



Training Services

Pressure Relief Systems I

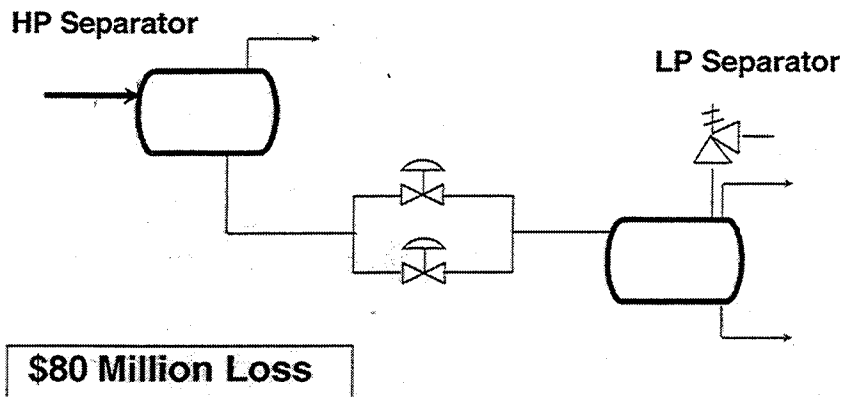
IPD-2004/RS-I-1

Day 1 Outline

- PRV's and Process Safety
- PRV's: What and Where
- Rules and Procedures
- Pressure Definitions
- Identifying Overpressure Cases

IPD-2004/RS-I-2

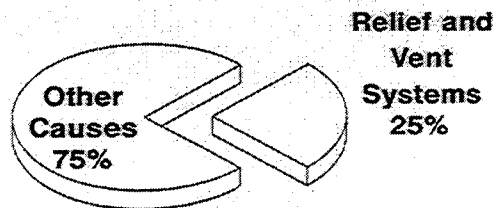
A Major Incident UK Hydrocracker



SCOTLAND

IPD-2004/RS-1-3

Process Industry Loss Experience

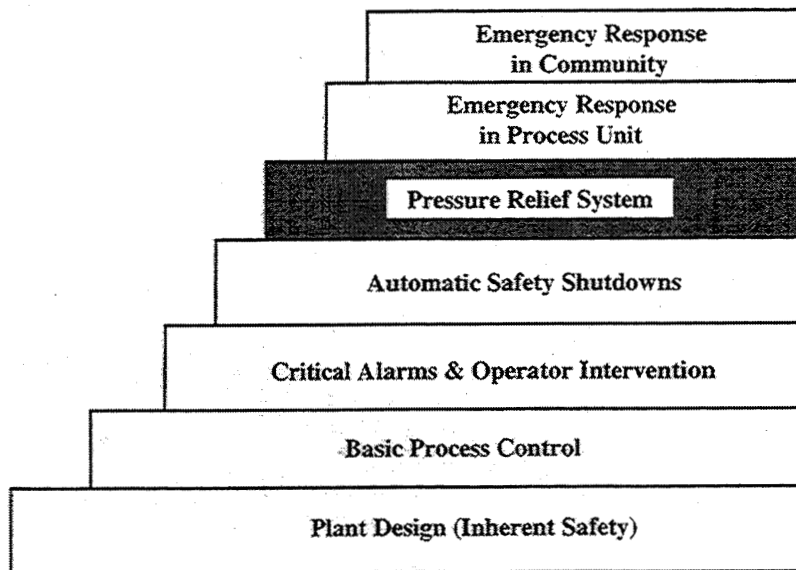


**100 Largest Losses
(1957-1987)**

**AVERAGE LOSS
\$40 MM (1987)**

IPD-2004/RS-1-4

Protection Layers



IPD-2004/RS-1-5

Pressure Relief Systems

- Essential part of process safety
- Governed by many codes & standards
- Simple operating concept
- Complex design process

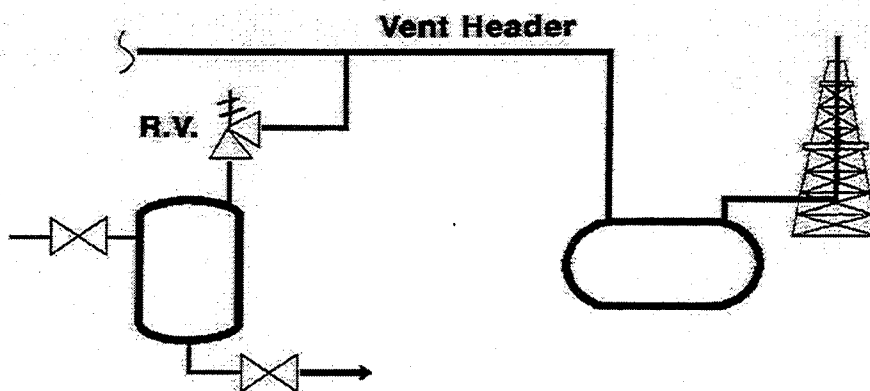
IPD-2004/RS-1-6

PRV's: What and Where

- Appearance
- Location

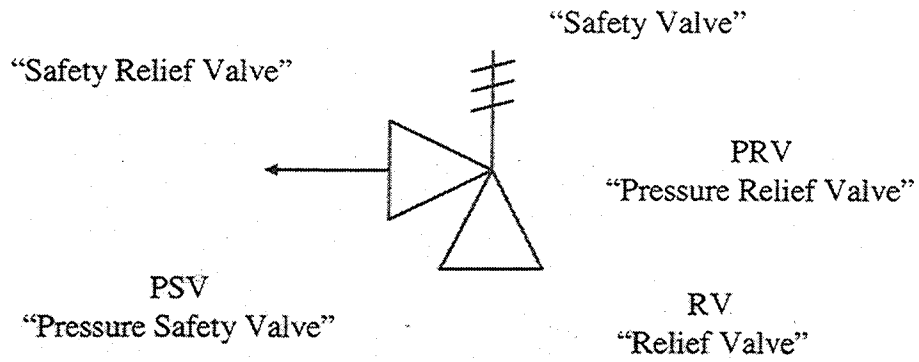
IPD-2004/RS-1-7

Typical Relief System



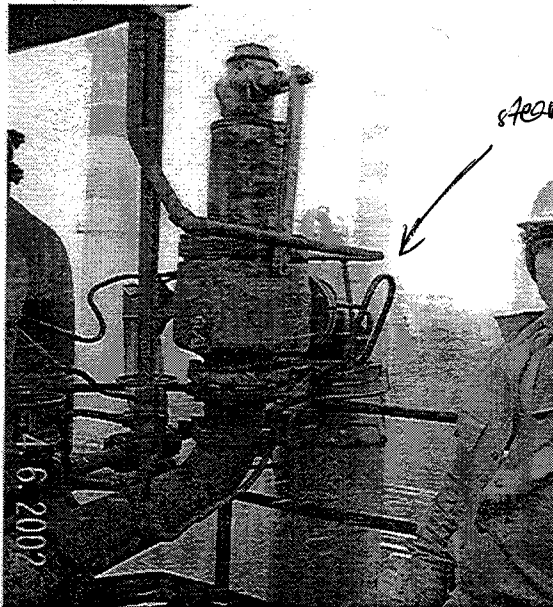
IPD-2004/RS-1-8

Common Names



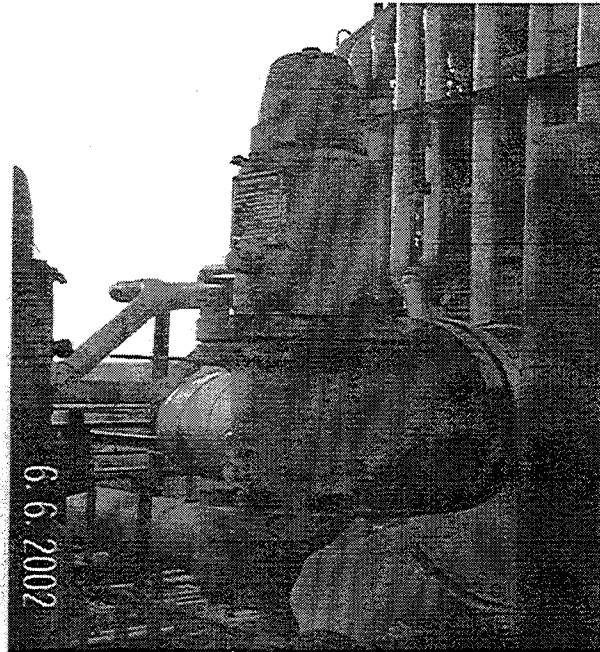
IPD-2004/RS-I-9

Spring-Loaded PRV



IPD-2004/RS-I-10

*Another
Spring-Loaded
PRV*



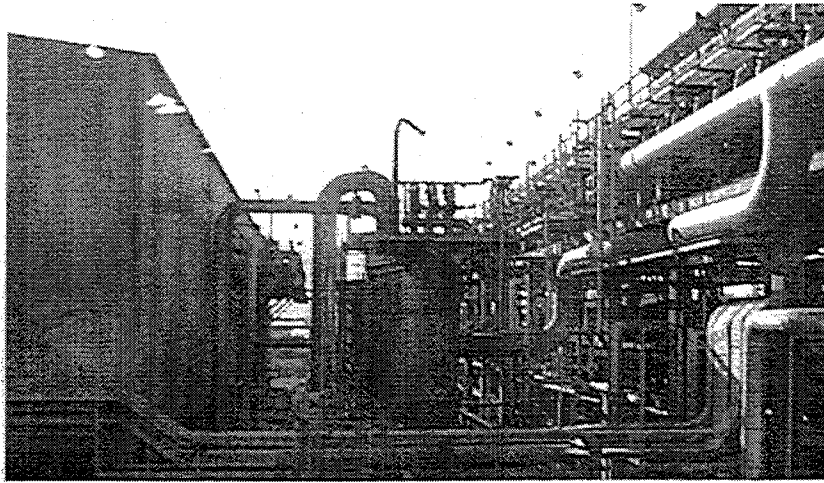
IPD-2004/RS-I-11

Location of Pressure Relieving Devices

- Location in Process : PRV Module
- Top
- Free-draining
- Close to vessel
- "Clean" location

IPD-2004/RS-I-12

Multiple Relief Valves
Platforming Unit Separator



IPD-2004/RS-I-13

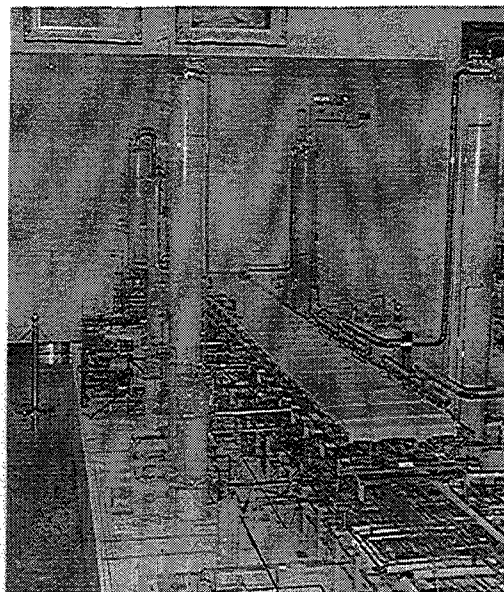
Model of the Same



IPD-2004/RS-I-14

UOP BTX Fract-Sulfolane Extraction

- Red lines are Overhead Vapor to air Cooler
- Yellow lines are relief to flare

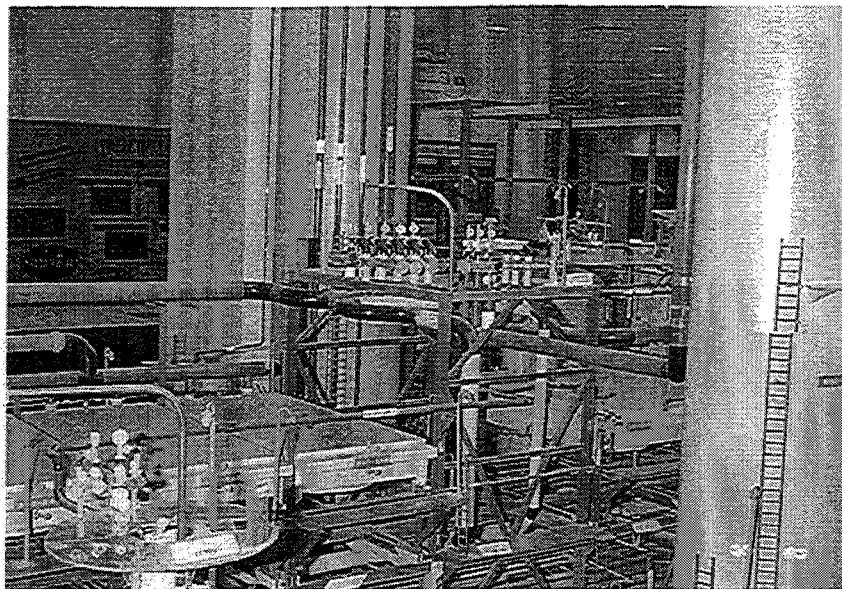


BTX Fractionation



WRS-1-16

Column Ovhd Piping (to Air Cooler) Multiple PRV's



IPD-2004/RS-17

Rules and Procedures

IPD-2004/RS-18

External Codes and Practices

- ASME Boiler & Pressure Vessel Code
- API Recommended Practices & Standards
- OSHA 1910.119 "Process Safety Management"
- State Regulations
- National Fire Protection Association
- AIChE Design Institute for Emergency Relief Systems

