a PSV. When a PSV is installed in the pump discharge to limit the design pressure of the downstream equipment, the quantity to be relieved is taken as the normal pump capacity (at trip speed for a variable speed drive).

- **centrifugal compressors** : if discharge system is designed for a pressure lower than 120% of maximum discharge pressure of the compressor, a PSV is installed at compressor outlet (supplier curves are theoretical, they can be pessimistic). The quantity to be relieved is taken at the maximum capacity of the machine with suction pressure equal to upstream PSV set pressure[1], suction temperature taken at minimum possible at compressor inlet and discharge pressure equal to downstream PSV set pressure[14] (extrapolate the curves if needed). Before taking decision to install no PSV, be sure considered suction pressure is the maximum one, and it is acceptable by local regulation.

- system with partial condenser : closing of outgoing vapor line results of total loss of condensing capacity (accumulation of non condensable of the condenser). See § 9.4.

9.3. VALVE OPENING

There may be single or multiple inlet lines fitted with control devices or block valves The scenario to consider is that one inlet valve will be in a fully opened position, <u>regardless of the control valve failure position</u>. If the system has multiples inlets, the position of any control device in those remaining lines shall be assumed to remain in its normal operating position. Inadvertent valve opening of any valve (control and bypass valves, block valves, drain valves…) must be considered.

9.3.1. RELIEF CAPACITY CALCULATION :

The quantity to be relieved is the difference between the maximum expected inlet flow and the normal outlet flow, adjusted for relieving conditions, assuming that the other valves in the system are still in operating position at normal flow.

In the case of a pressure vessel operating at high pressure where liquid bottom is on level control and discharges into a lower pressure system, it should be considered that vapors will flow into the low pressure system if loss of liquid level occurs in the vessel at higher pressure. In this case, serious

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[13] Stalling pressure calculation : inlet PSV set pressure + maximum suction hydraulic height + pump head at zero flowrate at maximum motor speed.
[14] Curves P2/P1=f(actual flowrate) can be obtained from supplier and are very useful for this type of calculation

overpressure can rapidly develop. Relief devices on the LP system should therefore be sized to handle the full vapor flow through the liquid control valve.

Flow through the fully open valve should be calculated for the actually installed valve (installed CV) with a differential pressure based on normal operating pressure upstream and set pressure downstream[15]. In case of liquid valve, the flash is to be considered. In case of liquid level loss, fluid to consider is the vapor normally contained in the upstream vessel.

Where limit stops are installed on valves, the wide open capacity rather than the capacity at the stop setting should be used, excepted if limit stop can be mechanically locked (it is not the case in the systems we generally use).

This pressure drop at initial conditions frequently results in critical flow and may cause the rate to be several times higher than the normal rate of vapor inflow to the LP system. Unless make up equals outflow, this condition will be of short duration as the upstream capacity is depleted. Nonetheless, the relief facilities that protect the LP system must be sized to handle peak flow, except if it can be proven that the additional flowrate may be absorbed without overpressure. If the LP side has a large vapor volume, it is also accepted to take credit for the following : the transfer of vapor from the HP system needed to raise the pressure of the downstream side from operating pressure to set pressure will lower the upstream pressure, producing a corresponding reduction of the flow. Where such credit is taken, an allowance must be made for the normal makeup of vapor to the HP system, which tends to maintain upstream pressure.

In case of a control valve equipped with a bypass, the largest contribution of either the fully open control valve or the fully open bypass valve shall be taken into account. If the event of both control and bypass valve being fully open is not unrealistic, both flow rates should be added.

9.4. COOLING OR REFLUX FAILURE

This can occur if a control valve fails, power is lost, or if the pumps fail. In many cases, failure of the reflux that results, for example from pump shutdown or valve closure will cause flooding of the condenser which is equivalent to total loss of coolant The required relieving rate is determined by a heat and material balance on the system at the relieving pressure. Because of the difficulty in calculating detailed heat and material balances, the simplified bases described here are accepted for determining relieving rates.

9.4.1. TOTAL CONDENSING

The relief requirement is the total incoming vapor rate to the condenser, recalculated at temperature that corresponds to set pressure.

[15] If maximum operating pressure upstream valve is less than set pressure downstream, no relief flow rate for valve opening is to be considered. Refer to § 11.2 for example.

9.4.2. PARTIAL CONDENSING

The relief requirement is the difference between the incoming and outgoing vapor rate at relieving conditions.

9.4.3. FAN FAILURE

Fans on air cooled heat exchangers or cooling towers occasionally become inoperative because of a loss of power or a mechanical breakdown. On cooling towers and air-cooled heat exchangers where independent operation of the louvers can be maintained, credit for the cooling effect may be obtained by convection and radiation in still air at ambient conditions. Credit for a partial condensing capacity of 20 percent of normal duty can be used, the capacity of safety valve is then based to the remaining 80 percent, depending on the service (total or partial condensing).

9.4.4. LOUVER CLOSURE

Louver closure on air-cooled condensers is considered to be total failure of the coolant with the resulting capacity established above.

9.4.5. PUMP-AROUND CIRCUIT

The relief requirement is the vaporization rate caused by an amount of heat equal to that normally removed in the pump-around circuit. Latent heat of vaporization would correspond to the latent heat under relieving conditions of temperature and pressure at the point of relief.

9.4.6. ACCUMULATION OF NONCONDENSABLES

Noncondensables do not accumulate under normal conditions, because they are released with the process streams. However, in case of blocked outlet, noncondensables will accumulate to the point that the condenser is blocked. This effect is equal to a loss of coolant.

9.5. ENTRANCE OF VOLATILE MATERIAL INTO THE SYSTEM

Entrance of water or light hydrocarbon into hot oil, of a hot fluid into a cryogenic liquid or of a cryogenic liquid into an ambient system are sources of potential overpressure. However, expansion in volume is so great (approximately 1:1400 at atmospheric pressure for water) and the speed of vapor generation is essentially instantaneous, normally no pressure relieving device is provided for this contingency. Proper design and operation of the process are essential in attempting to eliminate this possibility.

9.6. UTILITY FAILURE

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The consequence that may develop from the loss of any utility service, whether plantwide or local must be carefully evaluated. An evaluation of the effect of overpressure that is attributable to the loss of a particular utility service should include the chain of developments that could occur and the reaction time involved. In situation in which the equipment fails but operates in parallel with equipment that has a different energy source, operating credit may be taken for the unaffected and functioning equipment to the extend that service is maintained. No protective credit is to be taken for a standby equipment, even with separate energy source and equipped with automatic start up (not considered totally reliable).

9.6.1. ELECTRICAL OR MECHANICAL FAILURE

The failure of electrical or mechanical equipment that provides cooling or condensation in process streams can cause overpressure in process vessels.

Determination of relieving requirements resulting from power failures requires a careful plant or system analysis to evaluate what equipment is affected by that power failure and how failure affects plant operation.

Analysis should be done in the following three ways :

9.6.1.1 As a local power failure

It is a failure in which one piece of equipment is affected (pumps, compressors, valves, fans...). Most of these effects are covered in other section of this guide (for example a pump failure can cause a loss of cooling water or a loss of reflux).

- centrifugal pump trip : it is assumed that a stopped centrifugal pump offers no resistance to backward flow and that the check valve normally installed in the pump discharge is not a sufficient protection. Overpressure protection of the pump suction system should be based on PSV on the suction vessel but sizing of the PSV on the basis of the vapor backflow is considered impracticable. It is the reason why a protection system by instrumentation is adopted, including at least upstream the pump an automatic shut off valve (failure close).

- centrifugal compressor trip : whenever possible overpressure protection of the compressor suction system should be based on designing the suction system for the settling out pressure[16], otherwise a pressure relief valve is required. It is assumed that a stopped centrifugal compressor offers no resistance to backward flow and that the check valve normally installed in the compressor discharge is not a sufficient protection. It should also be checked if the compressor has a recycle line which could give a relief case. Overpressure protection of compressor suction depends on compressor function and is described in particular cases.

- reciprocating pumps : the suction and discharge valves within the pump are normally considered to be sufficient protection against backflow.

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[16] pression d'équilibrage

- reciprocating compressors : same as reciprocating pumps except in case of suction valve unloaders which automatically unload on shutdown : partial backflow is possible and should be considered as relief case for suction system.

9.6.1.2 As an intermediate power failure

It is a failure in which one distribution center, one motor control center or one bus is affected. Several individual equipment could be affected simultaneously and relief causes should be added.

9.6.1.3 As a total power failure

It is a failure in which all electrical operated equipment is simultaneously affected. All individual equipment are affected simultaneously and relief cases should be added. Credit should therefore be taken for loss of equipment which will reduce relief causes

9.6.2. LOSS OF INSTRUMENT AIR OR ELECTRIC POWER FOR INSTRUMENTATION

To minimize the likelihood of overpressure, each control valve should have its fail-safe characteristic properly established as an integral part of the plant design. The failure position of a control valve in itself is not considered to be sufficient relief protection as other failure in an instrument system can cause a control valve to move in a direction opposite its design failure position. Effects of these defaults are described in other sections (open valve, blocked outlet, loss of coolant…).

9.7. ABNORMAL HEAT INPUT FROM REBOILER OR VAPORIZER

Reboilers and vaporizers are designed with a specified heat input. In some abnormal events (failure of heat medium flow or pressure control, heat medium composition and/or temperature modification during transient phases…) additional heat input above the design can occur and vapor generation can exceed the normal condensing capacity of the process.

The increased duty is to be calculated considering the same UA as in normal operation. The required relief capacity is then the maximum rate of vapor generated at relieving conditions less the rate of normal condensation or vapor outflow. In shell and tubes heat exchangers, heat input should be calculated on the basis of cleaned rather than fouled conditions.

9.8. HEAT EXCHANGER TUBE FAILURE

In shell and tube heat exchangers, the tubes are subject to failure from a number of causes, including thermal shock, corrosion and vibration. The result is the possibility that the high pressure stream will overpressure equipment on the lower pressure side of the exchanger.

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9.8.1. PRESSURE CONSIDERATION : "2/3 RULE"

Since standard hydrostatic test pressure is 150 % of the equipment design pressure, equipment failure is unlikely to result from a tube rupture where the low pressure side (including upstream and downstream systems) is designed for at least 2/3 of the design pressure of the high pressure side. Pressure relief for tube rupture is not required where the low-pressure side (including upstream and downstream systems) is designed at or above this two-thirds criteria.

Where the actual test pressure of the low-pressure side is less than 150% of the design pressure, this lower pressure should be used to determine whether overpressure protection is needed[17]. In other words, the "2/3" rule should be generalized with : if test pressure of LP side exceeds or is equal to the design pressure of the HP side, PSV for tube rupture is not required.

Note that this rule cannot be applied if the high pressure side of the heat exchanger operates at 70 barg or more.

9.8.2. REQUIRED RELIEF FLOW RATE

In practice, an internal failure can vary from a pinhole leak to a complete tube rupture. For the purpose of determining the required flow rate, the following basis should be used :

- the tube failure is a sharp break in one tube
- the high-pressure fluid is assumed to flow both through the tube stub remaining in the tube sheet and through the other longer section of tube : a simplifying assumption of two orifices may also be used.

Allowance should be made for any liquid that will flash to vapor either as a result of pressure reduction and eventually vaporization as the fluid is intimately contacted by the hotter material on the low-pressure side.

If a steady-state method is used, the relief device size should be based on the gas and/or the liquid flow rate passing through an orifice which area is twice the tube section, with upstream pressure equal to normal operating HP pressure and downstream pressure equal to LP set pressure.

Following formulas[18] are used to calculate the flowrate through the ruptured tube :

- For a gas with $r \le \left(\frac{2}{k+1}\right)^{\frac{k}{k-1}}$ (critical flow) :

[17] For example where test pressure is 130% of design pressure, a PSV is not required if LP side is designed for at least 1/1.3 of the HP side design pressure.

[18] These equations correspond to orifice calculation with β=0. This hypothesis, obvious in the case of HP fluid in the shell, is also conservative when HP fluid is in the tube. In fact we consider then that restriction orifice is at the tube inlet and reduction of flow due to pressure drop is neglected.

Eq. 7 $$W = CA\sqrt{1.296kP_1\rho_1\left(\frac{2}{k+1}\right)^{\frac{k+1}{k-1}}}$$

- For a gas with $r > \left(\frac{2}{k+1}\right)^{\frac{k}{k-1}}$ (subcritical flow) or for a liquid :

Eq. 8 $$W = CYA\sqrt{1.296 * 2(P_1 - P_2)\rho_1}$$

Where :

-r = P_2/P_1

-W = fluid quantity to be relieved in kg/h.

-P_1 = normal operating pressure HP side in bar abs.

-P_2 = set pressure LP side in bar abs.

-k = C_p/C_v.

-C = discharge coefficient. Take a conservative value of 0.9 for a gas and 0.7 for a liquid.

-A = twice the cross sectional area of exchanger tube in mm^2.

-ρ_1 = density of fluid at upstream conditions in kg/m^3.

-Y = expansion factor, dimensionless. Y = 1 for a liquid and $Y = 1 - 0.41\left(\frac{1-r}{k}\right)$ for a gas

For a two-phase flow or a flashing liquid, the vapor ratio at orifice needs to be defined by isenthalpic expansion from upstream conditions to downstream set pressure or critical pressure of the vapor, whichever is greater. Vapor and liquid flow are then to be calculated separately using above equations and affecting for each phase the area proportional to its volumetric ratio.

If a steady state method is used, the relief device should be specified for this gas and/or liquid flowrate

Relieving temperature can be taken equal to upstream temperature.

9.8.3. DYNAMIC CONSIDERATION

A dynamic approach simulates the pressure profile and pressure transients developed in the exchanger from the time of the rupture, and generally will include the response time of the relief device. Where there is a wide difference in design pressure between the two exchanger sides, especially where the low-pressure side is liquid-full and the high-pressure side contains a gas, it will create an acceleration of the LP liquid, thus creating a pressure wave. Mathematical modeling has shown that under these circumstances, transient conditions may produce significant overpressure, even when protected by pressure relief device. Furthermore, in these cases, the initial fluid to be relieved is liquid at vapor volumetric flow[19]. In many cases, the resulting size is so large that this form of protection is impossible. The use of a rupture disk can then be considered or the low pressure design pressure should be raised.