IPD-2004/RS-1-19

UOP Design Practice *From PRV Mapping*

- Establish Design Basis
- Review PFD Module Layer
- Generate Relief Loads
- Size and Select Relief Device
- Update P&ID's and Equipment Specs

IPD-2004/RS-1-20

Pressure Definitions

IPD-2004/RS-1-21

Operating Pressure Margin

- **Required for vessels**
 - Leakage and Reclosing
- **Caused by relief device**
- **Magnitude Depends on Type of Relief Device**
- **Refinery Standards**

IPD-2004/RS-1-22

ASME/API

<u>Vessel Requirements</u>		<u>Pressure</u>	<u>PRV</u>
Maximum Accumulated P	Fire	121	Max Relieving Pressure (for Sizing)
	2+ Valves Process	116	
	1 Valve Process	110	2nd + Valves Fire 2nd + Valves 1 Valve } Max Set Pressure
Design Pressure		105	
		100	
Max Operating Pressure		90	Reclose Pressure (typ.)

accumulated *% of design* *set point*

IPD-2004/RS-I-23

Other Terms

- Max Allowable Working Pressure - MAWP
 - Applies more for revamp work
 - Replace "design pressure" with MAWP
- Blowdown -- reclose
- Back Pressure

IPD-2004/RS-I-24

Identifying Causes of Overpressure

IPD-2004/RS-1-25

Pressure Increases When:

- Mass and/or energy accumulates in a system
- PRV's job is to relieve excess mass and energy to prevent accumulation.
- Construct H&WB at relief

IPD-2004/RS-1-26

Identify Causes of Overpressure

- Need to do thorough job
- Check Similar Designs
- Process Design Manual
- Keep THINKING!

IPD-2004/RS-1-27

Causes of Overpressure

- Look for:
 - Equipment failure or malfunction
 - Control failure or malfunction
 - Loss of utilities (electric, steam, air, water)
 - Human error
 - External fire
 - Uncontrolled reactions

partial failure / plant-wide failure

IPD-2004/RS-1-28

Process Deviations

- Construct an envelope around the section of the process
- Notice Changes in:
 - mass in & out
 - heat in & out
 - stream composition
 - utility availability

IPD-2004/RS-I-29

Relationship Between Events

- Unrelated or isolated
- Simultaneous or common
- Sequential

IPD-2004/RS-I-30

Unrelated/Isolated Events

- Philosophy: design for all isolated failures
- Block valve closed in error
- Equipment item fails
- Many will have the same end result

IPD-2004/RS-1-31

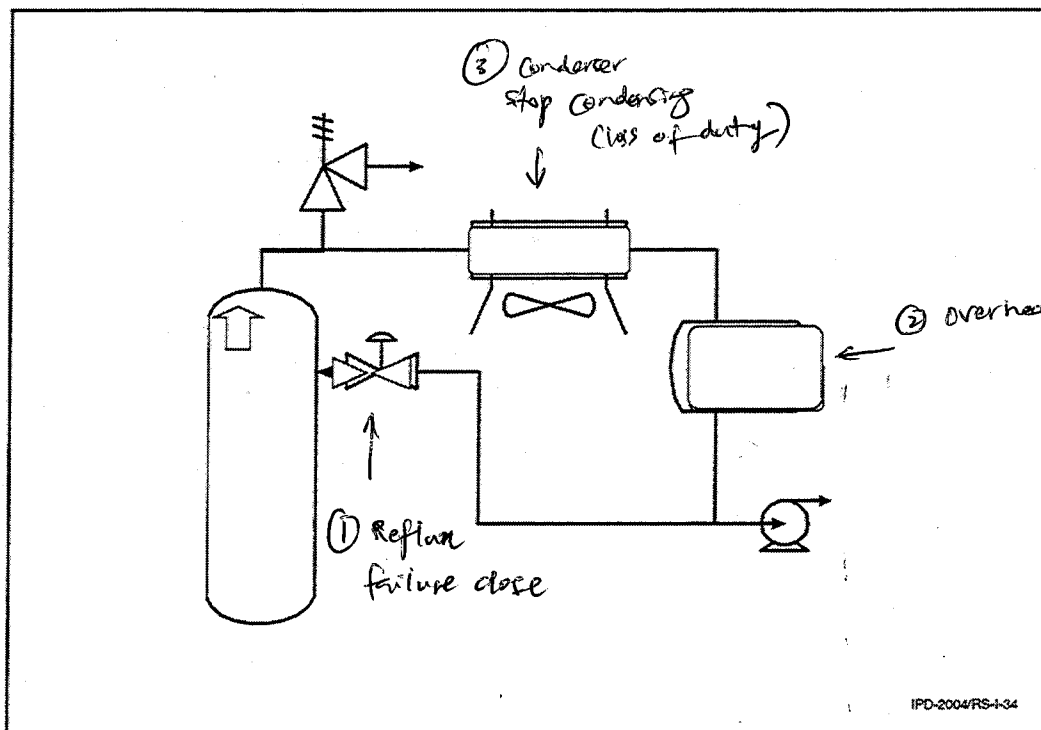
Simultaneous Events

- Utility failures:
 - Electric power
 - Cooling Water
 - Nitrogen
 - Instrument Air
- External fire
- Philosophy: design for the foreseeable effects of the common "root cause" event

IPD-2004/RS-1-32

Sequential Events

IPD-2004/RS-1-33



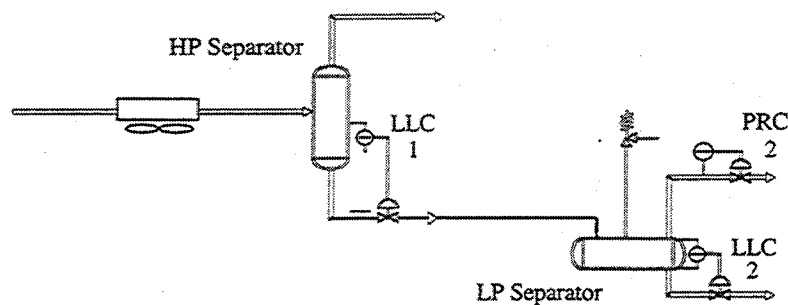
IPD-2004/RS-1-34

Examples of Sequential Events

- Reflux failure causes receiver to overfill with liquid, flood the condenser, and cause total loss of condenser duty
- PRC failure on net overhead vapor line causes blanketing of the condenser which results in total loss of condenser duty
- Loss of the liquid level in the products separator allows the reactor circuit to depressure to the stabilizer column, which results in blanketing of the overhead condenser

IPD-2004/RS-I-35

LP Flash Drum



IPD-2004/RS-I-36

Mass

Mass In & Out

LLC1 Fail	open X not overpressure close X
PRC2 Fail	open X close
LLC2 Fail	open X close
RV leaks	X
Condenser fail	X

IPD-2004/RS-137

Heat

Heat In & Out

LLC1 fail	open X close X
LLC2 fail	open X X close
PRC2 fail	open X close
Condenser fail	(no heat) ✓

Overall Fire

IPD-2004/RS-138

Composition Changes

LLC1 Fail → blowby
~~LLC2 Fail~~
PRC2 Fail
Condenser Fail
Reactor section upset

IPD-2004/RS-1-33

Reformate Splitter Problem

- What equipment is included in the scope of overpressure protection?
- Make a list of causes of overpressure for the splitter.
 - Refer to API RP 521, Section 3

IPD-2004/RS-1-40



Training Services

Pressure Relief Systems II

IPD-2004/RS-II-1

Overview

- Heat and Mass Balances at Relieving Conditions
- Sizing the Pressure Relief Device (PRV)
- PRV Inlet and Outlet Line Design

IPD-2004/RS-II-2



Training Services

Heat and Mass Balances at Relieving Conditions

IPD-2004/RS-II-3

Heat and Mass Balances

Steady State Balance:

$$\text{In} = \text{Out}$$

Unsteady State Balance:

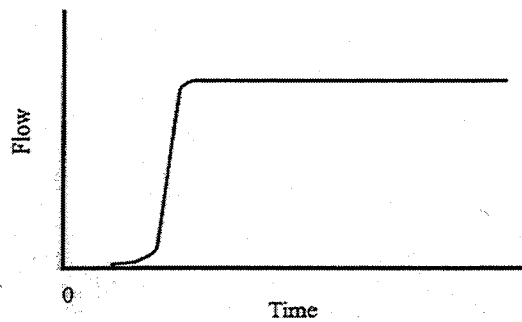
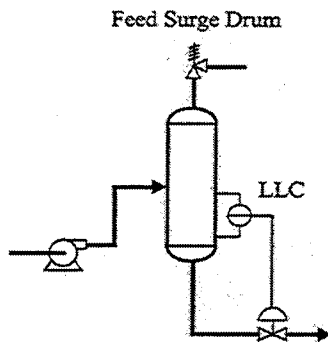
$$\text{In} + \text{Depletion} = \text{Out} + \text{Accumulation}$$

↑
inventory
evaporated

↑
inventory
condensed

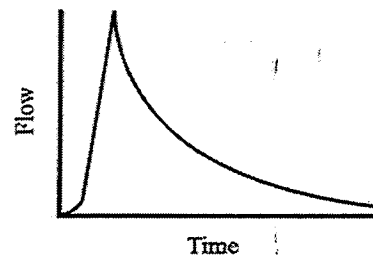
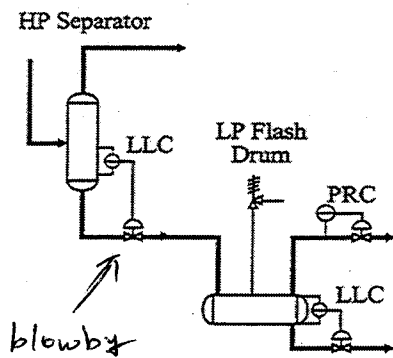
IPD-2004/RS-II-4

Blocked Outlet Transient



IPD-2004/RS-II-5
RS 31.4-R00-4

Control Valve Fails Open



IPD-2004/RS-II-6
RS 31.4-R00-5

Model Development

- Identify the credible causes for overpressure
- Determine the sequence of events
- *evaluate* → Rate equipment at relief conditions
- Consider instrumentation response
- Consider operator intervention

IPD-2004/RS-II-7

Simplifying Assumptions

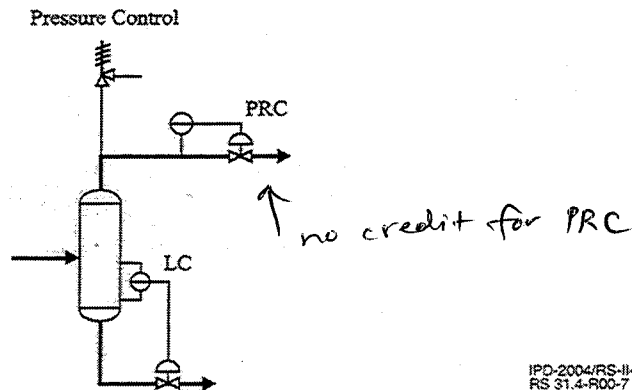
These simplifying assumptions should:

- Incrementally increase the relieving rate
- Have minimal economic impact
- Consider the risk associated with making a wrong assumption

IPD-2004/RS-II-8

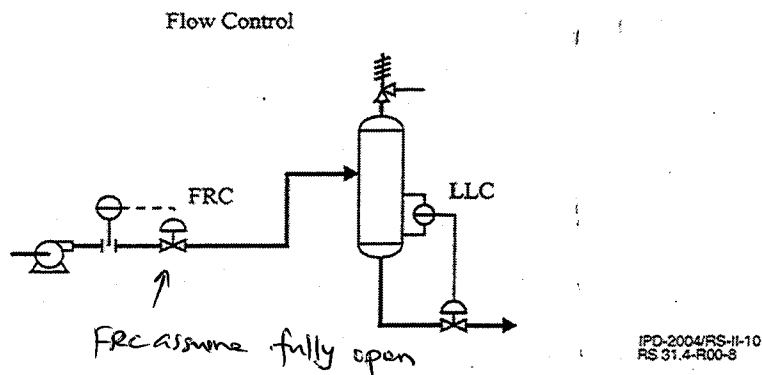
Instrumentation Response

API recommends that no credit for instrumentation response should be taken if that response reduces the relieving requirement.



Instrumentation Response

API recommends that instrumentation should be assumed to respond as designed if that response increases the relieving requirement.



Operator Response

UOP Approach:

- Almost never permitted
- Consider the time it takes to correct the problem
- Do not assume the operator will respond if the emergency affects many process variables simultaneously (multiple alarms)
- Assess operator effectiveness
- Consider the risk involved if the operator does not respond as assumed

IPD-2004/RS-II-11



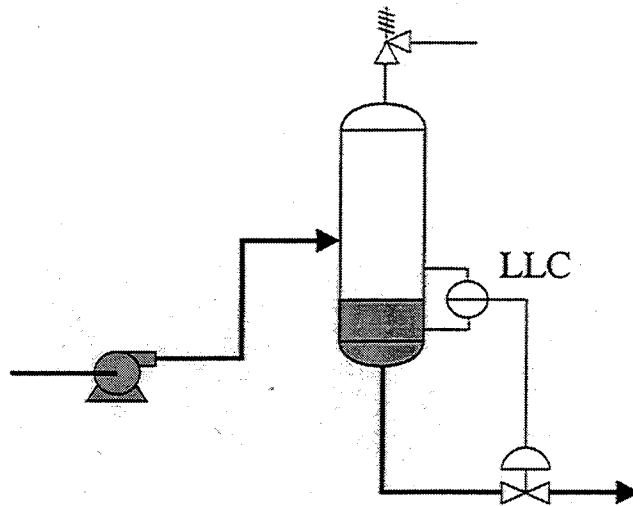
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External Fire

PRV Calculations

IPD-2004/RS-II-12

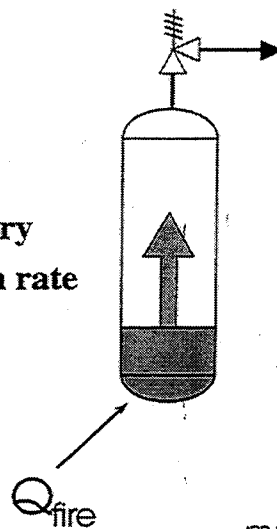
Feed Surge Drum



IPD-2004/RS-II-13
RS 31.5-R00-8a

Fire Case Model

- Feed stops
- Heat input vaporizes liquid inventory
- Relieving rate equal to vaporization rate



IPD-2004/RS-II-14
RS 31.5-R00-8c

Heat Input Due to External Fire

$$Q = 37140 FA^{0.82}$$

where:

Q = heat input due to fire, Kcal/hr

F = environmental factor $F = 1$ for bare vessel

A = wetted surface area, m²

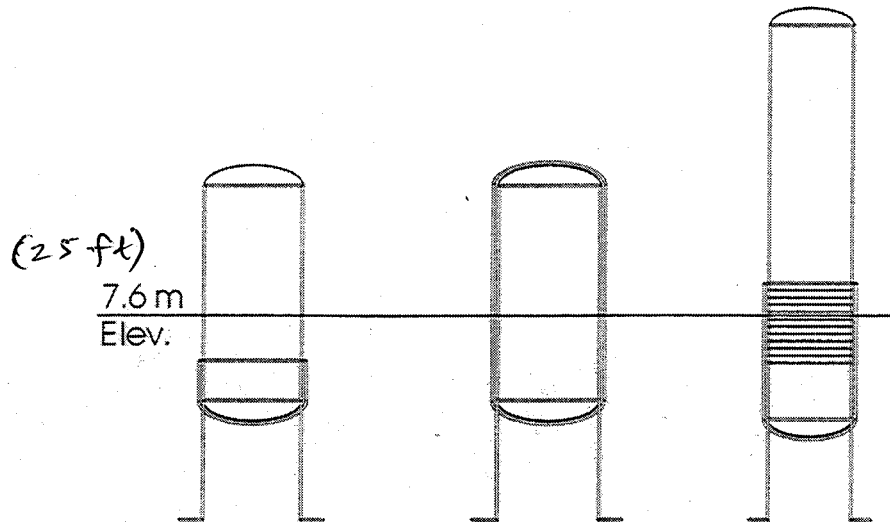
IPD-2004/RS-II-15

Environmental Factor

- Permits credit for insulation
- If insulation can be dislodged by impact of water spray, $F = 1.0$
- Assumes General Refining Conditions:
 - Ground sloped; no standing liquid pools
 - Adequate fire fighting equipment available
 - Trained personnel

IPD-2004/RS-II-16

Wetted Surface Area



IPD-2004/RS-II-17
RS 31.5-R00-10

Vaporization Rate

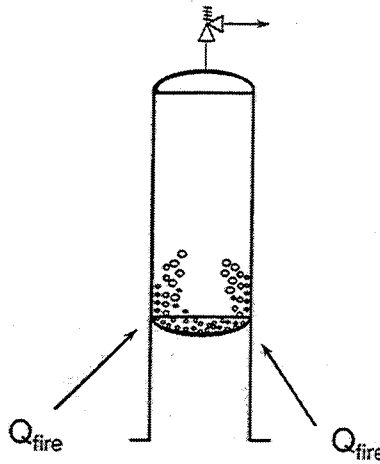
$$W_{fire} = \frac{Q_{fire}}{\Delta H_v}$$

- Minimum latent heat of vaporization is 28 Kcal/kg.
- Above the critical temperature, the expansion rate of the dense fluid is estimated using the vaporization approach.

IPD-2004/RS-II-18

Liquid Filled Vessel

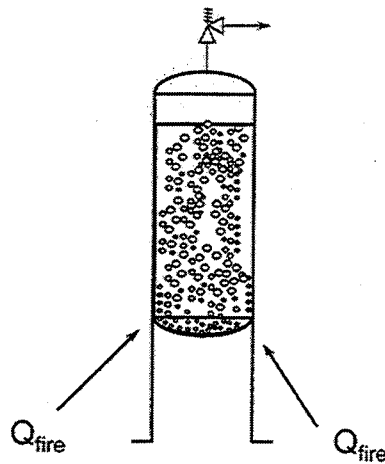
Initially, all liquid is displaced out the PRV



IPD-2004/RS-II-19
RS 31.5-R00-13a

Liquid Filled Vessel

Later on, sufficient space is available for vapor-liquid disengagement, and all vapor flows out the PRV



IPD-2004/RS-II-20
RS 31.5-R00-13b

Design Considerations *(for liquid full vessels)*

- Size PRV for all vapor flow (*total disengagement*)
- Rate PRV selected for all liquid flow and report liquid flow rate
- Determine the total volume of liquid discharged to relief header and report volume for sizing KO drum
 - Volume in vessel head plus 0.61 m of tangent length
- If auto-chilling occurs upon flashing, report minimum temperature

IPD-2004/RS-II-21



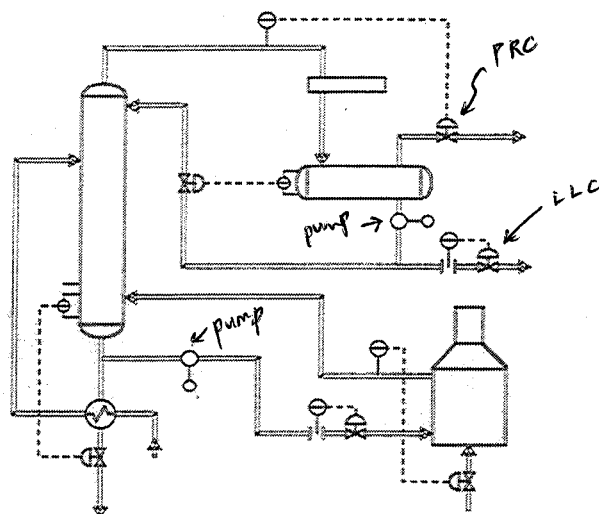
Training Services

Debutanizer

PRV Calculations

IPD-2004/RS-II-22

Debutanizer



IPD-2004/RS-II-23
RS 31.6-R00-15

Summary of Relief Scenarios for the Debutanizer

- Electrical power failure (refinery-wide & partial) (lose condenser → lose reflux)
lose pumps
- Loss of air-cooled condenser duty (ditto)
- Reflux failure
- PRC fails closed
- Product separator LLC fails open
- External fire

IPD-2004/RS-II-24

Total Power Failure Model

Basic Assumptions:

- Feed stops or eventually stops
- Reflux stops
- No fractionation occurs
- Heat balance constructed at end of relieving case

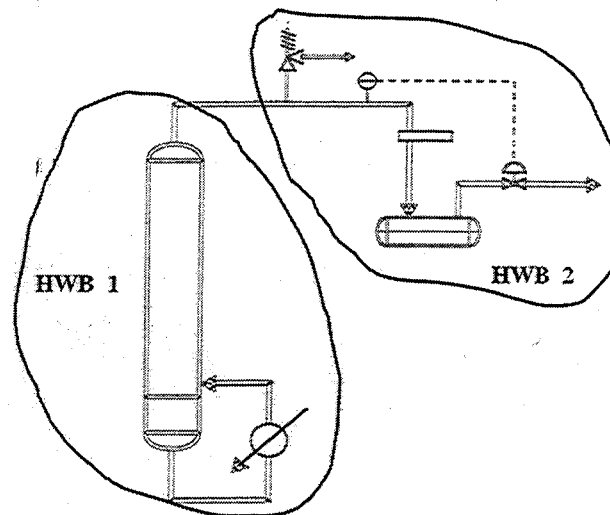
IPD-2004/RS-II-25

Partial Power Failure

- Reboiler circulating pumps on different electrical circuit or on different voltage system
- Various pumps may be located in another process unit
- Pumps driven by a source other than electrical power (Steam Turbine, Gas Turbine)

IPD-2004/RS-II-26

Debutanizer (Power Failure Model)



IPD-2004/RS-II-27
RS 31.6-R00-15

Power Failure HWB 1

	<u>Mass</u>	<u>Heat</u>
Reboiler Duty		$+ Q_{REB}$
Vaporized Inventory	$- W_I$	$- W_I h_v$
Depleted Liquid	$+ W_I$	$+ W_I h_l$
Total	<u>0</u>	<u>0</u>

IPD-2004/RS-II-28

Power Failure (working equations)

$$0 = Q_{REB} - W_I h_v + W_I h_l$$

$$W_I = \frac{Q_{REB}}{\Delta H_v}$$

**Vaporization rate based on simultaneous solution
of heat and mass balance**

IPD-2004/RS-II-29

Power Failure Relieving Rate

Residual Condenser Credit (natural convection):

$$W_{cond} = \frac{Q_{cond}}{\Delta H_v}$$

**Relieving Rate (difference between vaporization
rate and residual condensation rate):**

$$W_{rel} = W_I - W_{cond}$$

IPD-2004/RS-II-30



Training Services

Loss of Air-Cooled Condenser Duty

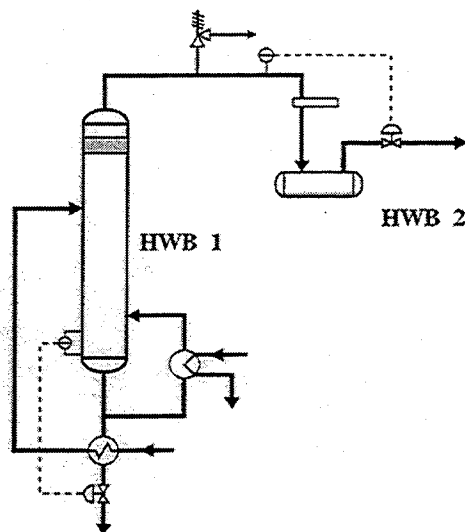
IPD-2004/RS-II-31

Loss of Condenser Duty

- PRC opens wide to attempt pressure control
- Debutanizer pressure increases in spite of controller action
- Liquid inventory in overhead receiver decreases
- LLC on overhead receiver stops reflux by resetting FRC on reflux
- Heavy material migrates to top section as liquid contacting in top section of debutanizer stops

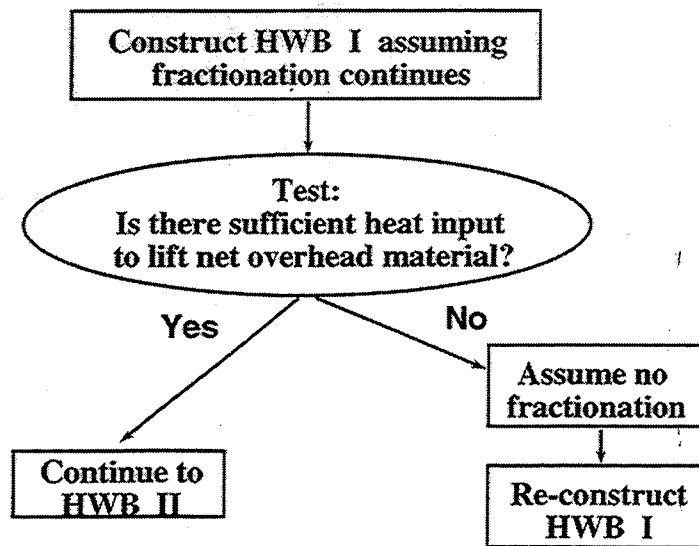
IPD-2004/RS-II-32

Debutanizer (loss of condenser duty)