9.8.4. RELIEF DEVICE LOCATION

The relieving device may need to be located, either directly on the exchanger or immediately adjacent on the connected piping. This is specially important if the low-pressure side is liquid-full. In that case, the time interval in which the shock wave is transmitted to the relieving device from the point of the tube failure will increase if the device is located remotely.

9.9. PLANT FIRE[20]

Pressure relief valve designed for fire case have to be installed on equipment (generally not on pipes) which can be blocked in and which are located in an area where a fire can be expected (following API recommendation, fire area to consider are limited to 500 m²).

9.9.1. EFFECT OF FIRE ON THE WETTED SURFACE OF A VESSEL

The surface area wetted by a vessel's internal liquid content is effective in generating vapor when the area is exposed to fire. To determine vapor generation, only that portion of the vessel that is wetted by its internal liquid and is equal to or less 8 m above the source of flame (ground or any level at which a substantial spill or pool fire could be sustained) needs to be recognized.

Table 2 : wetted surface

Type of vessel	Portion of liquid inventory	Remarks
liquid-full	all up to the height of 8 m	
surge drum, knock out drums, process vessels	normal operating level up to the height of 8 m	
distillation columns	normal level in bottom plus liquid holdup from all trays up to the height of 8m	level in reboiler is to be included if reboiler is an integral part of the column
working storages	maximum inventory level up to the height of 8 m (portions of wetted area in contact with foundations or the ground are normally excluded)	for tanks with operating pressure of 1 barg or less : see API 2000
spheres and spheroids	up to the maximum horizontal diameter or up to the height of 8 m, whichever is greater.	
air cooled exchanger :	wetted area = 0.3 times the bare-tube	

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[19] Vapor volumetric flow to be calculated at LP side set pressure
[20] Plant fire in case of vacuum insulated cryogenic equipment is described in § 9.13

condensing without subcooling	area up to the height of 8 m.	
air cooled exchanger : condensing with subcooling	wetted area = 0.3 times the bare tube area of condensing section plus bare tube area of subcooling section up to the height of 8 m.	
air cooled exchanger : gas or liquid cooling	wetted area = bare tube area up to the height of 8 m.	

It may be appropriate to add a percentage of the vessel area to account for vapor generation in piping associated with the vessel under consideration.

The surface area of an elliptical end is approximately given by $S = \Pi \frac{d^2}{4} * 1.41$ (where d is the column diameter).

9.9.1.1 Heat absorption equation

The amount of heat absorbed by a vessel exposed to an open fire is markedly affected by the type of fuel feeding the fire, the degree of which the vessel is enveloped by the flames (a function of vessel size and shape) and fireproofing measures. The following equation formulas are used to evaluate these conditions where there are prompt firefighting and drainage of flammable materials away from the vessels.

Eq. 9 $$Q = 37152\,FS^{0.82}$$

Where adequate drainage and firefighting equipment do not exist, following equation should be used :

Eq. 10 $$Q = 61036FS^{0.82}$$

Where :

-Q = total heat input to the wetted surface in kcal/h

-S = total wetted surface in m^2 (see Table 2).
For air cooled exchanger S is taken to an exponent of 1.0 instead of 0.82

-F = environment factor, dimensionless. F is equal to 1 for bare vessels, for air cooled exchangers and for all insulated vessels whose insulation is not specially fireproof designed. If it is proven that insulation is fireproof and will not be damaged by fire conditions and is well installed, environment factor can be obtained from following equation :

Eq. 11 $$F = \frac{\lambda(904 - T_f)}{56.9t}$$

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Where :

-λ = thermal conductivity of insulation at mean temperature between 904 °C [21] and process temperature at relieving conditions in kcal/h/m/°C

-t = thickness of insulation in mm

-T_f = temperature of vessel content at relieving temperature in °C

An environment factor F = 0.1 is taken for all vessels installed inside a cold box.

Note : these formulas and environment factor values are based on assumptions on firefighting efforts and adequate drainage facilities. When these assumptions are not appropriate, more rigorous methods of calculations may be warranted, considering fluid physical properties, vessel mass and insulation. If the fire is of sufficient duration, temperature will increase and vessel rupture occurs.

9.9.2. EFFECT OF FIRE ON THE UNWETTED SURFACE OF A VESSEL

Unwetted wall vessels are those in which the internal walls are exposed to a gas, vapor or super-critical fluid, or are internally insulated regardless of the contained fluids. These include vessels that contains separate liquid and vapor phase under normal conditions but become single phase (above the critical) at relieving conditions.

A characteristic of a vessel with an unwetted internal wall is that heat flow from the wall to the contained fluid is low as a result of the resistance of the fluid or any internal insulating material. Heat input from an open fire to the bare outside surface of an unwetted vessel may be sufficient to heat the wall to a temperature high enough to rupture the vessel

9.9.2.1 Heat absorption equations

The discharge areas for pressure relief devices on vessels containing super-critical fluids, gases or vapor exposed to open fires can be estimated using Eq. 12. In the use of this equation, no credit has been taken for insulation. It may be taken as in the case of wetted vessel.

Eq. 12 $$A = 18.23 \frac{F'S}{\sqrt{P_1}}$$

Where :

-A = effective discharge area of the PSV in cm^2

-S = exposed surface area of the vessel, up to the height of 8 m, in m^2

[21] 1660 °F

-P_1 = upstream relieving pressure in bar abs (accumulation = 21% for fire case. See § 7.3.4)

-F' can be determined from Eq. 13. Recommended minimum value of F' is 0.01; when the value is unknown, F'=0.045 should be used.

Eq. 13 $$F' = \frac{0.2}{CK_d}\left(\frac{(T_w - T_1)^{1.25}}{T_1^{0.6506}}\right)$$ where C is determined from Eq. 5.

where :

-K_d = coefficient of discharge (taken to 0.975 for sizing relief valves)

-T_w = vessel wall temperature in °K. The recommended maximum vessel wall temperature for the usual carbon steel plate materials is 593°C [22]. Where vessels are fabricated from alloy materials, the value for T_w should be changed to a more appropriate recommended maximum.

-T_1 = gas temperature absolute in °K, at the upstream relieving pressure, determined by the following equation :

Eq. 14 $$T_1 = \left(\frac{P_1}{P_n}\right)T_n$$

where :

-P_n= normal operating gas pressure, in bar abs

-T_n = normal operating gas temperature in ° K

The relief load can be calculated directly, in kg/hour by rearranging equation Eq. 1 and substituting Eq. 12 and Eq. 13 which results in equation as follows :

Eq. 15 $$W = 2.77S\sqrt{MP_1}\,\frac{(T_w - T_1)^{1.25}}{T_1^{1.1506}}$$

where :

-M = molecular weight of the gas in kg/kmol

-Z and K_b in Eq. 1 are assumed to be equal to 1

The derivations of Eq. 11, Eq. 12, and Eq. 15 are based on physical properties of air and perfect gas laws. The derivations assume that the vessel is uninsulated and has no mass, that the vessel wall

[22] 1100 °F

temperature will not reach rupture stress, and that there is no change in fluid temperature. These assumptions should be reviewed to ensure that they are appropriate for any particular situation.

9.9.3. FLUIDS TO BE RELIEVED

A vessel can contain liquid or vapor or both phases. The liquid phase may be subcritical at operating conditions and may pass into the supercritical range during the duration of a fire as temperature and pressure increase in the vessel.

The quantity and composition of the fluid to be relieved during a fire depend on the total heat input rate to the vessel and may be computed by means of one of the above formula, using appropriate values for wetted or exposed surfaces and environment factor. Once the total heat input is known, the quantity and composition of the fluid to be relieved can be calculated.

9.9.3.1 Vapor

For pressure and temperature conditions below the critical point, the rate of vapor formation is equal to the total rate of heat absorption (calculated per § 9.9.1.1) divided by the latent heat of vaporization. The vapor to be relieved is the vapor that is in equilibrium with the liquid under conditions that exist when the pressure relief device is relieving at its accumulated pressure.

The vapor and liquid composition may change as vapors are released from the system, resulting in temperature and latent heat values changes and affecting the required PSV size. Also, a multicomponent liquid may be heated at pressure and temperature that exceed critical for one or more components, vapors that are physically or chemically bound in solution may be liberated from the liquid upon heating. This is not a standard latent-heating effect but is more properly termed degassing or dissolution. Vapor generation is determined by the rate of change in equilibrium caused by temperature increase. For these and other multicomponent mixtures that have a wide boiling range, a time-dependant model may have to be developed where the total heat input to the vessel not only causes vaporization but also raises the temperature of the remaining liquid, keeping it at its boiling point.

The recommended practice of finding a relief vapor flow rate from the heat input to the vessel and from the latent heat of liquid contained in the vessel becomes invalid near the critical point, where the latent heat approaches zero and the sensible heat dominates. The rate of vapor discharge then only depends on the rate at which the fluid will expand as a result of the heat input. Equations of § 9.9.2.1 are to be used in this case.

9.9.3.2 Liquid

The hydraulic expansion formula given in § 9.12 may be used to determine the initial liquid relieving rate in a liquid-filled system when the liquid is still below its boiling point. However this rate is valid for a very limited time, after which vapor generation will become the determining contributor in the sizing.

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Should a pressure relief device be located below the liquid level of a vessel exposed to fire conditions, the pressure relief device should be able to pass a volume of liquid equivalent to the displacement caused by vapor generated by the fire.

9.9.4. PROTECTIVE MEASURES EXCLUDING INSULATION

Sometimes a pressure relief valve will not provide sufficient protection for an unwetted wall vessel or a vessel containing high-boiling point liquid. In these cases, additional protective measures should be considered such as water sprays, depressurizing, fireproofing, earth-covered storage and diversion walls.

9.10. PROCESS CHANGES, CHEMICAL REACTION

In some reactions and processes, loss of process control may result in significant change in temperature and/or pressure. The result could exceed the intended limits of the material selected or increase the pressure above the maximum allowable working pressure. Where normal pressure relieving devices cannot protect against these situations, controls are needed to prevent. For more details about the determination of vent system size for chemical reaction refer to API 521, § 3.13

9.11. INTERNAL EXPLOSION

Where overpressure protection against internal explosions caused by ignition of vapor-air mixtures is to be provided, rupture discs or explosion vent panels, not safety valve should be used (safety valve reacts too slowly to protect the vessel against the extremely rapid pressure buildup caused by internal flame propagation).

9.12. HYDRAULIC EXPANSION

It is the increase in liquid volume caused by an increase in temperature. It can result from several causes, most common of them are the following :

- Piping or vessel are blocked-in when they are filled with cold liquid and are subsequently heated by heat tracing, coils, ambient heat or fire.
- An exchanger is blocked-in on the cold side with flow in the hot side.
- Piping or vessel are blocked-in when they are filled with liquid at ambient temperature and are heated by direct solar radiation

The capacity requirement is not easy to determine. Since every application will be relieving liquid, the required capacity will be small : specifying an oversized device is therefore reasonable. A 3/4-inch x 1-inch nominal pipe size relief valve is commonly used.

Special cases : two general applications where thermal relieving devices larger than 3/4-inch x 1-inch nominal pipe size valve might be required are long pipelines of large diameter in uninsulated aboveground installations and large vessels or exchangers operating liquid-full.

For these cases, expansion rates for the sizing of relief devices that protects against thermal expansion of the trapped liquid can be approximated using the following formula :

Eq. 16 $$W = \frac{BQ}{C_p}$$

Where :

-W = flow rate at the flowing temperature in kg/h.

-B = cubical expansion coefficient per degree Celsius for the liquid at the expected temperature (value for water at 15.6°C is 0.0018 $°C^{-1}$)

-Q = total heat transfer rate, in kcal/hr. For heat exchanger, this can be taken as the maximum exchanger duty during operation.

-C_p = specific heat of the trapped fluid, in kcal/kg °C

9.13. VACUUM LOSS ON VACUUM INSULATED EQUIPMENTS

9.13.1. GENERAL

On vacuum insulated equipment and piping, a sudden vacuum loss will lead to a abnormal heat input.

On pipes, the flow is not calculated, a standard size for thermal expansion purpose is installed. <u>This type of safety valve are not considered as flow safety valve and are therefore not described in following chapters.</u> A 1/2 inch x 1/2 inch nominal pipe size relief valve is commonly used.

The maximum length protected by a thermal safety valve is given in Appendix 2 .

T(K) is the relief temperature: - For sub-critical fluid, T is the saturation temperature at relieving pressure Pa

- For critical or super-critical fluids, T is calculated in Équation d

For equipments (heat exchangers, adsorbers, pressure vessels), the flow to be relieved is calculated according to reference [8] and following equations.

Pa (bars abs) is the accumulation pressure (refer to 7.3.4)

Ta (K) is the maximum ambient temperature

T(K), the relief temperature is given above

The relief temperature is used in the following chapters for the heat input calculations, but also for the safety valve size selection (refer to 8.3.2), even in the case of a heat exchanger (see Appendix 1)

9.13.2. HEAT INPUT CALCULATION IN CASE OF VACUUM LOSS, NO FIRE

9.13.2.1 Heat input through insulation

$$\boxed{W_2 = (Ta - T) U_2 \Sigma}$$

W_2: heat transferred per unit time through the insulation in case of vacuum loss (W)

U_2: heat transfer coefficient of the insulating material under atmospheric pressure (W/m²/K) $U_2 = \lambda_2 / e$

e: insulation thickness (m)

Note: multi-layer insulation varies between 15 and 20 layers/cm.

- for T<20K, 30layers, e = 15 mm
- for 20K <T<80K, 20layers, e = 10 mm
- for T>80K, 10layers, e= 5 mm

For lower insulation thickness, refer to 11.12.