IPD-2004/RS-II-33
RS 31.6-R00-17

Decision Tree



IPD-2004/RS-II-34

Condenser Failure HWB I

	<u>Mass</u>	<u>Heat</u>
Feed	$+W_f$	$+W_f h_f$
Net bottoms	$-W_{nb}$	$-W_{nb} h_{nb}$
Reboiler Duty		$+Q_{reb}$
Net Overhead Vapor	$-W_{no}$	$-W_{no} h_{no}$
Vaporized Inventory	$-W_I$	$-W_I h_v$
Vaporized Inventory	$+W_I$	$+W_I h_l$
Depleted Liquid Inventory		
Total	<u>0</u>	<u>0</u>

IPD-2004/RS-II-35

Vaporized Liquid Inventory (working equation)

$$0 = W_F h_F - W_{NB} h_{NB} + Q_{REB} - W_{NO} h_{NO} - W_I h_v + W_I h_l$$

$$W_I = \frac{(W_F h_F - W_{NB} h_{NB} + Q_{REB} - W_{NO} h_{NO})}{\Delta H_v}$$

Inventory rate must have a positive sign for model assumption that fractionation continues to be valid

IPD-2004/RS-II-36

Required Calculations

Determine:

- Reboiler duty at relief
- Heat content of net streams
- Vaporization rate of liquid inventory
- Overhead condenser credit

IPD-2004/RS-II-37

Condenser Failure HWB II

	<u>Mass</u>	<u>Heat</u>
Overhead Vapor	$+W_o$	$+W_o h_v$
Relieving Rate	$-(W_o - W_{cond})$	$-(W_o - W_{cond})h_v$
Condenser Duty		$-Q_{cond}$
Accum. Liq. Inventory	$-W_{cond}$	$-W_{cond} h_l$
Total	0	0

IPD-2004/RS-II-38

Loss of Condenser Duty Relieving Rate

Residual condenser duty (natural convection):

$$W_{cond} = \frac{Q_{cond}}{\Delta H_v}$$

Relieving Rate (difference between total vapor rate leaving debutanizer and residual condensation):

$$W_{rel} = W_{total} - W_{cond}$$

IPD-2004/RS-II-39

Reality Check

Always check the reasonableness of the calculated relieving rates for each case.

- During a reflux failure, loss of condenser duty, or PRC failing closed, the relieving rate should be similar to the normal gross overhead rate.
- During a electrical power failure, the relieving rate should be similar to the normal reboiler vapor rate.

If the rates differ significantly from that given above, rationalize the difference, i.e. dramatic loss of reboiler duty due to low LMTD.

IPD-2004/RS-II-40

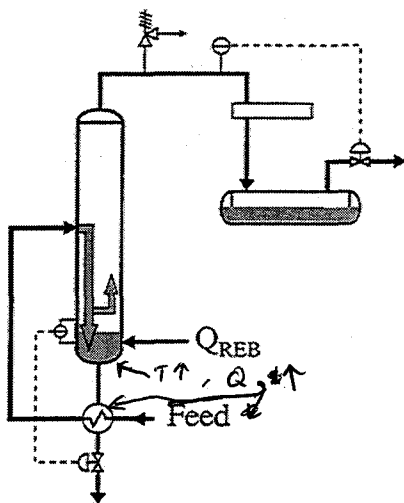
↑
process outlet temp increases,
LMTD decreases -
 $Q_{reb} = UA (LMTD)$

Alternate Calculation (when model assumption is not valid)

- If the "excess" heat is negative, the assumption that fractionation continues as normal is not valid.
- A negative value means that there is not sufficient heat to vaporize the net overhead material. Thus, the bubble and dew points used to determine the heat balance are not valid.
- Although some fractionation will occur, the problem is bracketed by assuming no fractionation occurs, i.e. only flashing occurs in the column.

IPD-2004/RS-II-41

No Fractionation Occurs



1. Assume a bottoms temperature.
2. Calculate reboiler duty using assumed temperature.
3. Calculate feed-bottoms exchanger duty using assumed temperature.
4. Flash feed at normal feed enthalpy plus reboiler and feed-bottoms exchanger duty.
5. Compare calc'd temperature to assumed temperature.
6. Vapor rate leaving column is vapor rate from the flash.

IPD-2004/RS-II-42
RS 31.6-R00-21



Training Services

Sizing the Pressure Relief Device

IPD-2004/RS-II-43

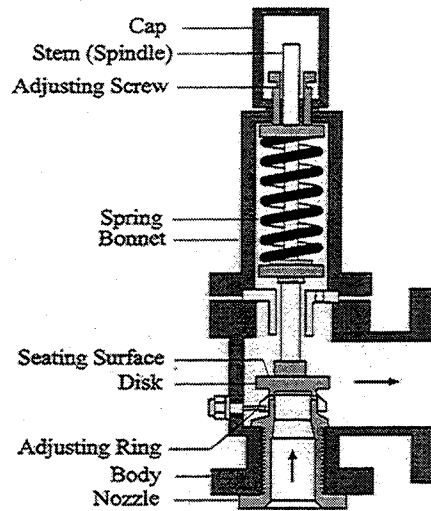
Types of Relief Devices

- Spring loaded pressure relief valve
- Pilot operated pressure relief valve
- Rupture disc
- Weighted pallet
- Vent line

conventional
balance bellows

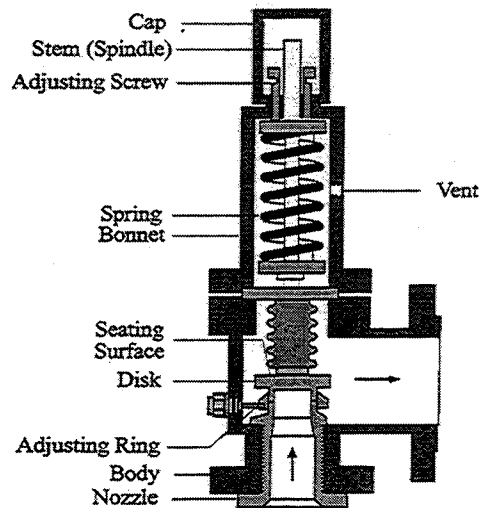
IPD-2004/RS-II-44

Conventional Pressure Relief Valve



IPD-2004/RS-II-45
RS-31.7-R00-22

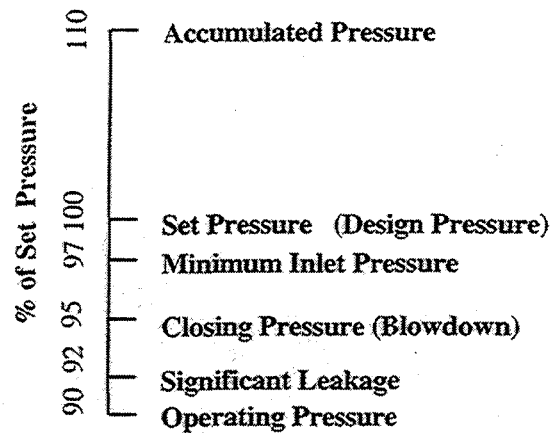
Balanced-Bellows Pressure Relief Valve



IPD-2004/RS-II-46
RS-31.7-R00-23

Summary

Spring-Loaded Pressure Relief Valves



IPD-2004/RS-II-47

Sizing for Vapor

Spring Loaded Relief Valve

$$RA = \frac{1.02 W}{CKP_{acc} K_b} \sqrt{\frac{TZ}{MW}}$$

RA = Required area, cm²

W = Flow rate, kg/hr

C = Coefficient based on C_p/C_v ratio

K = Coefficient of discharge (K = 0.975 for typical manufacturer)

K_b = Back pressure correction factor

P_{acc} = Accumulated pressure, kg/cm²(a)

T = Temperature °K

Z = Compressibility factor

MW = Molecular weight, kg/kg mol

If heat capacity ratio is unknown, assume CK = 387

IPD-2004/RS-II-48

Sizing for Vapor Spring-Loaded Relief Valve

$$RA = \frac{W}{CKP_{acc}K_b} \sqrt{\frac{TZ}{MW}}$$

If heat capacity ratio is
unknown, assume CK = 306

RA = Required area, in²

W = Flow rate, lb / hr

C = Coefficient based on C_p / C_v ratio

K = Coefficient of discharge (K = 0.97 for JOS / JBS series)

K_b = Back pressure correction factor

P_{acc} = Accumulated pressure, psia

T = Temperature, °R

Z = Compressibility factor

MW = Molecular weight, lb / lb mol

Sizing for Nonflashing Liquid Pressure Relief Valve with Liquid Trim

$$RA = \frac{Q}{5K} \sqrt{\frac{SpGr}{P_{acc} - P_b}}$$

RA = Required area, cm²

Q = Flow rate, m³/h

K = Coefficient of discharge

SpGr = Specific gravity of liquid

P_{acc} = Accumulated pressure, kg/cm²g

P_b = Total back pressure, kg/cm²g

Sizing for Liquid Pressure Relief Valve with Liquid Trim

$$RA = \frac{Q}{38K} \sqrt{\frac{SpGr}{P_{acc} - P_b}}$$

RA = Required area, in²

Q = Flow rate, gpm

K = Coefficient of discharge

$SpGr$ = Specific gravity of liquid

P_{acc} = Accumulated pressure, psig

P_b = Total back pressure, psig

Sizing for Two-Phase Flow UOP Practice

- **Select pressure for flash calculation**

Subcooled liquid - 90% of saturated pressure at inlet

Two phase @ inlet - 90% of inlet pressure

- **Flash relief stream at nozzle outlet pressure to get density ($\Delta S=0$)**

- **Calculate ω parameter**

$$\omega = 9 \left(\frac{P_{SAT}}{P_{FLASH}} - 1 \right) \quad \text{Subcooled}$$

$$= 9 \left(\frac{P_{INLET}}{P_{FLASH}} - 1 \right) \quad \text{Two-phase}$$

Sizing for Two-Phase Flow

Two-phase Inlet only!

- Calculate choke pressure ratio

$$\eta_c = \frac{P_c}{P_o} = 0.55 + 0.217 \ln(\omega) - 0.046(\ln(\omega))^2 + 0.004(\ln(\omega))^3$$

- Calculate required orifice area

$$RA = \frac{W_{rel} \sqrt{\omega}}{K_{d2} K_w \eta_c \sqrt{P_o \rho_o}} \quad (\text{consistent units required})$$

IPD-2004/RS-II-53



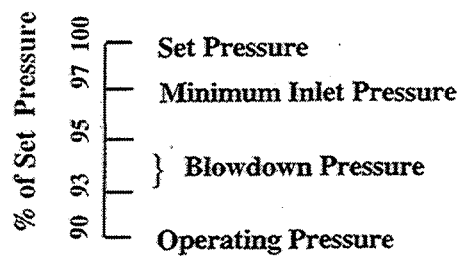
Training Services

PRV Inlet and Outlet Line Design

IPD-2004/RS-II-54

Inlet Pressure Loss

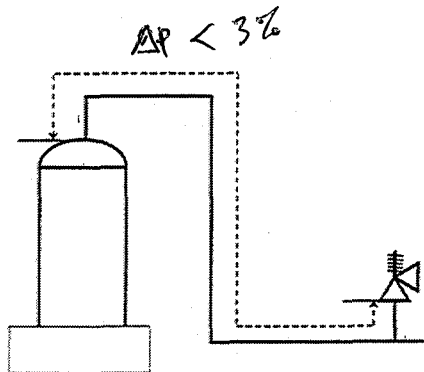
- Valve will “chatter” if the inlet pressure loss exceeds the blowdown pressure of valve
- Blowdown pressure varies between 5-7% of PRV set pressure (depending on valve model)
- Inlet pressure loss is limited to 3% of the PRV set pressure



IPD-2004/RS-II-55

Inlet Line Sizing

Pressure Drop Calculated Using Rated Capacity of Valve



Equivalent length includes:

- Sudden contraction
- Elbows
- Tee
- Reducer
- Length of piping

IPD-2004/RS-II-56
RS 31.7.R00-40

Design Options for Meeting 3% Inlet ΔP

- Do not ignore---this is a severe problem
- Locate PRV nearer to vessel (shorter inlet line)
- Increase diameter of inlet piping
- Select a pilot operated relief valve with remote pressure source

IPD-2004/RS-II-57

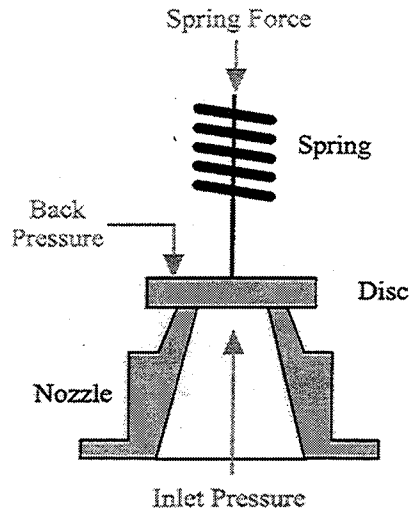


Training Services

PRV Outlet Line Design

IPD-2004/RS-II-58

Force Balance Conventional Pressure Relief Valve

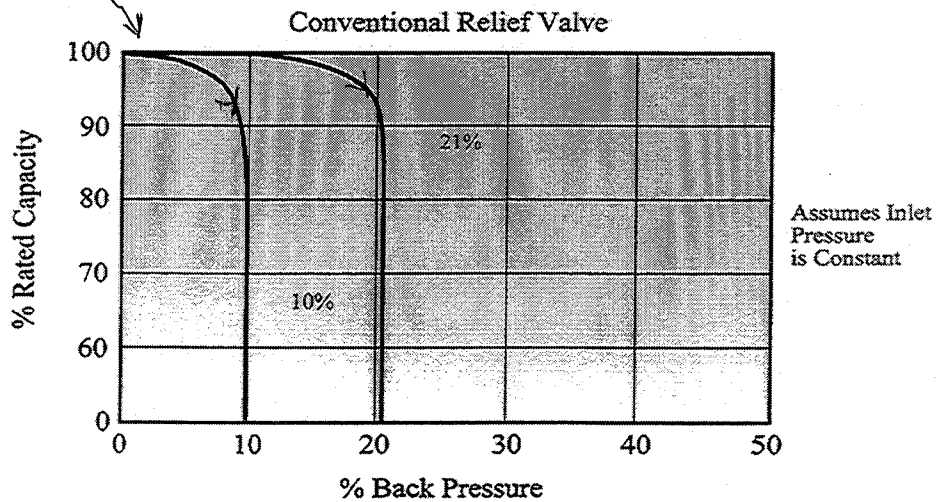


*Force balance:
Inlet pressure = Spring force
+ back pressure*

IPD-2004/RS-II-59
RS 31.7.R00-25

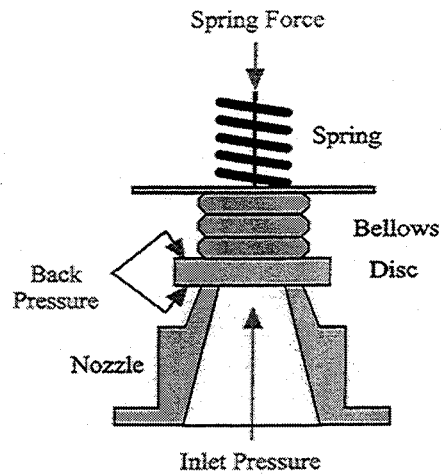
Effect of Back Pressure

*Rated capacity
decrease as
back pressure increase*



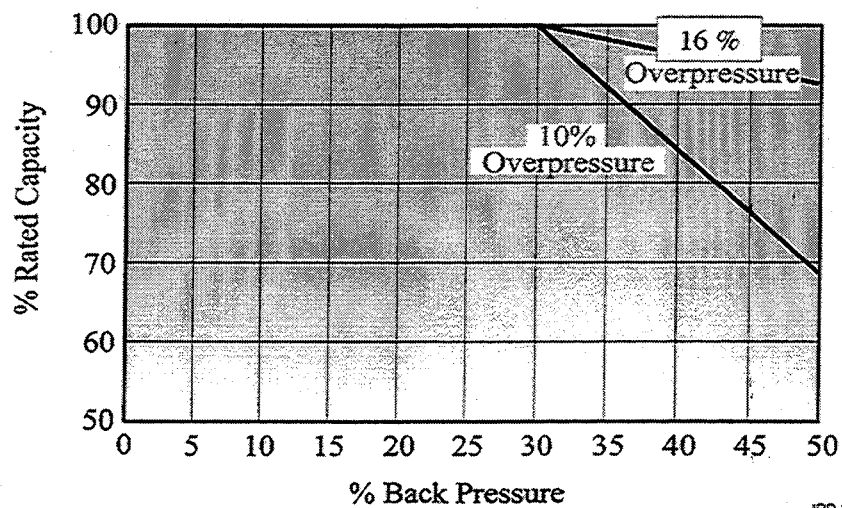
IPD-2004/RS-II-60
RS 31.7.R00-29

Force Balance *Balanced Pressure Relief Valve*



IPD-2004/RS-II-61
RS 31.7-R00-30

Effect of Back Pressure *Balanced Relief Valve*



IPD-2004/RS-II-62
RS 31.7-R00-31

Design Options for High Back Pressure

- Increase diameter of discharge piping
- Select a balanced bellows or pilot operated relief valve which can withstand a higher back pressure
- Offset the back pressure by lowering the spring setting of a non-balanced pressure relief valve (constant back pressure only)
- Lower set pressure of non-balanced pressure relief valve below vessel design pressure to offset the higher back pressure

IPD-2004/RS-II-63

Outlet Line Design

- Cannot reduce diameter below PRV outlet flange size
- PRV located at high point in closed piping system; free-draining to header
- Choked flow at expansions must be considered in ΔP calculations
- Thrust force used to design supports

IPD-2004/RS-II-64

Detailed Outlet Design

- Relieving to a process vessel
- Pump discharge back to inlet
- Acid Relief Neutralization
- Revamp

IPD-2004/RS-II-65

Pressure Relief Valves

Questions?

IPD-2004/RS-II-66



Training Services

Pressure Relief Systems III

IPD-2004/RS-III-1

Session III Outline

- Relief Device Characteristics
- Material Selection
- Discharge Location
- Other Design Issues
- Resources

IPD-2004/RS-III-2

Pressure Relief Device Characteristics

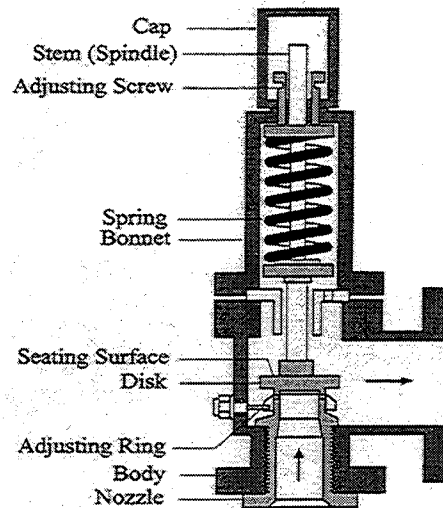
IPD-2004/RS-III-3

Types of Relief Devices

- Spring loaded pressure relief valve
- Pilot operated pressure relief valve
- Rupture disc
- Weighted pallet
- Vent line

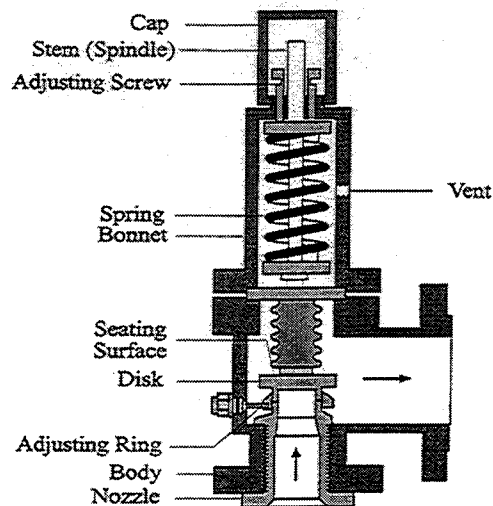
IPD-2004/RS-III-4

Conventional Pressure Relief Valve



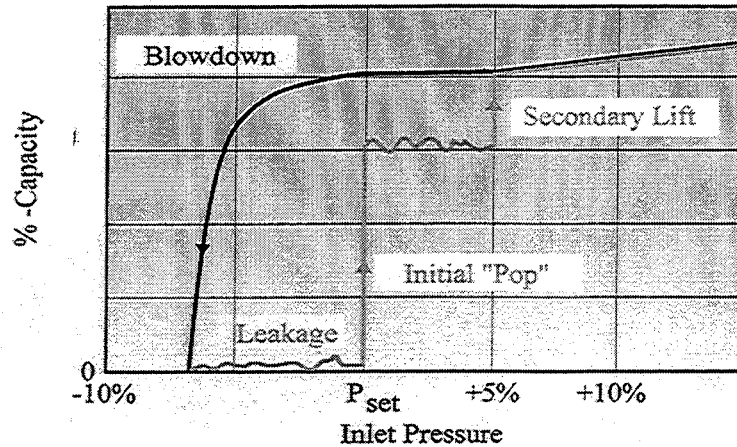
IPD-2004/RS-III-5
RS 31.7-R00-22

Balanced-Bellows Pressure Relief Valve



IPD-2004/RS-III-6
RS-31.7-R00-23

Opening & Closing Spring-Loaded PRV



IPD-2004/RS-III-7
RS31.7.R00-24

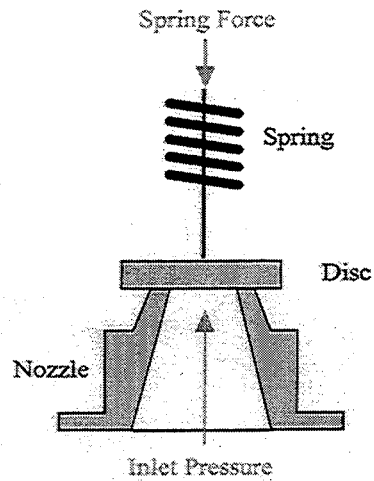
Impact on Design

- Spread between operating and set pressure (preventing leakage through valve)
- Inlet pressure loss (avoiding valve chattering)
- Capacity affected by back pressure (outlet line size, spring setting, valve type)
- Rapid opening (free draining inlet and outlet piping)

IPD-2004/RS-III-8

Force Balance

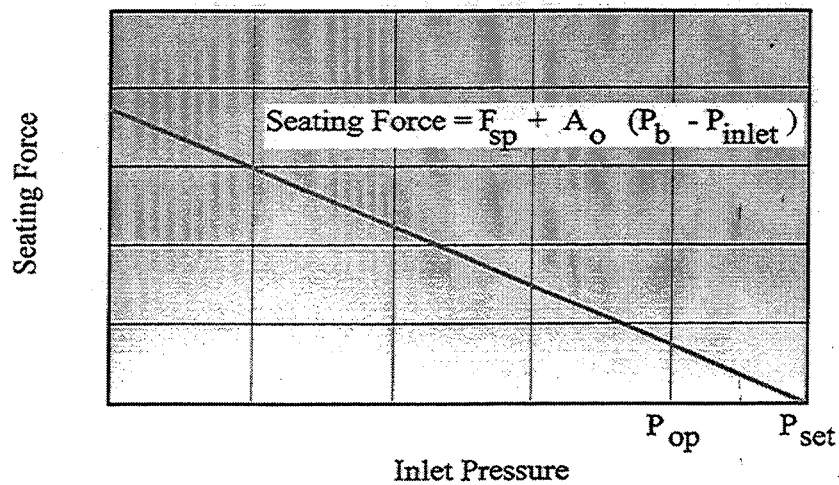
Conventional Pressure Relief Valve



IPD-2004/RS-III-9
RS 31.7.R00-27

Seating Force

Spring-Loaded Pressure Relief Valve

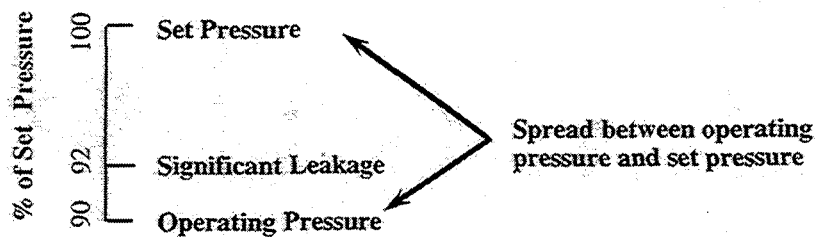


IPD-2004/RS-III-10
RS 31.7.R00-26

Valve Leakage

start to leak
↓

- Valve will “simmer” at 92% of set pressure
- Set pressure of valve is chosen relative to operating pressure and selected to minimize leakage, e.g. 10% or 1.75 kg/cm² whichever is greater



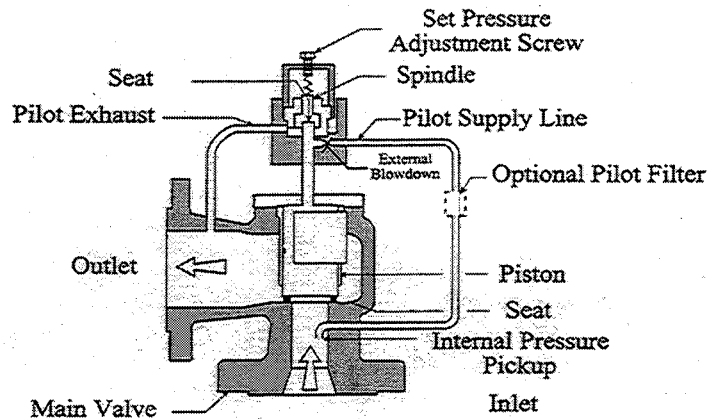
IPD-2004/RS-III-11

Design Options for Leakage

- Provide greater pressure differential between operating and set pressure
- Lap valve seat surfaces to improve tightness
- Install elastomeric seats
- Select a pilot operated relief valve
- Install spare relief valves to permit maintenance while unit is operating