λ_2:conductivity of the insulating material saturated with either process gas or air at atmospheric pressure, whichever gives the highest heat transfer coefficient, between T and Ta (W/m/K)

Σ is the arithmetic mean value of the insulating material inner and outer surfaces (m²)

9.13.2.2 Heat input through pipes and supporting elements

$$\boxed{W_3 = (Ta - T)(w_1 + w_2 + \ldots + w_n)}$$

$$w_n = \lambda_n \frac{S_n}{l_n}$$

W_3: heat transferred per unit time through the supporting elements (W)

w_n : heat input per K through one of the supporting elements (W/K)

λ_n: conductvity of the supporting element between T and Ta (W/m/K)

S_n: section f the supporting element between T and Ta (m²)

l_n: length of the supporting element in the vacuum jacket (m)

9.13.2.3 Heat input through Pressure Build Up system

$$\boxed{W_4 = U_4\ A\ (Ta - T)}$$

W_4 :heat input per unit time through the pressure build up system, control valve being open (W)

U_4: heater overall heat transfer coefficient (W/m²/K)

A: heater external surface available for heat transfer (m²)

9.13.2.4 Overall heat input

$$W = \sum_i W_i$$

W: overall heat input (W)

9.13.3. HEAT INPUT CALCULATION IN CASE OF VACUUM LOSS AND FIRE

9.13.3.1 Heat input through the vessel walls

9.13.3.1.1 Insulation remains partially or totally in place

$$\boxed{W_5 = 2.6\ (922 - T)\ U_5\ \Sigma}$$

W_5: heat transferred per unit time through the insulation in case of vacuum loss (W)

U_5: heat transfer coefficient of the insulating material under atmospheric pressure (W/m²/K) $U_5 = \lambda_5 / e$

e: remaining insulation thickness (m)

λ_5:conductivity of the insulating material saturated with either process gas or air at atmospheric pressure, whichever gives the highest heat transfer coefficient between T and 922 K (W/m/K)

Σ is the arithmetic mean value of the remaining insulating material inner and outer surfaces (m²)

9.13.3.1.2 Insulation is destroyed

$$\boxed{W_6 = 7.1\ 10^4\ \sigma^{0.82}}$$

W_6: heat input to the vessel per unit time (W)

σ is the outer surface of the inner vessel (m²)

9.13.3.2 Heat input through pipes and supporting elements

In case of fire, the heat input trough the pipes and supporting elements can be neglected.

9.13.3.3 Overall heat input

$$W = \sum_i W_i$$

W: overall heat input (W)

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9.13.4. MASS FLOW TO BE RELIEVED

9.13.4.1 P < 40%Pc

Équation a $$\boxed{\mathrm{Qm} = 3.6\frac{W}{L}\cdot}$$

T : relief temperature = saturation temperature at relieving pressure (K)

Qm: mass flow to be relieved (kg/h)

L: latent heat at relieving conditions (kJ/kg)

9.13.4.2 40%Pc ≤ P < Pc

Équation b $$\boxed{\mathrm{Qm} = 3.6\left[\frac{vg - vl}{vg}\right]\frac{W}{L}\cdot}$$

T : relief temperature = saturation temperature at relieving pressure (K)

Qm: mass flow to be relieved (kg/h)

W: overall heat input (W)

L: latent heat at relieving conditions (kJ/kg)

vg: specific volume of saturated gas, at relieving pressure P (kg/m3)

vl: specific volume of saturated liquid, at relieving pressure P (kg/m3)

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9.13.4.3 P > Pc

Équation c $\boxed{Q_m = 3.6 * \frac{W}{L'}\cdot}$

T : relief temperature (K)

Équation d T is such that $\dfrac{\sqrt{\nu}}{\nu.\left(\dfrac{\partial \mathrm{h}}{\partial \nu}\right)_p}$ is maximum

ν = specific volume of the fluid, at relieving pressure and any temperature within the operating range (kg/m3)

h = enthalpy of the fluid at relieving pressure P and any temperature within the operating range (kJ/kg)

Qm: mass flow to be relieved (kg/h)

L' : specific heat $\nu\left(\dfrac{\partial \mathrm{h}}{\partial \nu}\right)_P$ at relieving temperature T (kJ/kg)

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10. AIR LIQUIDE UNITS PARTICULAR CASES

10.1. THERMAL EXPANSION

Thermal expansion will occur whenever a system can be blocked in. This is particularly acute with cryogenic liquid services which can vaporize but also with blocked in cryogenic gas (pressure increases proportionally to absolute temperature). **Any cryogenic service lines and equipment should therefore be considered for thermal expansion purposes.** All possible blocked-in cases should be reviewed and a safety valve should be installed.

On equipment (vessels, columns, heat exchangers…) the flow to be relieved in case of thermal expansion is to be calculated. It is equal to the heat input (used on equipment for thermodynamic calculation) divided per liquid heat of vaporization at set pressure. Relief temperature is equal to boiling temperature at set pressure. For vacuum insulated equipment, vacuum break must be taken into account and the flow to be relieved must be calculated according to 9.13

On pipes, the flow is not calculated, a standard size for thermal expansion purpose is installed. Relief temperature is calculated per Eq. 14 for a gas and is equal to boiling temperature at set pressure for a liquid. <u>This type of safety valve are not considered as flow safety valve and are therefore not described in following chapters</u>.

10.2. PLATE FIN EXCHANGER PARTING SHEET FAILURE

It is considered that any possible leaks in these exchangers is small and will not constitute a valid failure option (non-significant flows). Possible leaks will be detected through standard purity analyses and pressure monitoring systems.

10.3. INTRODUCTION OF DERIME OR INERTING GAS TO LOW PRESSURE CIRCUIT

The maximum flow through the derime or inerting valves must be considered as a possible relief scenario for the low pressure circuits (calculation per § 9.3.1).

10.4. INTRODUCTION OF PROCESS GAS INTO DERIME OR INERTING CIRCUIT

This contingency is normally not considered if sufficient protection against back flow is provided (isolation valve plus check valve, double block and bleed…).

10.5. INTRODUCTION OF LIQUID NITROGEN TO LOW PRESSURE CIRCUIT

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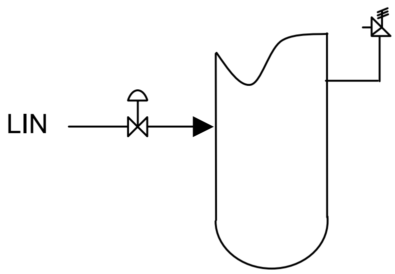
The maximum flow through the liquid nitrogen inlet valve must be considered as a possible relief scenario for the low pressure circuit.

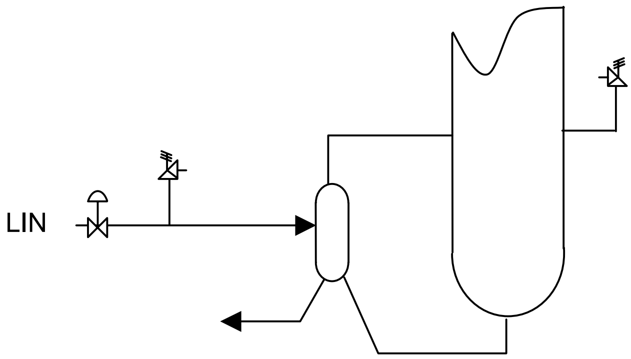
Case 1 : LIN is introduced into a large vessel (column, vaporizer shell…) which is protected by a safety valve. This PSV must handle the maximum between :

- gaseous flowrate through the valve
- volume of gas equivalent to liquid volume through the valve plus generated flash

Case 2 : LIN is introduced into a line or a vessel which is not directly protected by a PSV and whose degassing line is small (for example quick load changing capacity), a PSV has to be added downstream the valve, designed to handle the maximum liquid flowrate through the valve. This liquid relief PSV must be collected.



CASE 1



CASE 2

10.6. WATER COOLERS

In our units, application of the so called "2/3 rule", installation or not of safety valve and design of this PSV depend on many parameters (gas involved, pressure, fluid in the tube,…). Refer to applicable Water Coolers Standard W-EP-3-6-1. If a PSV is installed, relief flow is to calculate following § 9.8.2.

If water can be blocked in the exchanger (isolation valves upstream and downstream) a safety valve has to be installed :

- where gas inlet temperature is higher than water boiling temperature at water side set pressure, PSV is sized for vapor flowrate based on water vaporized with a duty corresponding to the normal gas flow rate cooled down from normal inlet temperature up to water boiling temperature. This PSV has to be installed on a high point.
- thermal expansion safety valve if there is no possibility of water vaporization.

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10.7. PURGING CIRCUITS[23]

It should be checked that during all phases, including transient phases such as circuit purging (depressurizing - pumping – helium re-filling), there is no possibility of connecting two circuits with different design pressures by operating one only valve.

On purging circuits, the chosen valves have large Cv, that it is not reasonable to consider the "valve opening case" described in 9.3, it would lead to very large safety valves operating in parallel.

Double isolation valve is the chosen safety rule in that case.

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[23] Circuits de pompage / conditionnement

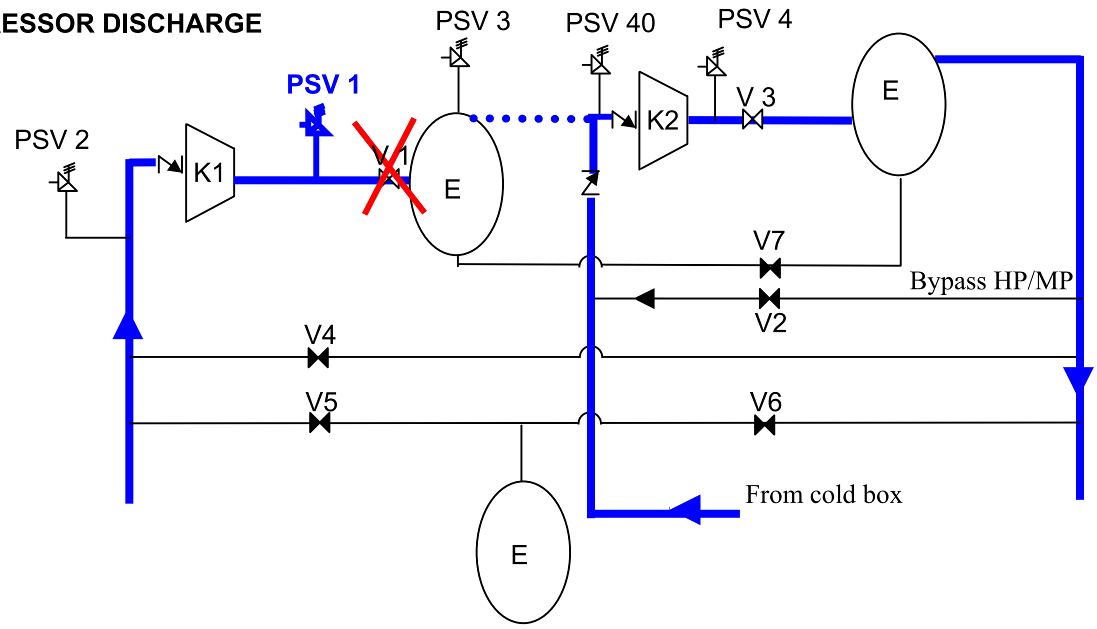
11. HELIUM REFRIGERATION / LIQUEFACTION UNITS PARTICULAR CASES

11.1. PLANT FIRE

In helium refrigeration / liquefaction plants, the external fire contingency is generally not considered, except in very specific cases where the **risk analysis** has shown that this contingency is likely. It may be the case if there is an opportunity for a fuel source to ignite (lube oil systems for compressors are not included as fire fuel sources, as BREOX B35 is a non flammable product) or in case of local regulation requirement.

11.2. LIQUEFACTOR CYCLE COMPRESSOR

11.2.1. COMPRESSOR DISCHARGE



It is assumed that the design pressure of the K1 and E1 are equal, and that the design pressures of K2, E2 and E4 are also equal.

K1 discharge protection

The relief case at compressor discharge is a blocked outlet. Relief flow rate calculation is compressor capacity with suction pressure equal to normal pressure and discharge pressures equal to PSV1 set pressure (refer to § 9.2.1).

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If V1 does not exist (the isolation valve is further on the HP line), PSV1 can be located on the oil separator E1.

If V1 exists, PSV1 must be located upstream of V1.

Note: The compressor is a pressure source. It is allowed to locate the pressure relief valve on a pipe and not on the compressor itself (see ASME VIII UG 125 (g)(2)), but it is not allowed to protect the compressor with a safety relief valve located on the oil separator unless there is no block valve in between the compressor and the oil separator. If this block valve is under positive control (car seal open valve, with all due documentation and personnel training), it is tolerated see (ASME VIII appendix M) but still not recommended to protect the compressor with a safety relief valve located on the oil tank (refer to 9.2)

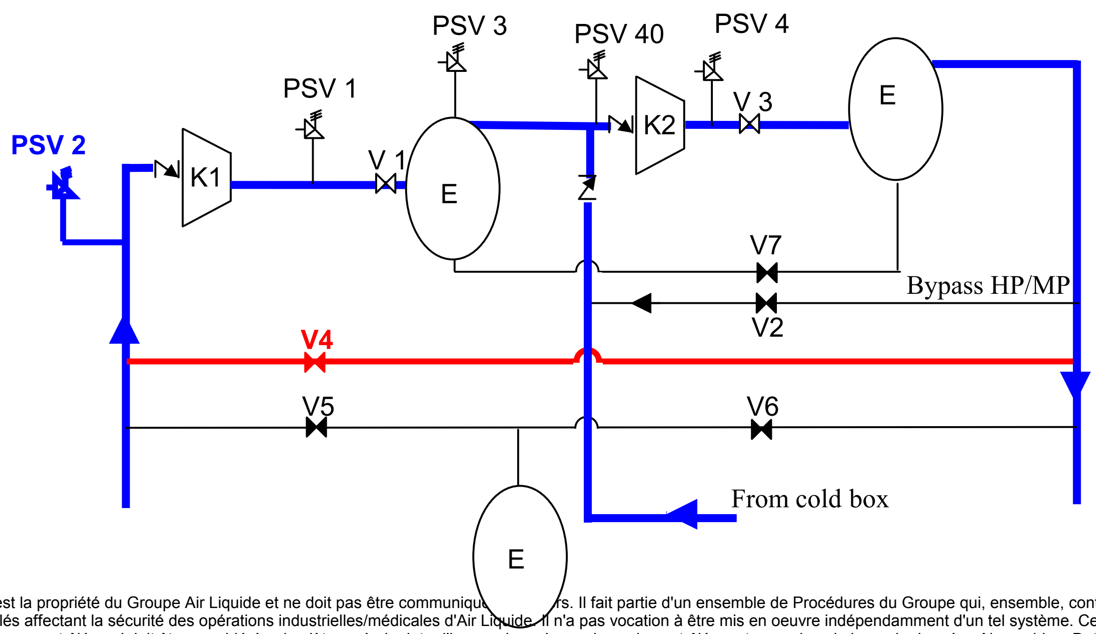
K2 discharge protection Proceed as for K1