IPD-2004/RS-III-12

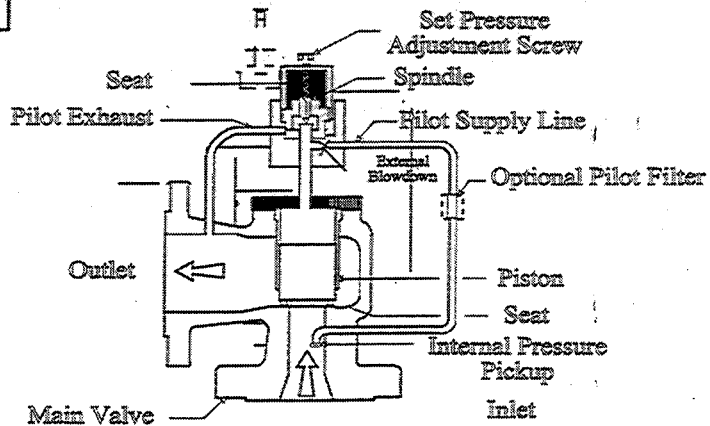
Pop-Action Pilot-Operated Valve (Flowing Type)



IPD-2004/RS-III-13
RS 31.7-.R00 -34

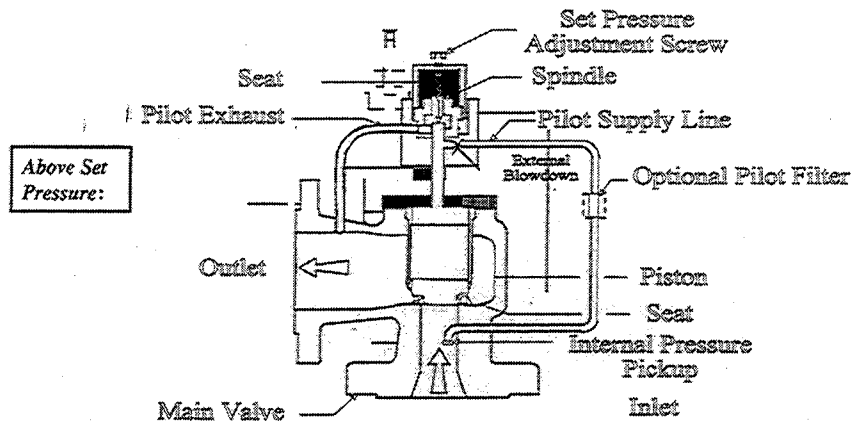
Pop-Action Pilot-Operated Valve (Flowing Type)

**Below Set
Pressure:**



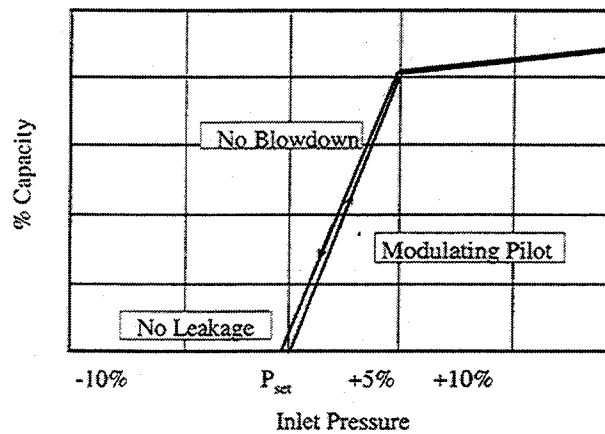
IPD-2004/RS-III-14
RS 31.7-.R00 -34

Pop-Action Pilot-Operated Valve (Flowing Type)



IPD-2004/RS-III-15
RS 31.7-R00-34

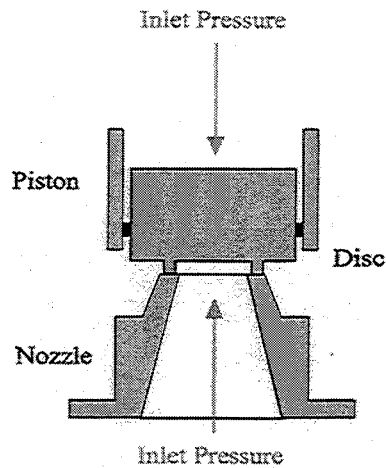
Flow Characteristic Open and Closing



IPD-2004/RS-III-16
RS 31.7-R00-35

Force Balance

Pilot Operated Pressure Relief Valve



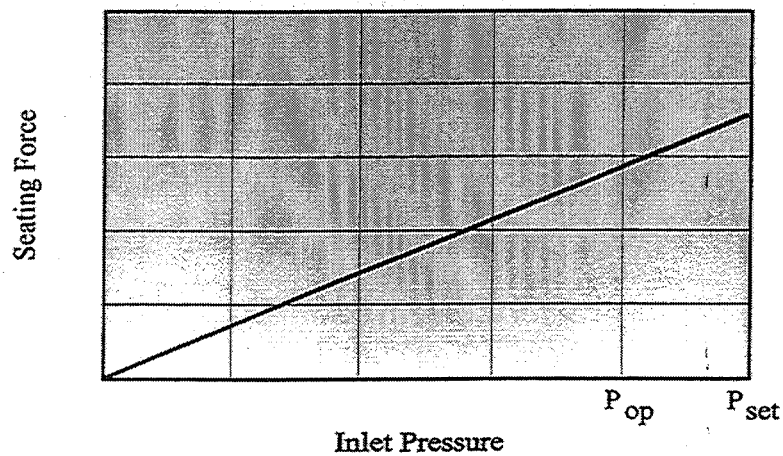
Exposed area on top of piston
is greater than the exposed area
at bottom of piston.

$$\text{Seating Force} = P_{\text{inlet}} (A_{\text{top}} - A_{\text{bott}})$$

IPD-2004/RS-III-17
RS 31.7-R00-36

Seating Force

Pilot Operated Pressure Relief Valve



IPD-2004/RS-III-18
RS 31.7-R00-37

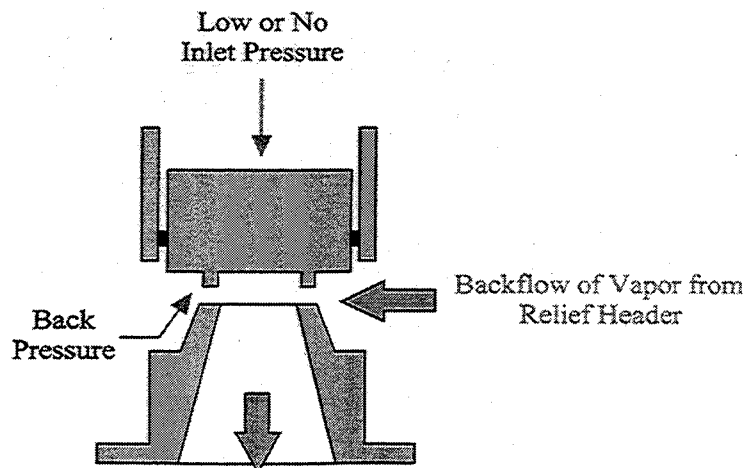
Unique Considerations

Pilot Operated Relief Valves

- Backflow preventer
- Remote sense line
- Material compatibility (elastomeric seats and seals, lower metallurgy availability)
- More restrictive temperature limits

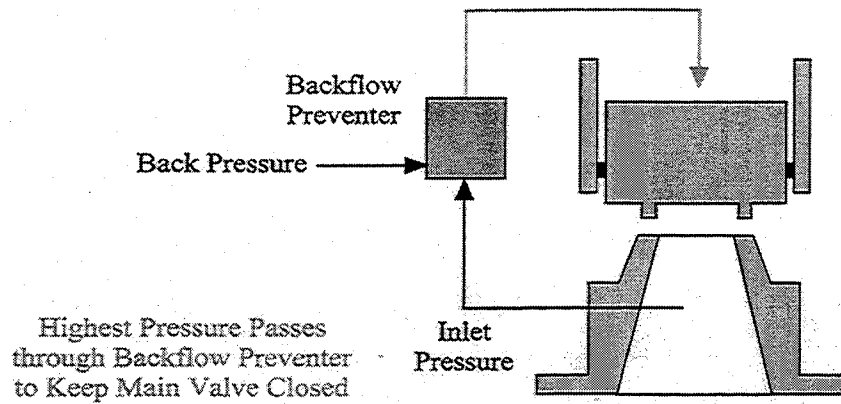
IPD-2004/RS-III-19

Backflow Through a PORV



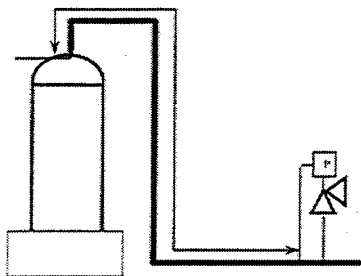
IPD-2004/RS-III-20
RS 31.7-R00-36

Backflow Preventer



IPD-2004/RS-III-21
RS 31.7-R00-39

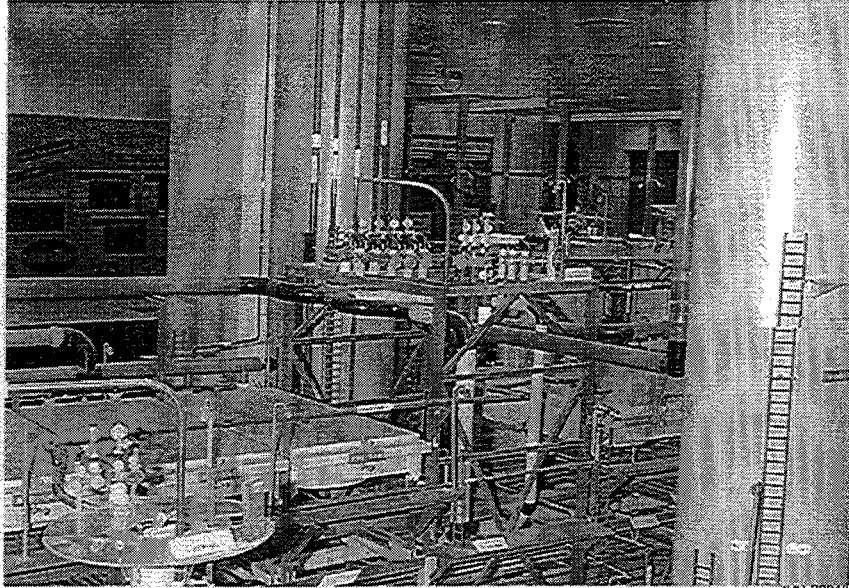
Remote Sensing *Applying the 3% Pressure Loss Constraint*



- 3% rule applies to pressure drop from vessel to sense point
- Sense point should be on relatively low velocity line (Bernoulli effect)
- Capacity of PORV may have to be adjusted for high inlet pressure losses ($\Delta P > 5\% P_{set}$)

IPD-2004/RS-III-22

Column Ovhd Piping (to Air Cooler) Multiple PRV's



IPD-2004/RS-III-23

PORV Advantages

- Seat tightness
- Remote pressure sensing
- Less sensitive to back pressure
- Field testing of pilot set point
- Remote depressuring
- Ability to modulate flow

IPD-2004/RS-III-24

PORV Disadvantages

- Difficult to specify
- Temperature constraint
- Elastomeric compatibility
- Limited metallurgy selection
- Unfamiliar technology

IPD-2004/RS-III-25

PORV Applications

- High pressure (above 70.0 kg/cm²)
- Low pressure (below 1.4 kg/cm²)
- Revamps (operating pressure close to set pressure)
- Remote sense required for 3% rule
- Minimize valve leakage
- Large liquid relief

IPD-2004/RS-III-26

PORV Sizing for Vapor

$$RA = \frac{1.34 W}{CKP_{acc}} \sqrt{\frac{TZ}{MW}}$$

RA = Required area, cm^2

W = Flow rate, kg/hr

C = Coefficient based on C_p / C_v ratio

K = Coefficient of discharge ($K = 0.86$ for 400 series PORV)

P_{acc} = Accumulated pressure, $\text{kg/cm}^2\text{a}$

T = Temperature, $^{\circ}\text{K}$

Z = Compressibility factor

MW = Molecular weight, kg/kg mol

IPD-2004/RS-III-27

PORV Sizing for Liquid

$$RA = \frac{Q}{5K} \sqrt{\frac{SpGr}{P_{acc} - P_b}}$$

RA = Required area, cm^2

Q = Flow rate, m^3/hr

K = Coefficient of discharge ($K = 0.67$ for 400 series PORV)

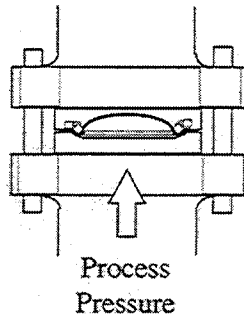
P_{acc} = Accumulated pressure, $\text{kg/cm}^2\text{g}$