11.2.2. COMPRESSOR SUCTION

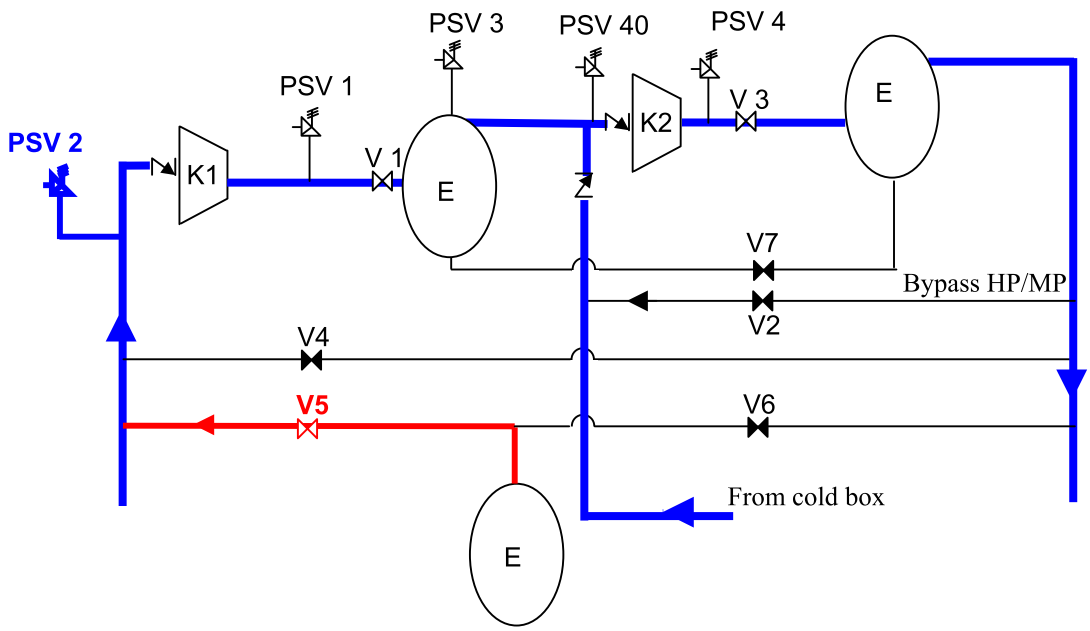
K1 suction protection

PSV2 must be sized considering the following cases :

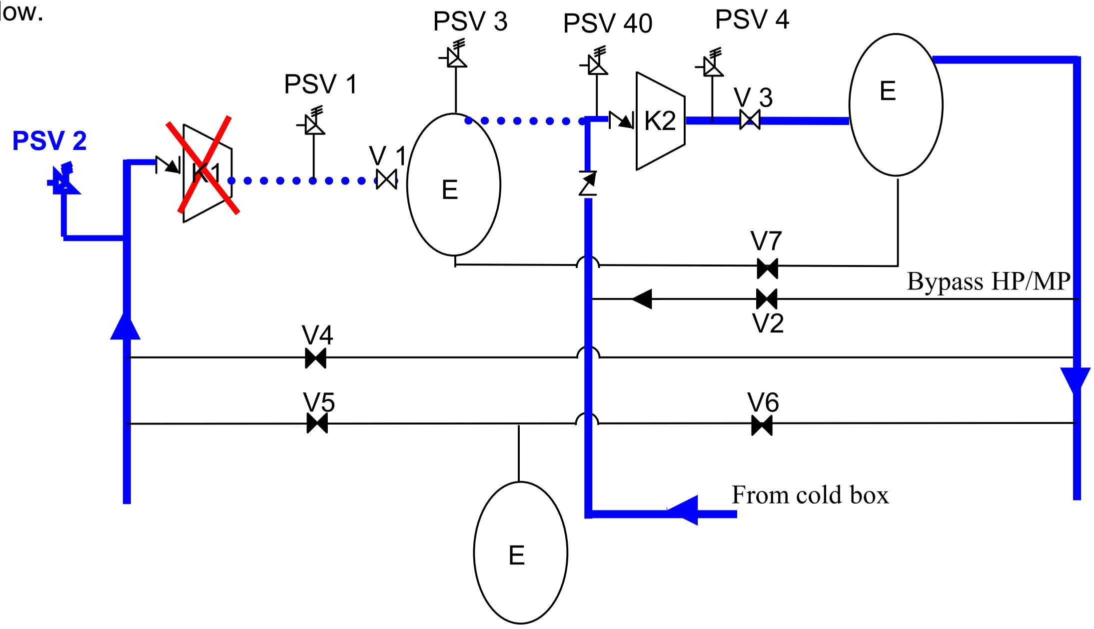
- Scenario 1 full opening of V4 (HP to LP bypass valve), high pressure level being normal, low pressure level equal to PSV2 set pressure



- Scenario 2 full opening of V5 (cycle loading valve) V5 upstream pressure being equal to the buffer maximum operating pressure, low pressure level equal to PSV2 set pressure



- Scenario 3 compressor K1 stops: the flow-rate to be evacuated through PSV2 is the compressor nominal flow.



The largest contribution must be taken into account

(refer to § 9.3.1)

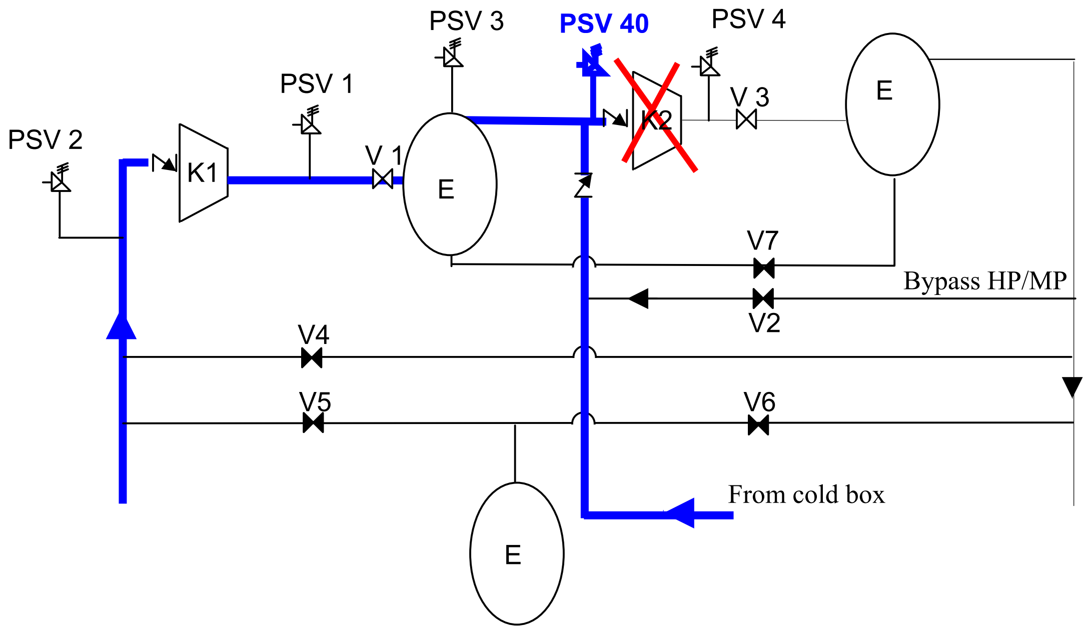
K2 suction protection

In most cases, the set pressure of PSV40 will be the same as the set pressure of PSV1.

In that case, PSV40 protects both E1 and K2 suction.

PSV40 must be sized considering the following cases :

Scenario 1: Compressor K2 stops



The flow to be relieved is K2 nominal flow-rate. This must be relieved through PSV40.

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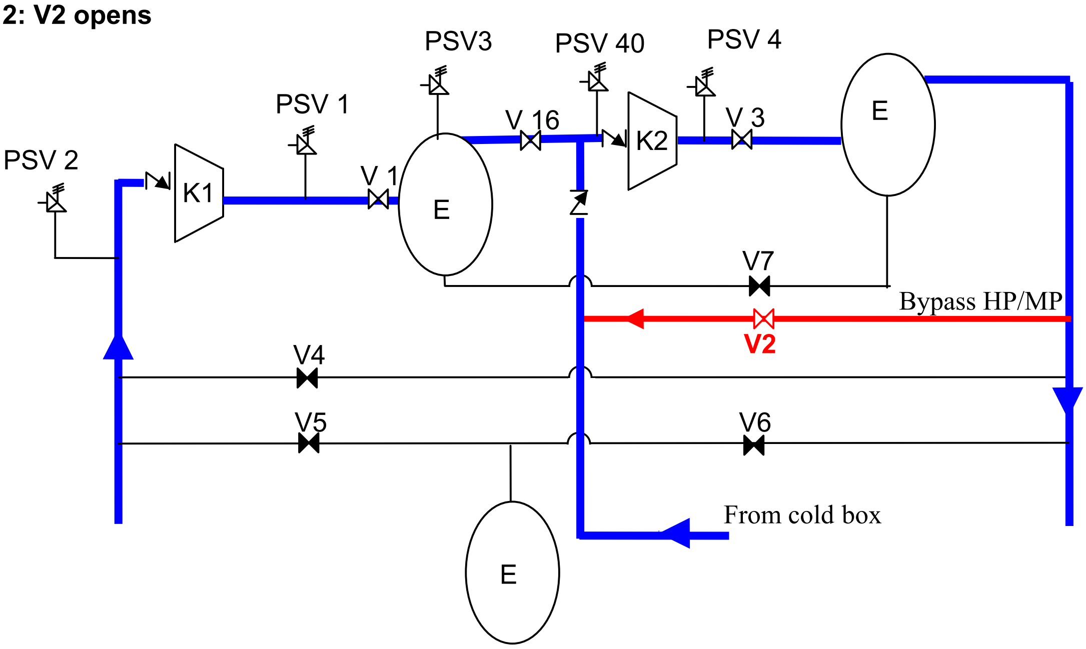
Scenario 2: V2 opens

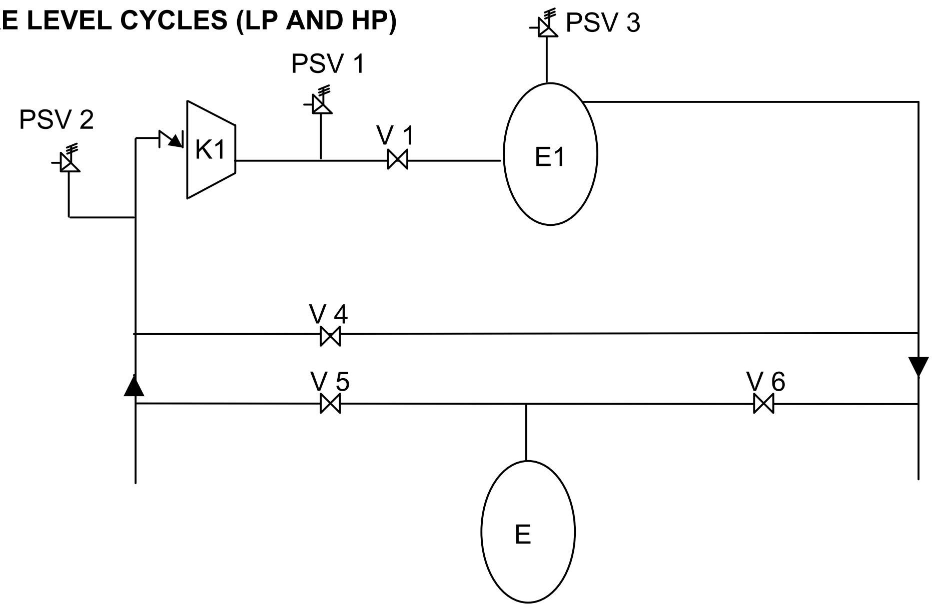


PSV40 must evacuate the flow coming from V2. Refer to 9.3.1, Pressure upstream of V2 = nominal High Pressure, Pressure downstream of V2 = PSV 40 set pressure, Cv = Cvmax V2 (valve V2 fully open).

The largest contribution must be taken into account (refer to § 9.3.1)

11.3. OIL SEPARATOR

11.3.1. TWO PRESSURE LEVEL CYCLES (LP AND HP)



It is assumed that the design pressure of the K1 and E1 are equal, and lower than the design pressure of E4.

If V1 does not exist, oil separator is protected by PSV1.

If V1 exists, and if the design pressure of both circuits (upstream and downstream of V1) are equal,

- If thermal expansion is to be considered, one should consider the risk of potential overpressure when the oil separator is isolated due to temperature increase between normal operating temperature (and normal operating pressure) and maximum outside temperature. If it is possible to reach maximum design pressure, PSV3 should be installed. PSV3 will be a standard thermal PSV.

- If fire is to be considered, every equipment that can be blocked in and possibly submitted to fire must be protected by a PSV: PSV3 is necessary and protects E1 from overpressure caused by fire. For PSV3 sizing, refer to §9.9.

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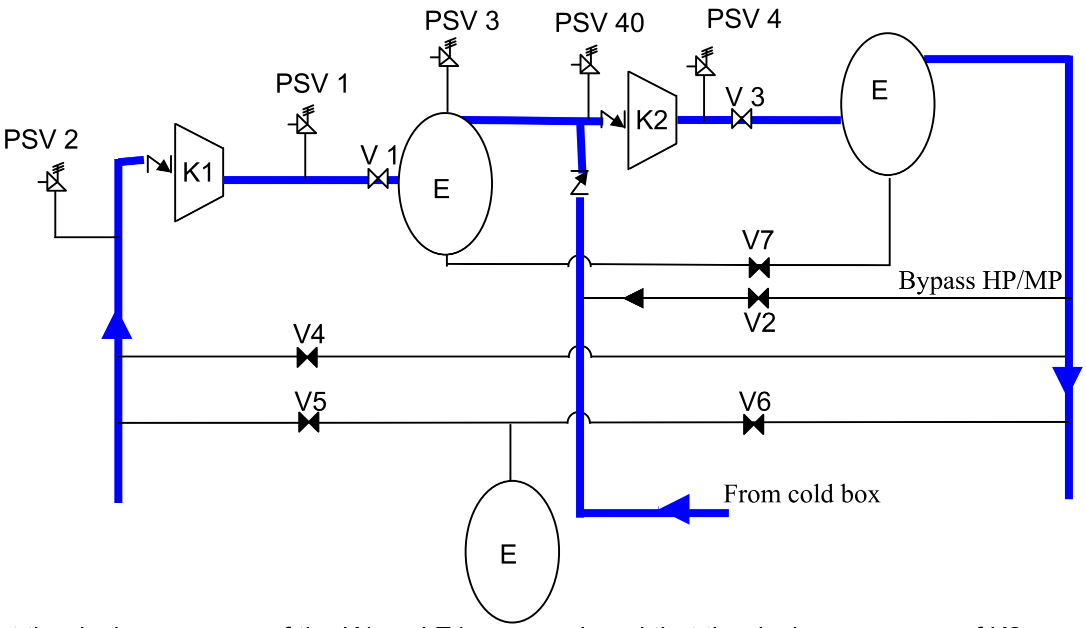
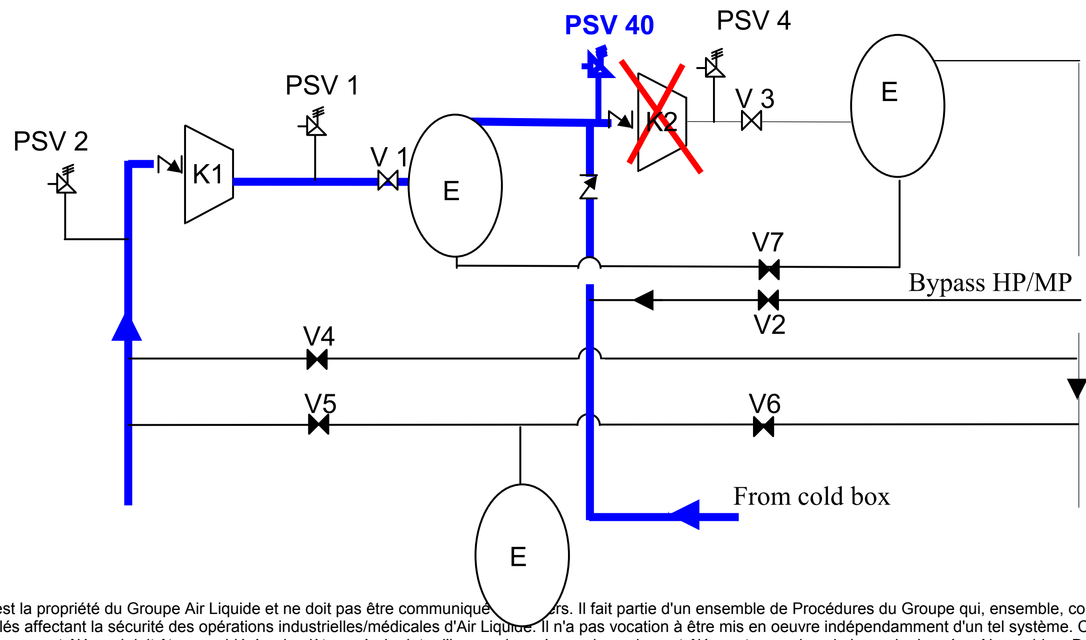
11.3.2. THREE PRESSURE LEVEL CYCLES (LP, MP AND HP)



It is assumed that the design pressure of the K1 and E1 are equal, and that the design pressures of K2, E2 and E4 are also equal and greater.

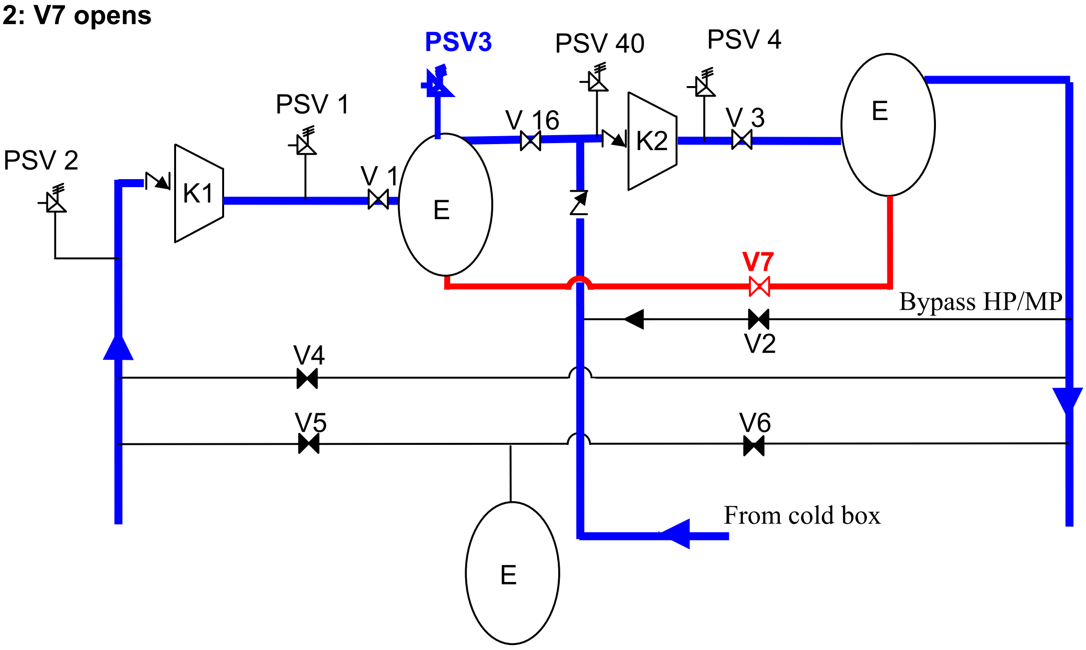
The oil separator E1 must be protected from overpressure coming from the following scenarios:

Scenario 1: Compressor K2 stops



In that case, E1 is protected from an overpressure by PSV 40 . Refer to §11.2.2, protection of K2 suction.

Scenario 2: V7 opens



PSV3 must be installed and sized to evacuate the flow due to V7 opening. (V7 opening can happen when the unit is sopped and the oil separator E1 is isolated.

Scenario 3: Blocked outlet (V16 closed)

E1 is protected through PSV1

Scenario 4 : Thermal expansion

If thermal expansion is to be considered, then one should consider the risk of potential overpressure when E1 is isolated due to temperature increase between normal operating temperature (and normal operating pressure) and maximum outside temperature. PSV3 will be a standard thermal PSV.

Scenario 5: Fire

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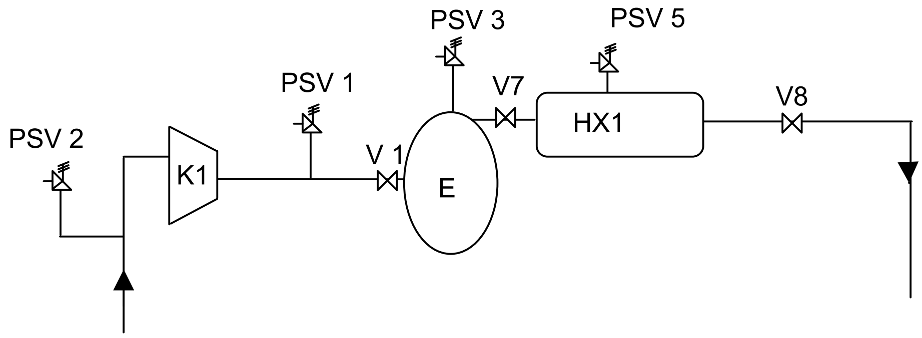
If fire is to be considered, every equipment that can be blocked in and possibly submitted to fire must be protected by a PSV: PSV3 also protects E1 from overpressure caused by fire. For PSV3 sizing, refer to §9.9.

For PSV3 sizing, the largest contribution must be taken into account.

For the E2 protection against overpressure, refer to 11.3.1.

11.4. HELIUM / WATER HEAT EXCHANGER

Refer to §10.6. See example in Appendix 3.



It is assumed that the design pressures of K1, E1, and HX1 are equal to the highest design pressure in the plant.

11.4.1. PROTECTION OF THE LP SIDE

Generally, but not always, the LP side is the water side.

LP side = water side

As a minimum, as described in §10.6, if water can be blocked in, a thermal expansion valve shall be installed.

Application of the so-called 2/3 rule: According to ASME VIII, an internal failure of shell and tubes heat exchangers has to be considered.

- If the LP side is tested at a pressure equal or above the HP side design pressure, it is considered that in case of a tube rupture, the LP side will be able to contain the high pressure. In that case there is no need to protect the LP side against a tube rupture with a flow safety valve. This 2/3 rule applies to the whole LP circuit: the whole LP circuit and not only the LP side of the heat exchanger must be tested at a pressure equal or above the HP side design pressure.

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- If the LP side is tested at a pressure lower than the HP side design pressure, the tube rupture has to be taken into account and the LP side has to be protected by a flow safety valve. Refer to §10.6.

LP side = gas side

(a) tube rupture: application of the 2/3 rule: see above.

(b) In addition to a tube rupture, the gas side must be protected against overpressure coming from the compressor.

If V1 and V2 do not exist, HX1 is protected through PSV1. PSV5 is not necessary.

If V1 and V2 exist, when V1 and V2 are open, HX1 is protected through PSV1. When V1 and V2 are closed, HX1 is isolated from the pressure source. PSV5 is not necessary.

(c) If thermal expansion is to be considered, then one should consider the risk of potential overpressure when the heat exchanger is isolated due to temperature increase between normal operating temperature (and normal operating pressure) and maximum outside temperature. If it is possible to reach maximum design pressure, PSV5 should be installed. PSV5 will be a standard thermal PSV.

(d) If fire is to be considered, every equipment that can be blocked in (V7 and V8 exist) and possibly submitted to fire must be protected by a PSV: PSV5 protects HX1 from overpressure caused by fire. For PSV5 sizing, refer to §9.9.

The largest contribution shall be taken into account to size the safety valve.

11.4.2. PROTECTION OF THE HP SIDE

HP side = gas side

(a)The gas side must be protected against overpressure coming from the compressor.

If V1 and V2 do not exist, HX1 is protected through PSV1. PSV5 is not necessary.

If V1 and V2 exist, when V1 and V2 are open, HX1 is protected through PSV1. When V1 and V2 are closed, HX1 is isolated from the pressure source. PSV5 is not necessary.

(c) If thermal expansion is to be considered, then one should consider the risk of potential overpressure when the heat exchanger is isolated due to temperature increase between normal operating temperature (and normal operating pressure) and maximum outside temperature. If it is possible to reach maximum design pressure, PSV5 should be installed. PSV5 will be a standard thermal PSV.

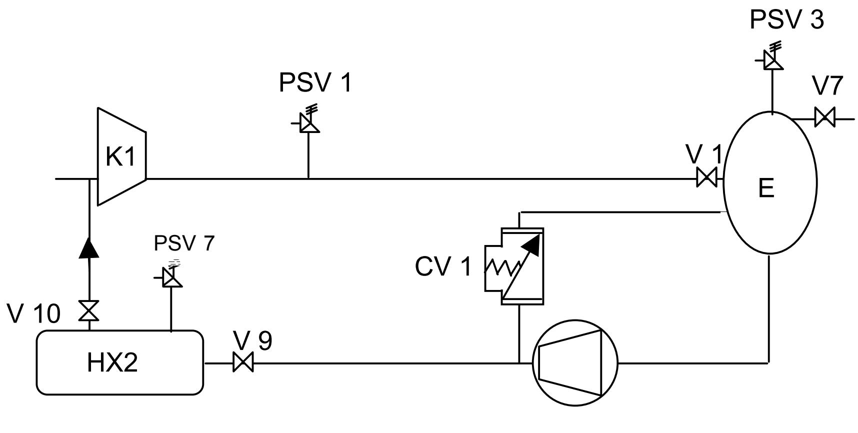
(d) If fire is to be considered, every equipment that can be blocked in (V7 and V8 exist) and possibly submitted to fire must be protected by a PSV: PSV5 protects HX1 from overpressure caused by fire. For PSV5 sizing, refer to §9.9.

The largest contribution shall be taken into account to size the safety valve.

HP side = water side: As a minimum, as described in §10.6, if water can be blocked in, a thermal expansion valve shall be installed.

11.5. OIL/WATER HEAT EXCHANGER

Refer to §10.6, See example in Appendix 3.



It is assumed that the design pressure of the K1, E1, HX2 are equal.

11.5.1. PROTECTION OF THE LP SIDE

Generally, but not always, the LP side is the water side.

LP side = water side

As a minimum, as described in §10.6, if water can be blocked in, a thermal expansion valve shall be installed.

Application of the so-called 2/3 rule: According to ASME VIII, an internal failure of shell and tubes heat exchangers has to be considered.

- If the LP side is tested at a pressure equal or above the HP side design pressure, it is considered that in case of a tube rupture, the LP side will be able to contain the high pressure. In that case there is no need to protect the LP side against a tube rupture with a flow safety valve. This 2/3 rule applies to the whole LP circuit: the whole LP circuit and not only the LP side of the heat exchanger must be tested at a pressure equal or above the HP side design pressure.

- If the LP side is tested at a pressure lower than the HP side design pressure, the tube rupture has to be taken into account and the LP side has to be protected by a flow safety valve. Refer to §10.6.

<u>LP side = oil side</u>

(a) tube rupture: application of the 2/3 rule: see above.

(b) If the heat exchanger can be isolated (V9 and V10 exist), hydraulic expansion has to be taken into account (refer to 9.12), a thermal expansion valve shall be installed.

(c) If fire is to be considered, every equipment that can be blocked in (V7 and V8 exist) and possibly submitted to fire must be protected by a PSV: PSV7 protects HX2 from overpressure caused by fire. For PSV7sizing, refer to §9.9.

The largest contribution shall be taken into account to size the safety valve.

The oil side of heat exchanger HX2 is protected again a pressure increase due to the oil pump through CV1.

11.5.2. PROTECTION OF THE HP SIDE

<u>HP side = oil side</u>

(a) If the heat exchanger can be isolated (V9 and V10 exist), hydraulic expansion has to be taken into account (refer to 9.12), a thermal expansion valve shall be installed

(b) If fire is to be considered, every equipment that can be blocked in (V7 and V8 exist) and possibly submitted to fire must be protected by a PSV: PSV7 protects HX2 from overpressure caused by fire. For PSV7 sizing, refer to §9.9.