P_b = Total back pressure, $\text{kg/cm}^2\text{g}$

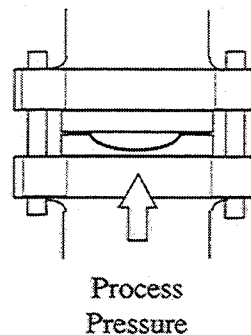
IPD-2004/RS-III-28

Rupture Discs

Tension Loaded



Reverse Buckling



Safety Head

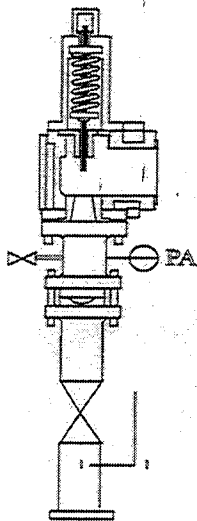
IPD-2004/RS-III-29
RS 31.7-R00-41

Rupture Disc Applications

- Protecting temporary piping used for inert gas regeneration operation
- Installed under a PRV
 - to prevent leakage
 - to protect valve from corrosive environment
- Fast response time required, e.g. Tube rupture in a high pressure exchanger which is liquid filled on LP side
- Runaway reaction (highly exothermic) where pressure is intentionally reduced below normal operating pressure to regain control

IPD-2004/RS-III-30

In Combination with PRV

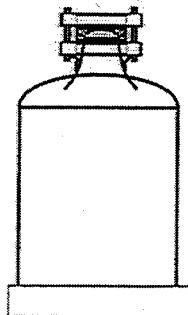


- Capacity of PRV derated by 10%
- Means of isolation for disc replacement
- Inlet pipe size may have to be increased
- Telltale indicator monitoring pressure between disc and PRV

IPD-2004/RS-III-31
RS-31.7-R00-42

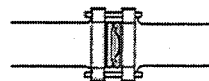
Rupture Disc Sizing

Atmospheric Discharge



Critical Flow Equation

Piping System



Assume Equivalent Length = 75 L/D

IPD-2004/RS-III-32
RS 31.7-R00-43

Material Selection for PRV's

IPO-2004/RS-III-33

Spring-Loaded PRV Metallurgy

- **PRV Sizing and Publisher Tool**
 - Guidance in TD-807-01 (UOP Standard)
 - Choose metallurgy based on corrosion and/or temperature
 - Tool does not check material overrides
 - Consult catalog outside of ranges
- **PRV Network Representative**
- **PRV Specialist**

IPO-2004/RS-III-34

Spring-Loaded PRV Disc Seals

(soft)

- **DON'T specify resilient seals unless needed**
 - Vacuum services
 - Fluids corrosive to relief header
 - Small PRV with screwed connections

IPD-2004/RS-III-35

Pilot-Operated PRV Materials

- **Materials for both pilot and main valves**
- **Both metals and non-metals present**
- **Sizing Tool provides some automatic selections**
- **Can override selections, but**
- **PRV Specialist must check all PORV specifications**

IPD-2004/RS-III-36

Discharge Location

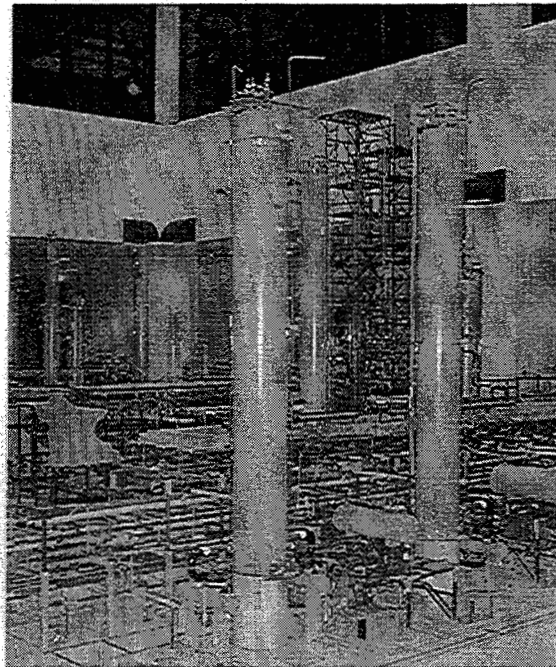
IPD-2004/RS-III-37

BTX Fractionation



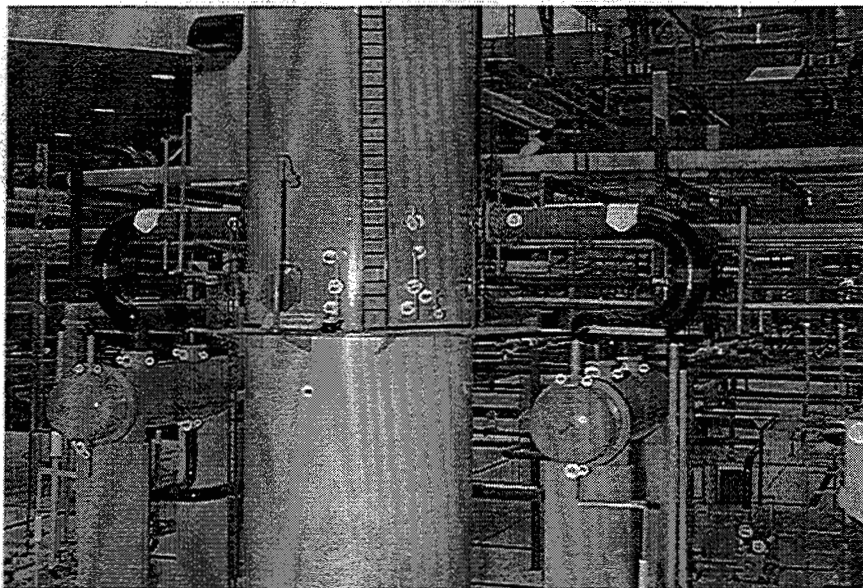
IPD-2004/RS-III-38

■ BTX Fractionation



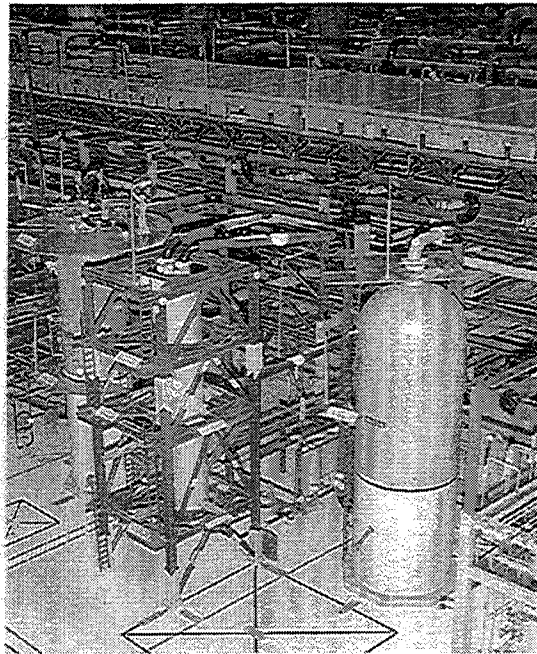
IPD-2004/RS-III-39

Two Horizontal Thermoshipon Reboilers



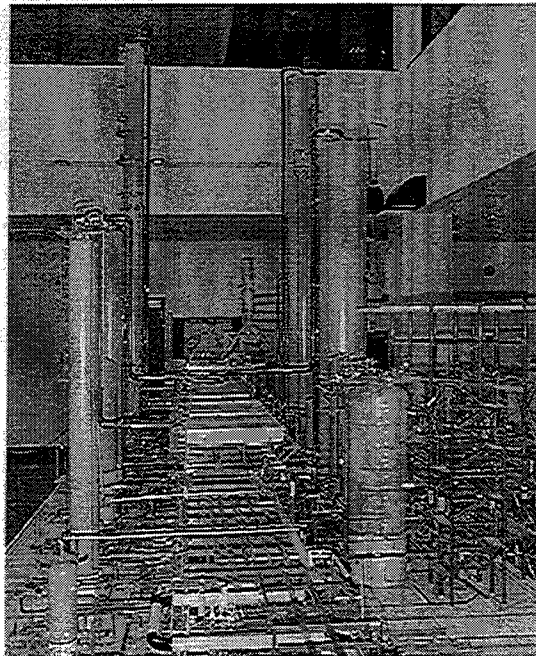
IPD-2004/RS-III-40

- Packinox combined feed exchanger is on the left
- Tatoray Reactor is on the right
- Isomar Separator is on the far left



IPD-2004/RS-III-41

- Xylene Fract and Parex overview
- Sieve Chamber and Rotary valve are front right



IPD-2004/RS-III-42

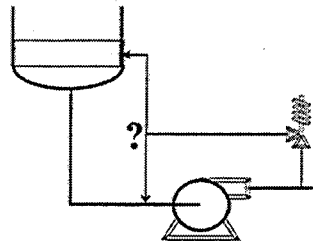
Discharge to Relief Header

- For most HCBN PRV's
- Design Issues
 - Free drainage
 - Back Pressure Limits
 - Rate and Quantity of liquid

IPD-2004/RS-III-43

Discharge to Process Vessel

- Design Issues
 - Back pressure varies
 - Line sizing
 - Discharge system at higher pressure
 - Small pumps: suction line
 - Larger pumps: suction vessel



IPD-2004/RS-III-44

Atmospheric Discharge

- To The Air
 - NOT for HCBN PRV's!
 - Low pressure systems
 - Innocuous materials
 - To a safe location
 - Low-point drain required
- Ground/drain
 - Thermal Expansion of cooling water
 - If HCBN, find a home for it.

IPD-2004/RS-III-45

Other Design Issues

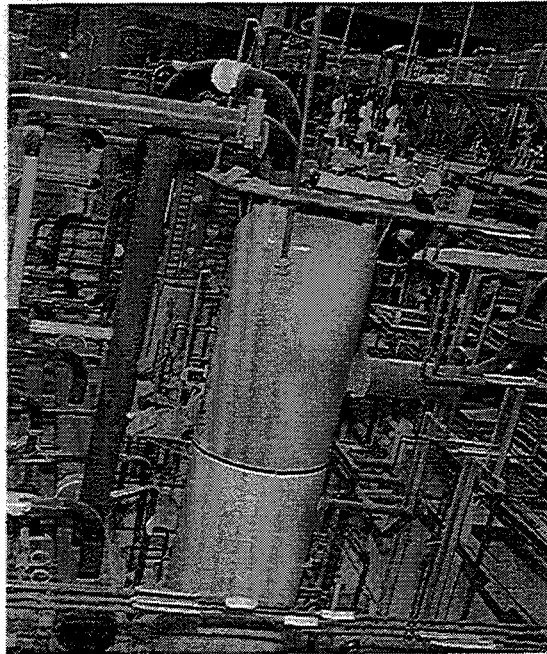
IPD-2004/RS-III-48

Multiple Valves

- Select minimum number of PRVs for a given service
- Select multiples of same size PRV
- At least one valve must be set at design pressure; the other valves can be set up to 5% above design
- Code permits accumulated pressure of 116% if

IPD-2004/RS-III-47

- Platformer Separator
- Dual relief valves with block valves



IPD-2004/RS-III-48

Example

Required Area	54.8 cm ²	180.6 cm ²	419.3 cm ²
Number of Valves	One	Two	Three
Area per Valve	71.3 cm ²	90.3 cm ²	141.9 cm ²
Orifice Designation	6Q8	6R10	8T10
Accumulation	110%	116%	116%

IPD-2004/RS-III-49

Spare Valves

- UOP design practice is not to spare PRVs
- Spare valves may be requested when a customer's experience indicates:

Fouling or plugging of PRV

Severe corrosion

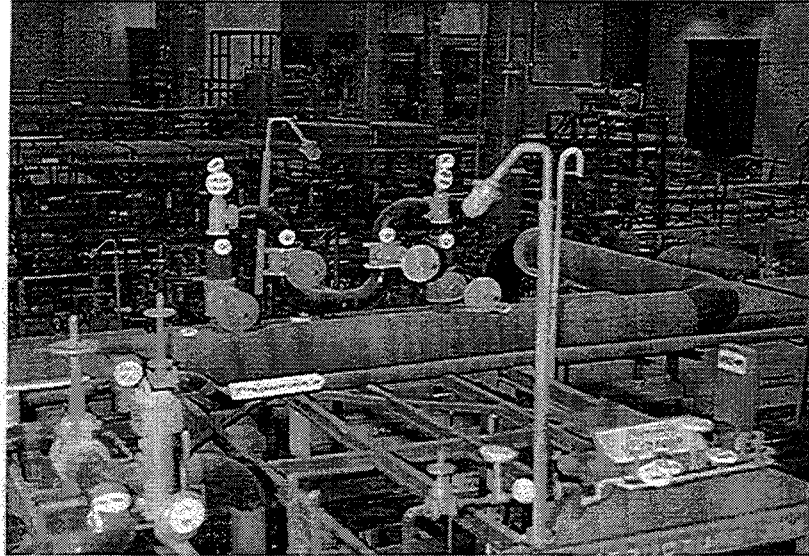
Valve leakage is likely

PRV testing period is shorter than turnaround

- UOP will specify spare PRV in accordance with customer design practice

IPD-2004/RS-III-50

Dual RVs on Ovhd line to air cooler



IPD-2004/RS-III-51

Isolating Block Valves Design Considerations

- Isolating block valves and vents permits maintenance of leaking valve while operating
- Inlet line size must ensure no more than 3% inlet pressure loss
- For multiple valves applications, install one spare set at design pressure

IPD-2004/RS-III-52

Resources - People

- Refinery Specialists
- Vendors

IPD-2004/RS-III-53

Resources Reference Documents

- "Process Safety booklet" (1/23/97)
- Gas Processors Suppliers Association
 - Engineering Data Book
- API RP 520, RP 521, and Std. 2000
 - up-to-date copies held by the PRV Specialists
- AGCO and Crosby Catalogs
 - download (in .pdf format) from
<http://www.andersongreenwood.com/docnet/pressprd.asp>.