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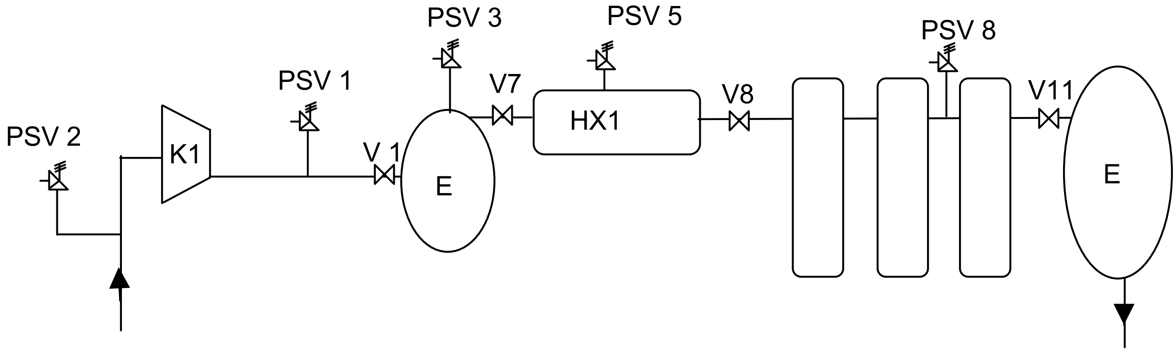
The largest contribution shall be taken into account to size the safety valve.

The oil side of heat exchanger HX2 is protected again a pressure increase due to the oil pump through CV1.

<u>HP side = water side:</u> As a minimum, as described in §10.6, if water can be blocked in, a thermal expansion valve shall be installed.

11.6. OIL REMOVAL SYSTEM: COALESCORS



It is assumed that K1, E1, HX1, the coalescors and E3 have the same design pressure.

If the coalescors cannot be isolated from the compressor, they are protected against an overpressure by PSV1.

If the coalescors can be isolated (V1, V7 or V8 exist, and V11 exists either upstream or downstream of E3)

- If thermal expansion is to be considered, one should consider the risk of potential overpressure when the coalescors are isolated due to temperature increase between normal operating temperature (and normal operating pressure) and maximum outside temperature. If it is possible to reach maximum design pressure, PSV8 should be installed. PSV8 will be a standard thermal PSV.

- If fire is to be considered, every equipment that can be blocked in and possibly submitted to fire must be protected by a PSV: PSV8 is necessary and protects the coalescors from overpressure caused by fire. For PSV8 sizing, refer to §9.9

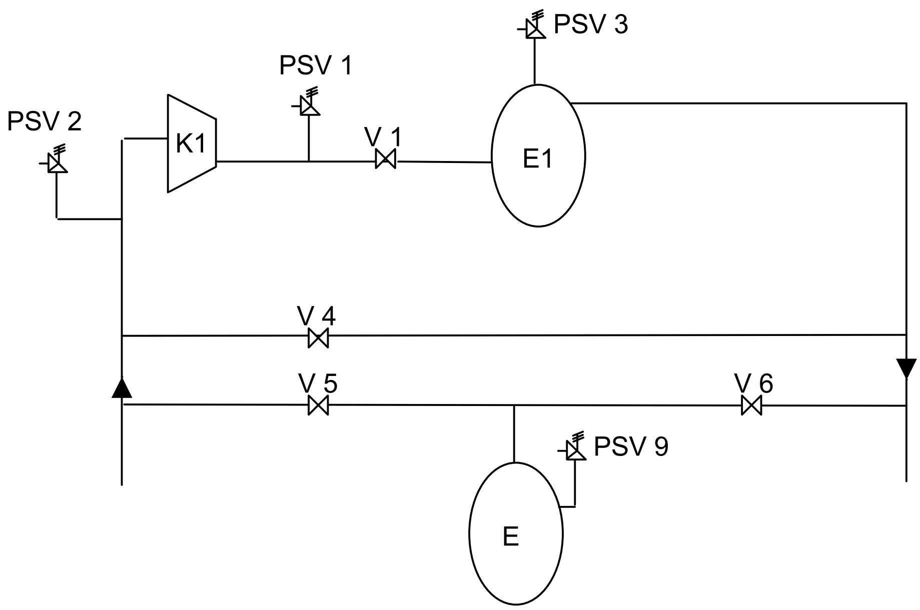
11.7. OIL REMOVAL SYSTEM: CHARCOAL ADSORBER

Refer to 11.6

The charcoal adsorber is protected in the same way as the coalescors.

It must be checked that the initial drying of the charcoal will not bring about other causes of overpressure. If so, these situations must be taken into account.

11.8. HELIUM BUFFER AND OTHER PRESSURE VESSELS



It is assumed that the design pressures of K1, E1 and E4 are equal to the highest design pressure in the plant .

For the helium buffer and for other equipments operating at room temperature, which can be isolated (i.e. filters), following has to be considered:

- If thermal expansion is to be considered, one should consider the risk of potential overpressure when the buffer is isolated due to temperature increase between normal operating temperature (and normal operating pressure) and maximum outside temperature. If it is possible to reach maximum design pressure, PSV 9 should be installed. PSV 9 will be a standard thermal PSV.

- If fire is to be considered, every equipment that can be blocked in and possibly submitted to fire must be protected by a PSV: PSV 9 is necessary and protects the helium buffer E4 from overpressure caused by fire. For PSV 9 sizing, refer to §9.9

It must be checked that the re-filling of the helium buffer will not bring about other causes of overpressure. If so, these situations must be taken into account.

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11.9. SUPER INSULATED PLATE FIN HEAT EXCHANGERS

Refer to 10.1.

The heat exchanger surface is to be calculated as follows:

$$Surface\quad HPcircuit\quad = \quad External\quad Overall\quad Surface$$

$$Surface\quad LPcircuit\quad = \quad External\quad Overall\quad Surface$$

As the heat exchangers are made of aluminum, which is a highly conductive material, the overall external surface must be considered when calculating the heat input to the exchangers.

11.10. SUPER INSULATED LINES

Refer to 10.1.

11.11. SUPER INSULATED ADSORBERS

Refer to 10.1.

11.12. NON SUPER INSULATED SURFACES OR THIN SUPER INSULATION UNDER VACUUM

11.12.1. FLUID TEMPERATURE BELOW 77K

On non super-insulated surfaces operating at low temperatures (T < 77K), in case of vacuum loss, air condensation will occur. This phenomenon will enhance the heat transfer coefficient and result in an increased heat input to the system.

The following heat inputs must be considered in that case (Refer to [31]).

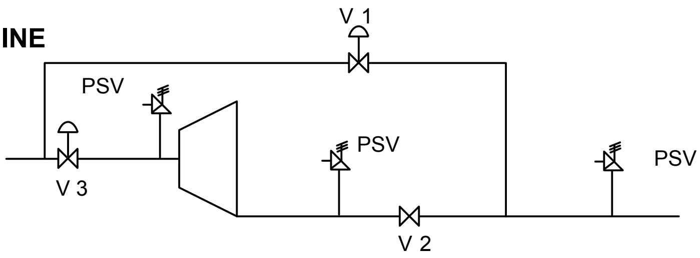
Operating Temperature (K)	Heat input to the system Q (W/m²)
4	38 000
20	36 000
77	28 000

11.12.2. FLUID TEMPERATURE BETWEEN 77K AND 300K

Air condensation will not occur.

11.13. TURBINE

11.13.1. CLAUDE TURBINE



If block valve V 2 doesn't exist, PSV 10 is not required and PSV 20 is designed for the most constraining of the two following cases :

- total opening of bypass valve V 1 (calculation per § 9.3.1)
- turbine flow with suction and discharge pressures equal to set pressures.

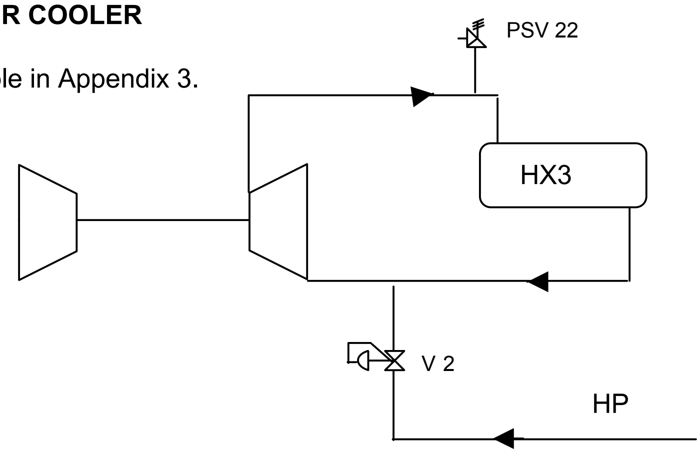
If valve V 2 exists (for maintenance or for turbine deriming) and design pressure at discharge up to V 2 is the same as suction side design pressure then this is equivalent to first case.

If valve V 2 exists (for maintenance or for turbine deriming) and design pressure at discharge is lower than suction side design pressure then PSV 10 is designed for turbine flow with suction and discharge pressures equal to set pressures and PSV 20 is designed for total opening of bypass valve V 1.

In the helium refrigeration/liquefaction facilities, the claude turbine is vacuum insulated. The turbine is not a pressure vessel, but as for all vacuum insulated piping, a vacuum break must be taken into account (refer to 10.1). PSV21 is a standard thermal PSV.

11.13.2. TURBINE WATER COOLER

Refer to §10.6. See example in Appendix 3.



11.13.3. PROTECTION OF THE LP SIDE

The LP side is the water side.

As a minimum, as described in §10.6, if water can be blocked in, a thermal expansion valve shall be installed.

Application of the so-called 2/3 rule: According to ASME VIII, an internal failure of shell and tubes heat exchangers has to be considered.

- If the LP side is tested at a pressure equal or above the HP side design pressure, it is considered that in case of a tube rupture, the LP side will be able to contain the high pressure. In that case there is no need to protect the LP side against a tube rupture with a flow safety valve. This 2/3 rule applies to the whole LP circuit: the whole LP circuit and not only the LP side of the heat exchanger must be tested at a pressure equal or above the HP side design pressure.

- If the LP side is tested at a pressure lower than the HP side design pressure, the tube rupture has to be taken into account and the LP side has to be protected by a flow safety valve. Refer to §10.6.

11.13.4. PROTECTION OF THE HP SIDE = GAS SIDE

(a)The gas side must be protected against overpressure coming from the brake compressor. The suction side of the brake compressor can be as high as the high pressure level, in case of V2 failure. As a consequence, the pressure on the discharge side of the compressor could be even higher. PSV 22 protects HX3 against an overpressure in case of V2 failure. The flow to be evacuated is the flow through the compressor with Pdischarge = PSV22 set pressure, at nominal speed.

(b) HX3 is protected against thermal expansion through PSV 21.

(c) If fire is to be considered, every equipment that can be blocked in and possibly submitted to fire must be protected by a PSV: PSV 22 protects HX3 from overpressure caused by fire. For PSV 22 sizing, refer to §9.9.

The largest contribution (among (a) and (b) shall be taken into account to size the safety valve PSV 22.

11.14. CRYOGENIC CENTRIFUGAL PUMPS AND COMPRESSORS (SHE, GHE)

In most cases those pumps are not “transfer” pumps, but circulators. They ensure a certain flow as they balance the pressure drop in a closed loop.

11.14.1. SUCTION SIDE PROTECTION

If the pressure drop across the pump is low (< 500mb), and if the fluid volume in the loop is small, then the back flow through the pump caused by a pump stop can be neglected.

If not, then refer to § 9.6.1.1. for protection against back flow.

11.14.2. DISCHARGE SIDE PROTECTION

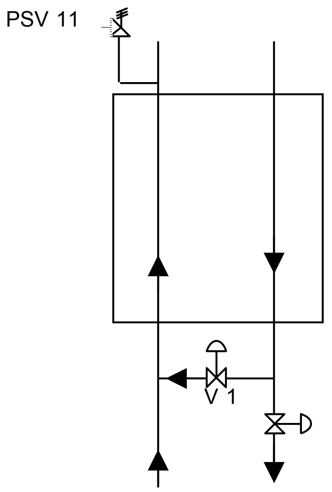
In the case of reciprocating pump, a PSV is installed at the discharge of the pump. It is sized for nominal pump flow rate.

11.15. CRYOGENIC STORAGE

Refer to Process Design Rules A0043 : process calculation on cryogenic storage tank for non vacuum insulated tanks.

Refer to 9.13 for vacuum insulated tanks.

11.16. START UP BYPASS



The LP line must be protected against overpressure in case of start-up bypass opening (V1). Refer to 9.3.

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PSV 11 must be able to evacuate the flow through V1 fully open, with V1 upstream conditions = nominal conditions (operation, cool down or warm up of the cold box, whichever gives the worst scenario) and V1 downstream conditions = PSV set pressure.

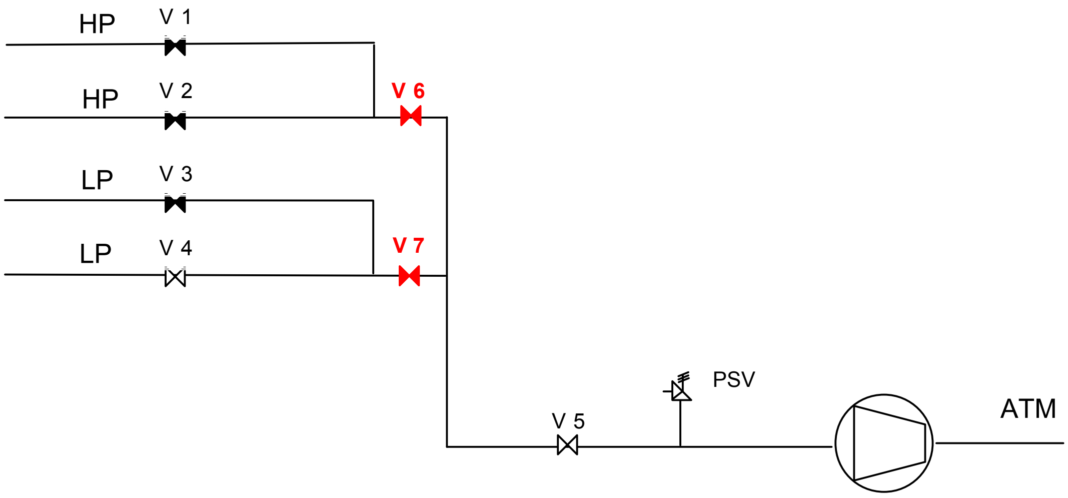
Note: cool down phases must be taken into account when sizing PSV 11

11.17. .

11.18. VACUUM PUMPS

Refer to 10.7

11.18.1. PART OF THE CIRCUITS CAN BE PURGED WHILE THE UNIT IS UNDER PRESSURE



This is the case on large refrigeration units, where the cold box can be purged while the compressor station is running, or where one of the turbo-expanders can be purged while the rest of the unit is running.

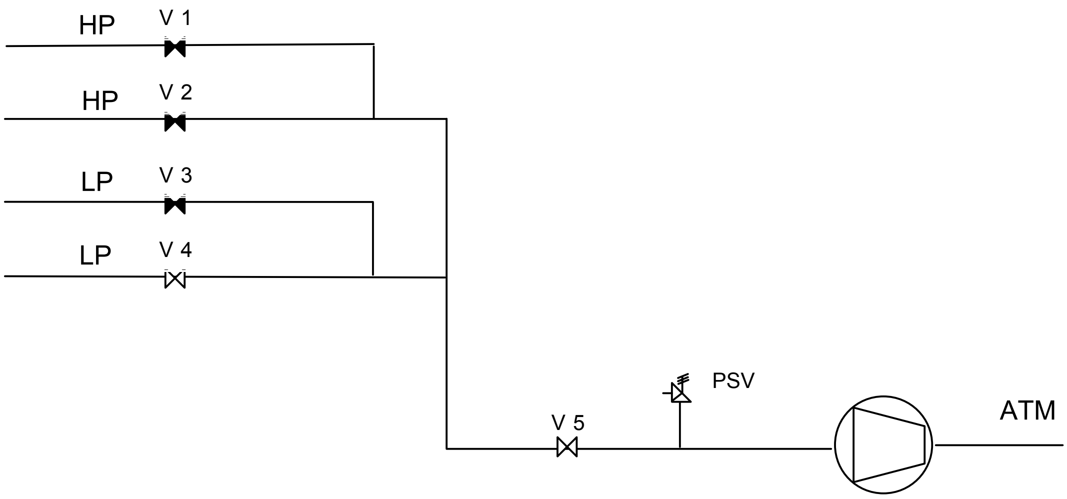
In that case, when a LP circuit is being purged (V4 and V5 are open), the opening of one valve (V1 or V2) leads to an overpressure in the purging circuit and in the “LP circuit”.

PSV12 would be far too large to account for this incident. An additional isolation valve is put in place (V6) so that the incident is only possible by the opening of two valves (V1 and V6 or V2 and V6).

Likewise , when a HP circuit is being purged (V1, V6 and V5 are open), the opening of one valve (V3 or V4) leads to an overpressure in the purging circuit. If this overpressure cannot be handled by the purging circuit, a supplementary isolation valve, V7, is installed.

PSV 12 is a standard leak tight thermal PSV which protects the pump from a slight overpressure (due to a valve leakage or to thermal expansion of blocked in gas). PSV 11 is approximately set at 0.2 barg.

11.18.2. THE CIRCUITS ARE PURGED WHILE THE UNIT IS DEPRESSURIZED



This is the case on small refrigeration units (HELIAL), and on fully automatic units, where the cold box cannot be purged while the compressor station is running, or where no turbo-expander can be purged while the rest of the unit is running.

The purging circuit is used when the HP and LP circuits are depressurized: there is no risk to overpressure neither the purging circuit nor the LP circuit.

PSV 12 is a standard leak tight thermal PSV which protects the pump from a slight overpressure (due to a valve leakage or to thermal expansion of blocked in gas). PSV 12 is approximately set at 0.2 barg.

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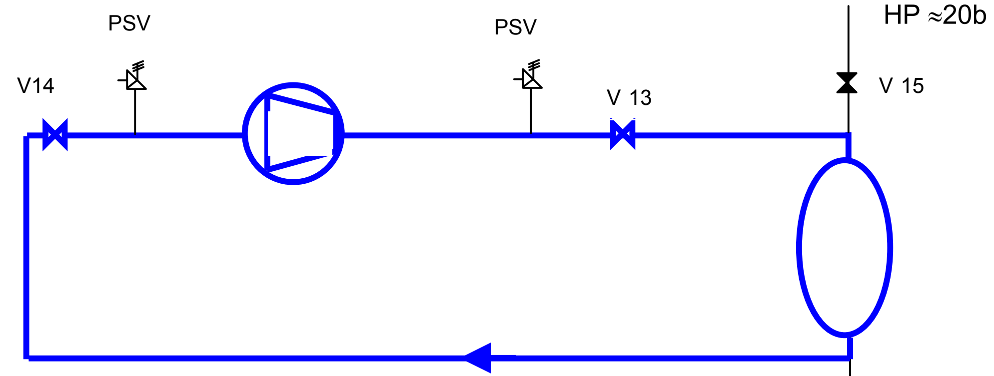
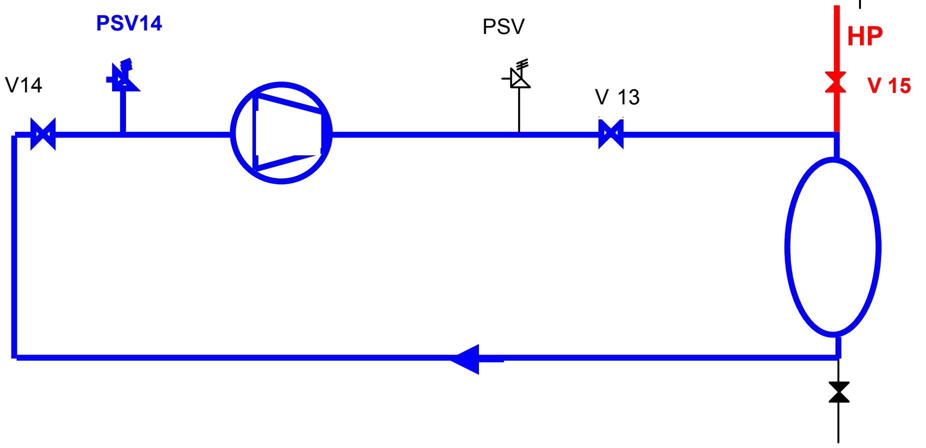
11.19. ROOTS COMPRESSORS

The roots compressor generally operates between 1 and 2 bars abs.



11.19.1. SUCTION SIDE



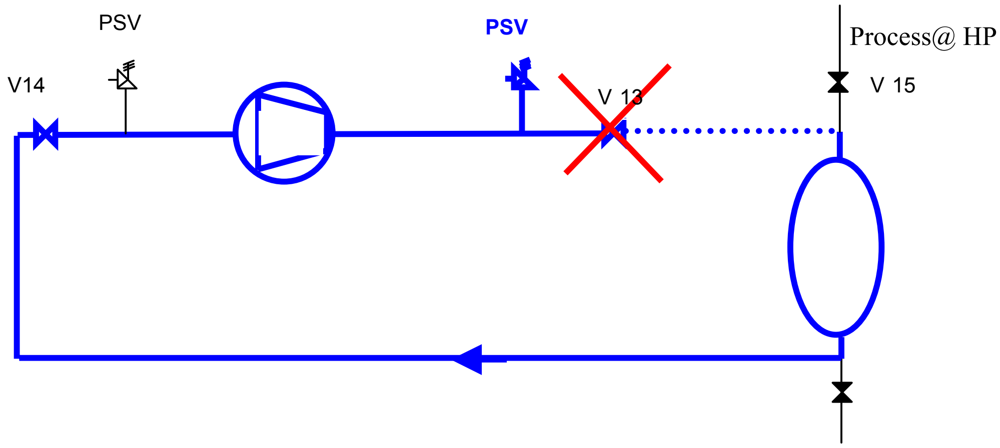
PSV14 protects K3 from an overpressure coming from the upstream circuit.

The scenario to be taken into account is that V15 opens when K3 is running. The regeneration loop is then in communication with the cycle high pressure.

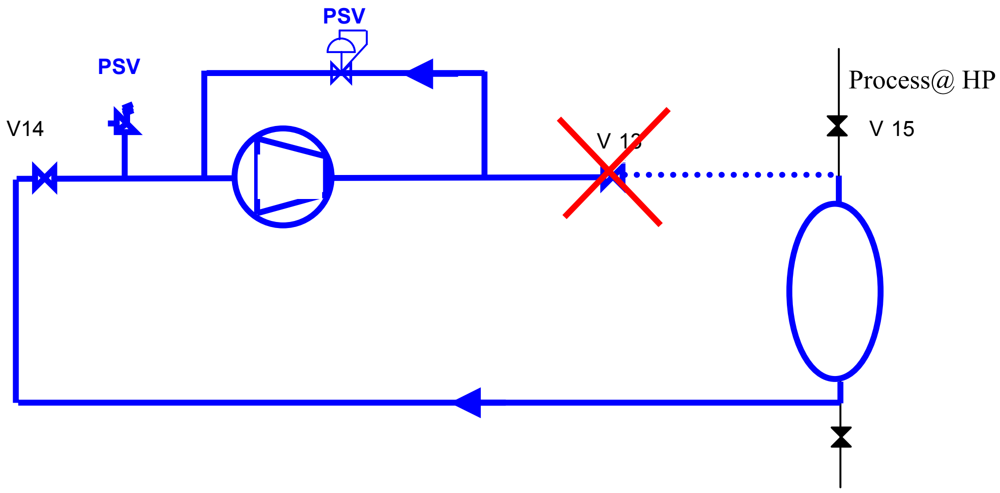
PSV14 must evacuate the flow through the upstream circuit, at set pressure, temperature is either 80K or 20K (the worst case must be considered)

11.19.2. DISCHARGE SIDE

The scenario to be considered is a blocked outlet . PSV13 protects the compressor against an overpressure caused by V13 closed. Refer to 9.2.1.



Note



K3 can also be protected against an overpressure caused by a blocked outlet with a balanced safety valve (PSV15).

PSV15 is set to allow an admissible DP.

PSV15 must be able to flow the flow-rate of the roots compressor, calculated with suction pressure = PSV14 set pressure and discharge pressure = PSV14 set pressure + admissible DP.

11.20. JACKET RELIEF DEVICE

The jackets of containers shall be protected by a suitable pressure relief device to release internal pressure. This relief device shall function at a pressure not exceeding the design pressure of the jacket calculated in accordance with the applicable code, maximum external collapse pressure on the inner vessel calculated with a safety factor of two or 1.72 b, whichever is less.

The total discharge area of vacuum jacket relief devices on container shall be at least 0.3414mm²/kg of water capacity of the container.

$$\text{Set P of the relief device} \geq \text{MIN} \left\{ \begin{array}{l} \text{Jacket internal design P} \\ \text{Cold box equipments external design P x 2} \\ \text{1.72 bars abs – 1.013 bars abs = 0.707b} \end{array} \right\}$$

DTA recommendation is :	- Jacket internal design P = 0.5b	($\Delta P_{in\text{-}out}$ = 0.5b)
	- Cold box equipments external design P = 1.5b	($\Delta P_{in\text{-}out}$ = -1.5b)

11.21. ROUTING OF THE SAFETY VALVE DISCHARGE

The following paragraphs concern helium refrigeration / liquefaction units. DTA standard is not to collect safety valves which are meant to discharge inert gases.

However, it is to be noted that in case of toxic or flammable gases, this philosophy does no longer apply.

11.21.1. HIGH OIL CONTENT IN THE FLUID TO BE DISCHARGED

This concerns most PSVs located in the compressor station (PSV1, 3, 4, 5, 7, 8), upstream of the oil removal system.

DTA recommendation is to route these safety valves discharge to a safe location, to a containment tub. The discharge of oil or oily gas in the compressor's area would endanger the personnel safety (burns, slipping).

11.21.2. FLUID TO BE DISCHARGED = AMBIENT TEMPERATURE HELIUM

This concerns PSVs located in the compressor station, downstream of the oil removal system (PSV2, PSV9)

DTA recommendation is not to route these safety valves discharge, but to ensure that the safety valves will not discharge on a walk path. It is generally not of DTA responsibility to provide all the necessary means (i.e. extraction fan, oxygen content detector) to avoid anoxia risk.

11.21.3. FLUID TO BE DISCHARGED = COLD HELIUM (T< 273K)

This concerns most PSVs located on the cold box warm panel.

DTA recommendation is not to route these safety valves discharge, but to ensure that the safety valves will not discharge on a walk way, or into a maintenance area, or onto carbon steel (typically, the cold box shell)

It is generally not of DTA responsibility to provide all the necessary means (i.e. extraction fan, oxygen content detector) to avoid anoxia risk.

11.21.4. FLUID TO BE DISCHARGED = LN2

DTA recommendation is to route these safety valves discharge to a safe location, outdoors, to avoid the risk of anoxia and risk of cold burns.

11.21.5. WATER SAFETY VALVES

DTA recommendation is not to route these safety valves discharge. Water draining is generally not supplied by DTA.

11.22. ISOLABLE SAFETY VALVES

DTA recommendation is to avoid to install isolable safety valves.

In some situations, when a customer requests the possibility to isolate a safety valve for maintenance, it can be tolerated that the safety valves are isolated. In such situations, isolation valves can be accepted provided trained personnel and adequate procedures are put in place.

If so, then the isolation valve upstream (and downstream of the PSV, if the PSV discharge is routed to a header) must be CSO (car seal open), and located very close to the safety valve. A pressure indicator must also be located close to the safety valve and must be watched during the maintenance phase.

Flow safety valves will be redundant, so that there will always be a PSV to protect the equipment while the second PSV is being maintained.

11.23. THERMAL ISOLATION OF UPSTREAM AND DOWNSTREAM PIPING

The piping located upstream of PSV (from cold box to safety valve) and downstream of PSV will not be thermally isolated.

APPENDICES

1. Calculation of the flow capacity of the safety valve which protects the whole HP line on a HELIAL 1000 in case of vaccum loss
2. Maximum pipe length to be protected by a standard thermal PSV.
3. Protection of a watercooler

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APPENDIX 1

Calculation of the flow capacity of the safety valve which protects the whole HP line on a HELIAL 1000.

This safety valve protects the heat exchangers (operating temperatures from 300K to 5K) and the 80K adsorber.

1. DEFINE THE SET PRESSURE AND FLOWRATING PRESSURE

Design Pressure = 20 b

MAWP = 19 barg (the heat exchangers are vacuum insulated)

According to §7.3.4: Set pressure = 19 barg

Prelief = 110% MAWP = 110% (19 barg) + 1.013 = 21.9 bars abs

2. CALCULATE THE SURFACES

Heat exchanger 1 / 2

Width: 230mm, Length:1750mm, Stacking height: 238mm, Nb of HP passages :15, Overall nb of passages:31

Insulation thickness: 10 layers, i.e. 5mm

According to §9.13.2.1 , the protected surface Σ is the arithmetic mean value of the insulating material inner and outer surfaces (m²)

Σ inner =(0,23*0,238*2+1,75*0,238*2+1,75*0,23*2)

Σ outer = [(0,23+0.005*2)*(0,238+0.005*2)*2+(1,75+0.005*2)*(0,238+0.005*2)*2+(1,75+0.005*2)*(0,23+0.005*2)*2]

Σ = 1.8m²

Heat exchanger 3 / 4

Width: 200, Length: 1800, Stacking height: 206, Nb of HP passages : 4, Overall nb of passages: 25

Insulation thickness: 10 layers, i.e. 10mm

Σ = 1.6m²

Heat exchanger 5 / 6

Width: 170, Length: 1850, Stacking height: 137, Nb of HP passages : 8, Overall nb of passages:17

Insulation thickness: 10 layers, i.e. 15mm

Σ =1.3 m²

80K Adsorber

Outside diameter : 168.3 mm, Height: 731 mm, Insulation thickness: 10 layers, i.e. 10mm

$$\Sigma \text{ inner} = \pi *0.1683*0.731 + 2*\pi \frac{0.1683^2}{4} *1.41 = 0.45m^2$$

Σ outer = π (0.1683+0.01*2)*(0.731 +2*0.01)+ 2*π $\dfrac{(0.1683 + 0.01 * 2)^2}{4}$ *1.41 = 0.52m²

Σ = 0.49m²

3. CALCULATE THE RELIEF TEMPERATURE

Heat exchanger 1 / 2

The relief pressure is greater than the critical pressure.

According to §9.13.4.3, the relief temperature is such that $\dfrac{\sqrt{\nu}}{\nu.\left(\dfrac{\partial \mathrm{h}}{\partial \nu}\right)_p}$ is maximum and T is within the operating range of the protected equipment.

Define the minimum process temperature: 80K, but as the safety valve protects the whole HP line which minimum temperature is 5K, the minimum process temperature is taken as 5K for the whole line.

Relief temperature = 12.26K

Note: if HX1/2 had its own safety valve, the min process temperature and as a consequence the relief temperature would have been 80K.

Heat exchanger 3 / 4, Heat exchanger 5 / 6, 80K Adsorber

As explained above, the minimum process temperature is taken as 5K for the whole line.

Relief temperature = 12.26K

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4. DETERMINE THE HEAT TRANSFER COEFFICIENT AND HEAT INPUT

DEFINE THE AMBIENT TEMPERATURE
Ta is taken as 45°C = 318K

DETERMINE U2 AND W2

HEAT EXCHANGER 1 / 2
Mean temperature between T and Ta = (T+Ta)/2 = (12.26+318)/2 = 165K
λ air at 165K, 1atm = 0.058 W/m/K
λ helium at 165K, 1atm = 0.103 W/m/K
λ_2 = max (λ air, λ helium) = 0.103 W/m/K

$$U_2 = \frac{\lambda_2}{e}$$

U_2 = 0.103 /0.005 = 20.7 W/m²/K
W_2 = (318-12.26)*20.7*1.8 = 11364 W

Heat exchanger 3 / 4
U_2 = 0.103 /0.01 = 10.4 W/m²/K
W_2 = (318-12.26)*10.4*1.6 = 5177 W

Heat exchanger 5 / 6
U2 = 0.103 /0.015 = 6.9 W/m²/K
W2 = (318-12.26)*6.9*1.3 = 2776 W

80K Adsorber
U_2 = 0.103 /0.010 = 10.4 W/m²/K
W_2 = (318-12.26)*10.4*0.49 = 1539 W

Overall heat input
W = 11364 + 5177 + 2776 + 1539 = 20857
In our case, the heat input through piping and supporting elements is neglected.

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5. DETERMINE THE FLOW TO BE EVACUATED

The relief pressure is above the critical pressure.

According to § 9.13.4.3

$$\boxed{Q_m = 3.6 * \frac{W}{L'}.}$$

L' = 69.4 kJ/kg, at 12.26K, P = 21.9 bars abs
Qm = 1081 kg/h

6. DETERMINE THE INDICATIVE DIAMETER OF THE SAFETY VALVE

The safety valve shall be specified as follows:

Fluid to be relieved = helium

Set pressure = 19 barg

Flow-rating pressure = 21.9 bars abs

Relief temperature = 12.26K, even though the equipment proeted is a set of heat exchangers, with a relief valve located on the warm end of the heat exchangers.

Helium flow-rate = 1081 kg/h

The diameter is calculated by the supplier and can be checked by DTA.

According to §8.3.2

Eq 1 $A = 1.316 \frac{W}{CK_d P_1 K_b} \sqrt{\frac{TZ}{M}}$

C = 378 (kstd = Cpstd /Cv std = 1.67)

Kd is taken as 0.509.

This is a very low coefficient, but applicable for small safety valves known as thermal expansion valves.

$$A = 1.316 \frac{1081}{378 * 0.509 * 21.9 * 1} \sqrt{\frac{12.26 * 0.935}{4}} = 0.57\ cm^2$$

D = 8.5mm

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APPENDIX 2

This appendix gives the maximum length which can be protected with a standard thermal safety valve (1/2" - 1/2"), depending on the set pressure, temperature level, and pipe diameter.

Pipe length to be proptected by a thermal safety valve (1/2" - 1/2"), depending on pipe diameter and insulation thickness, Set Pressure = 4 barg

HELIUM		
Molecular weight	4,003	g/mol
Ta	300	K
Pc	2,275	bars abs
MAWP	4	barg
Prelief	110	% MAWP
Prelief	5,4	bars abs (110 % MAWP)+1,01325

Tprocess min	Insulation thickness	Maximum pipe length (m) which can be protected by a standard thermal PSV 1/2" - 1/2", depending on pipe diameter and temperature level													D
K	mm	DN8	DN10	DN15	DN20	DN25	DN32	DN40	DN50	DN65	DN80	DN100	DN125	DN200	cm
4	15	25	22	19	17	15	12	11	9	8	7	5	4	3	1,00
20	10	38	33	29	24	21	17	15	13	11	9	7	5	4	1,00
80	5	55	46	39	32	27	22	19	16	13	11	9	6	5	1,00

Example: For a design pressure of 5 bar (MAWP = 4barg), a DN50 pipe which flows helium à 4K must be protected by a standard PSV every 9m

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Pipe length to be proptected by a thermal safety valve (1/2" - 1/2"), depending on pipe diameter and insulation thickness, Set Pressure = 9 barg

HELIUM		
Molecular weight	4,003	g/mol
Ta	300	K
Pc	2,275	bars abs
MAWP	9	barg
Prelief	110	% MAWP
Prelief	10,9	bars abs (110 % MAWP)+1,01325

Tprocess min	Insulation thickness	Maximum pipe length (m) which can be protected by a standard thermal PSV 1/2" - 1/2", depending on pipe diameter and temperature level													D
K	mm	DN8	DN10	DN15	DN20	DN25	DN32	DN40	DN50	DN65	DN80	DN100	DN125	DN200	cm
4	15	64	57	50	44	38	32	29	24	21	18	14	10	8	1,00
20	10	76	66	58	49	42	35	31	26	22	18	15	10	8	1,00
80	5	111	94	79	65	55	44	39	32	27	22	17	12	9	1,00

Example: For a design pressure of 10 bar (MAWP = 9barg), a DN50 pipe which flows helium à 4K must be protected by a standard PSV every 24m

Pipe length to be proptected by a thermal safety valve (1/2" - 1/2"), depending on pipe diameter and insulation thickness, Set Pressure = 19 barg

HELIUM		
Molecular weight	4,003	g/mol
Ta	300	K
Pc	2,275	bars abs
MAWP	19	barg
Prelief	110	% MAWP
Prelief	21,9	bars abs (110 % MAWP)+1,01325

Tprocess min	Insulation thickness	Maximum pipe length (m) which can be protected by a standard thermal PSV 1/2" - 1/2", depending on pipe diameter and temperature level													D
K	mm	DN8	DN10	DN15	DN20	DN25	DN32	DN40	DN50	DN65	DN80	DN100	DN125	DN200	cm
4	15	165	148	131	113	99	83	75	63	54	46	37	26	20	1,00
20	10	158	138	120	101	87	72	65	54	45	38	30	21	16	1,00
80	5	225	190	160	132	111	90	79	65	54	45	35	24	19	1,00

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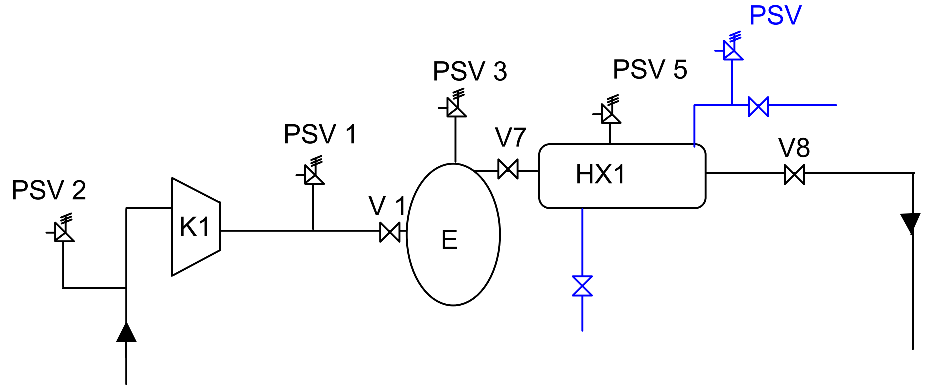
Example: For a design pressure of 20 bar (MAWP = 19barg), a DN50 pipe which flows helium à 4K must be protected by a standard PSV every 63m

APPENDIX 3

Protection of a watercooler



LP side = water-side, Design pressure = 10 barg, Test pressure = 1.3 x10 barg =13 barg

Tubes internal diameter = 19.6 mm

HP side = gas side = shell side, Design pressure = 30 barg.

HP side normal operating pressure = 25 barg = 26.01 bars abs

HP side normal operating temperature = 20°C

Extreme ambient temperature: 4°C / +50°C

Protection of the HP side:

1. Overpressure caused by K1

The heat exchanger HX1 is protected against an overpressure coming from the compressor through an upstream PSV (PSV1), refer to 11.4.2.

2. Fire is considered as an unlikely contingency.

3.. Thermal expansion is not significant:

Initial state Pgas side = 25 barg = 26.01 bars abs

T gas side = 4°C

ρHe = 4.462 kg/m3 (HYSYS MBWR)

Final state Tgas side = 50°C, ρHe = 4.462 kg/m3 $\Rightarrow$ Pgas side = 30.3 bars abs

Pgas side < Pdesign

No safety valve is necessary on HX1 to protect the HP side (PSV 5 is not necessrey)

Protection of the LP side:

1. Thermal expansion

As water can be blocked in, the LP side must be at minimum equipped with a thermal safety valve : PSV50 is necessary.

2. In case of tube rupture, the HP side (helium) will flow into the LP circuit (refer to § 11.4.1)

LP side test pressure (13 barg) < HP side design pressure (30 barg)

The LP side is not tested at a pressure equal or greater than the HP side design pressure. The LP side must be protected by a flow safety valve. This safety valve is sized as follows:

Refer to § 10.6 and 9.8.2 .

Orifice = 2 tube section = 2 Π (19.6)²/4 = 603 mm²

Upstream pressure= normal operating HP pressure = P_1 = 25 barg = 26.01 bars abs

Downstream pressure = LP set pressure = P_2= 10 barg = 11.01 bars abs

r = P_2/P_1 = 11.01 / 26.01 = 0.42

k = Cp / Cv = 1.664

$$\left(\frac{2}{k+1}\right)^{\frac{k}{k-1}} = 0.49$$

$r \le \left(\frac{2}{k+1}\right)^{\frac{k}{k-1}}$, critical flow, use Equation 7 $W = CA\sqrt{1.296kP_1\rho_1\left(\frac{2}{k+1}\right)^{\frac{k+1}{k-1}}}$, W = 4701 kg/h

According Table 1 of document W-EP-3-6-1, as the cooling water circuit is a closed circuit, as water flow is on the tube side, as helium is not a dangerous gas, whatever the size of the heat exchanger, the level of protection is defined as 1, i. e. the relief valve is calculated in gas and installed on water pipe close downstream of cooler.

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When calculating the safety valve size with a helium flow, equation 1 must be used:

LP set pressure = 10 barg

LP relief pressure = 110% 10barg = 12.01 bars abs

k	1,67	
P1	12,01	bars abs
P2	1	bar abs
r =P2/P1	0,083	

$$\left(\frac{2}{k+1}\right)^{\frac{k}{k-1}} \quad 0{,}49$$

CRITICAL FLOW

Eq 1 $$A = 1.316 \frac{W}{CK_d P_1 K_b} \sqrt{\frac{TZ}{M}}$$

kst	1,67	

$$C = 520 \sqrt{k_{st} \left(\frac{2}{k_{st}+1}\right)^{\frac{k_{st}+1}{k_{st}-1}}}$$

C	378	
Kd	0,9	
Kb	1	
T	293	K
Z	1,01	
M	4	g/mol
A	13	cm²
D	4	cm

Note 1 If the LP side design pressure is 20 barg, with a test pressure of 26 barg, the “2/3” rules applies:

The LP is tested at a pressure greater (26 barg) than the HP design pressure (25 barg).

In that case, the LP side does not need to be equipped with a flow safety valve. Nevertheless, if water can be blocked in, a thermal PSV must be put in place.

Note 2 With semi-open or open circuit, it is not necessary to provide protection on the water circuit because it is assumed that the water will be discharged at the cooling towers before the opening of a relief valve. Refer to document W-EP-3-6-1 (see § 10.6)

Nevertheless, if water can be blocked in, a thermal PSV must be put in place.

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