IPD-2004/RS-III-54

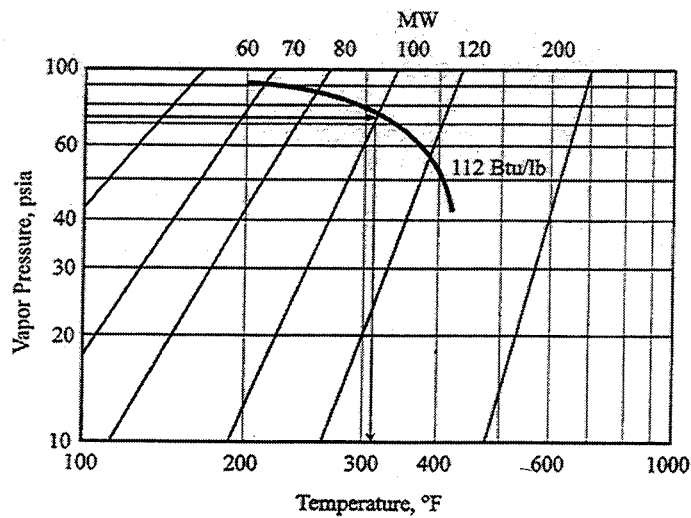


Training Services

Pressure Relief Systems IV

IPD-2004/RS-IV-1

Latent Heat of Vaporization



API RP 521
Fig D-3

RS 31.5-R00-11
IPD-2004/RS-IV-2

Alternative Method (Watson)

$$\lambda_T = \lambda_B \left[\frac{T_c - T}{T_c - T_B} \right]^{0.38}$$

where: λ_T is ΔH_v at relief temperature T

λ_B is ΔH_v at normal boiling point temperature

T_c is critical temperature

T is temperature at relief

T_B is normal boiling point

IPD-2004/RS-IV-3

Vaporization Rate

$$W_{fire} = \frac{Q}{\Delta H_v}$$

- Minimum latent heat of vaporization is 50 Btu/lb
- Above the critical temperature, the expansion rate of the dense fluid is estimated using the vaporization approach

IPD-2004/RS-IV-4

Fire Case Relieving Rate (by volume balance)

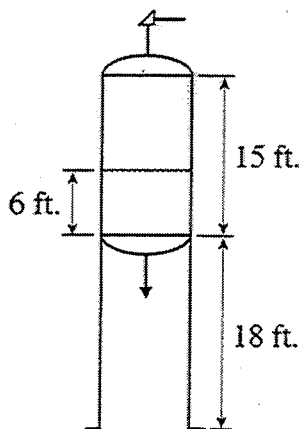
$$W_{fire} = \frac{Q_{fire}}{\Delta H_{vap}} \left\{ 1 - \frac{\rho_{vap}}{\rho_{liq}} \right\}$$

Note: If the fluid approaches critical conditions, the latent heat of vaporization approaches zero and the density ratio becomes one.

At critical conditions, UOP assumes a "pseudo" latent heat equal to 50 Btu/lb.

IPD-2004/RS-IV-5

Fire Case Problem



- Design Pressure = 50 psig
- Vessel Tangent Length = 15 ft.
- Mol. wt. of Liquid is 100
- Vessel Diameter = 5 ft.
- Liquid Height = 6 ft.
- Vessel Elevation = 18 ft.

RS 31.5-R00-14
IPD-2004/RS-IV-5

Fire Case Solution

$$P_{acc} = 121 \% (50) + 14.7 = 75.2 \text{ psia}$$

Wetted Surface = Head & Shell

$$= 1.089 (D)^2 + \pi DL$$

$$= 1.089 (5)^2 + \pi(5)(6) = 121.5 \text{ sq.ft.}$$

$$\text{Heat Input} = 21,000 (1.0)(121.5)^{0.82}$$

$$= 1.075 \times 10^6 \text{ Btu/h}$$

Heat of Vaporization = Use Figure $\frac{D}{A-3}$

$$P = 75.2 \text{ \& MW} = 100$$

$$\text{Then } T = 310^\circ\text{F}$$

$$\Delta H_v = 112 \text{ Btu/lb}$$

$$\text{Vapor Rate} = \frac{1.075 \times 10^6}{112} = 9598 \text{ lb/h}$$

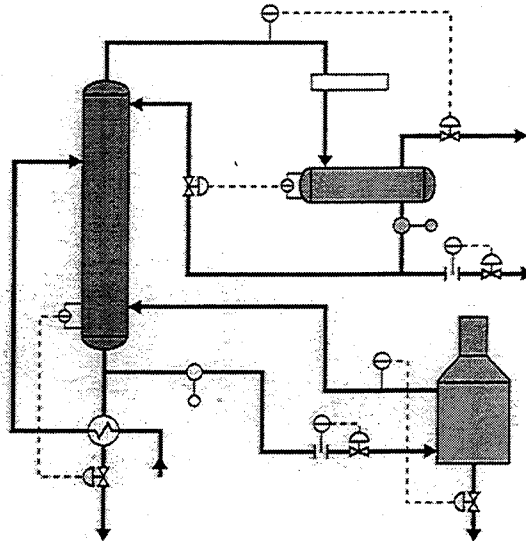
$$\text{Use } 9600 \text{ lb/h}$$

IPD-2004/RS-IV-7

Debutanizer PSV Calculations

IPD-2004/RS-IV-8

Debutanizer



RS 31.6-R00-15
IPD-2004/RS-IV-9

Summary of Relief Scenarios for the Debutanizer

- Electrical Power Failure (Total & Partial)
- Loss of Air-cooled Condenser Duty
- Reflux Failure
- PRC Fails Closed
- Product Separator LLC Fails Open
- External Fire

IPD-2004/RS-IV-10

Electrical Power Failure

- Refinery Wide (Total) Power Failure
- Intermediate (Partial) Power Failure

IPD-2004/RS-IV-11

Total Power Failure Model

Basic Assumptions

- Feed Stops
- Reflux Stops
- No Fractionation Occurs
- Heat Balance Constructed at End of Relieving Case

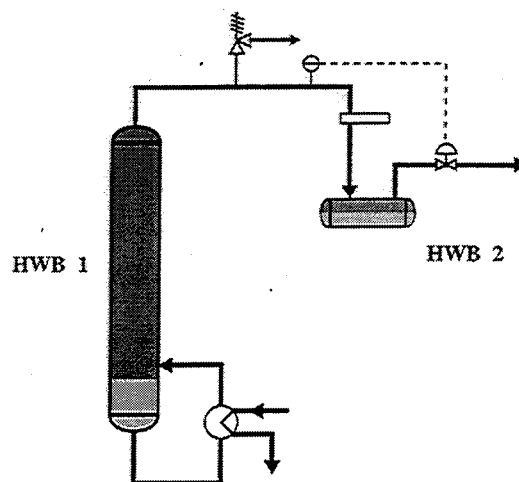
IPD-2004/RS-IV-12

Partial Power Failure

- Pumps on different electrical circuit or on different voltage system
- Various pumps may be located in another process unit
- Pumps driven by a source other than electrical power (steam turbine, gas turbine)

IPD-2004/RS-IV-13

Debutanizer (Power Failure Model)



RS 31.6-P00-16
IPD-2004/RS-IV-14

Power Failure HWB I

	<u>Mass</u>	<u>Heat</u>
Reboiler Duty		$+ Q_{REB}$
Vaporized Inventory	$- W_I$	$- W_I h_v$
Depleted Liquid	$+ W_I$	$+ W_I h_l$
Total	0	0

IPD-2004/RS-IV-15

Power Failure (working equations)

$$0 = Q_{REB} - W_I h_v + W_I h_l$$

$$W_I = \frac{Q_{REB}}{\Delta H_v}$$

Vaporization rate based on simultaneous solution
of heat and mass balance

IPD-2004/RS-IV-16

Power Failure Relieving Rate

Residual Condenser Credit (natural convection):

$$W_{cond} = \frac{Q_{cond}}{\Delta H_v}$$

Relieving Rate (difference between vaporization rate and residual condensation rate):

$$W_{rel} = W_I - W_{cond}$$

IPD-2004/RS-IV-17

Loss of Air-Cooled Condenser Duty

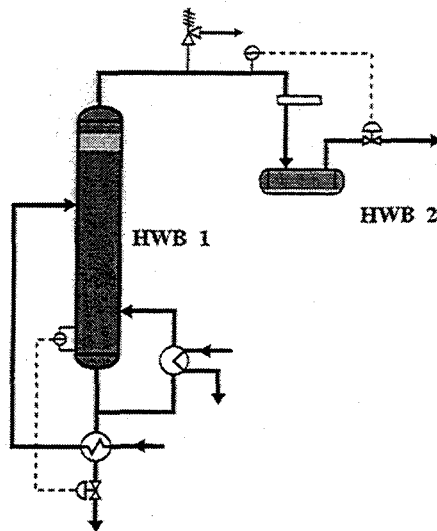
IPD-2004/RS-IV-18

Loss of Condenser Duty

- PRC opens wide to attempt pressure control
- Debutanizer pressure increases in spite of controller action
- Liquid inventory in overhead receiver decreases
- LLC on overhead receiver stops reflux by resetting FRC on reflux
- Heavy material migrates to top section as liquid contacting in top section of debutanizer stops

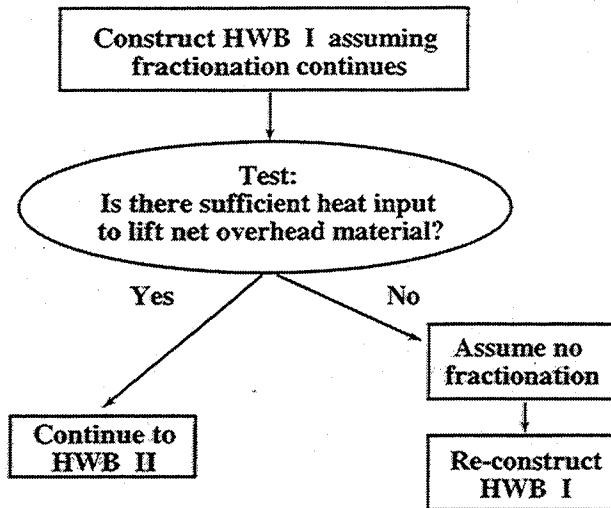
IPD-2004/RS-IV-19

Debutanizer (loss of condenser duty)



RS 31.6-R00-17
IPD-2004/RS-IV-20

Decision Tree



IPD-2004/RS-IV-21

Assume Fractionation Continues

- Feed, net overhead, and net bottoms mass rates are constant
- Composition of net streams is unchanged; thus dew and bubble points at relief can be determined

NOTE: If a heat and mass balance determines that this assumption is invalid, the calculational model for loss of condenser duty will have to be modified.

IPD-2004/RS-IV-22

Condenser Failure HWB I

	<u>Mass</u>	<u>Heat</u>
Feed	$+W_f$	$+W_f h_f$
Net Bottoms	$-W_{nb}$	$-W_{nb} h_{nb}$
Reboiler Duty		$+Q_{reb}$
Net Overhead Vapor	$-W_{no}$	$-W_{no} h_{no}$
Vaporized Inventory	$-W_I$	$-W_I h_v$
Depleted Liquid Inventory	$+W_I$	$+W_I h_l$
Total	<u>0</u>	<u>0</u>

IPD-2004/RS-IV-23

Condenser Failure HWB I

	<u>Mass</u>	<u>Heat</u>
Feed		
Net Bottoms		
Reboiler Duty		
Net Overhead Vapor		
Vaporized Inventory		
Depleted Liquid Inventory		
Total	<u>0</u>	<u>0</u>

IPD-2004/RS-IV-24

Vaporized Liquid Inventory ***(working equation)***

$$0 = W_F h_F - W_{NB} h_{NB} + Q_{REB} - W_{NO} h_{NO} - W_I h_v + W_I h_l$$

$$W_I = \frac{(W_F h_F - W_{NB} h_{NB} + Q_{REB} - W_{NO} h_{NO})}{\Delta H_v}$$

Inventory rate must have a positive sign for model assumption that fractionation continues to be valid

IPD-2004/RS-IV-25

Required Calculations

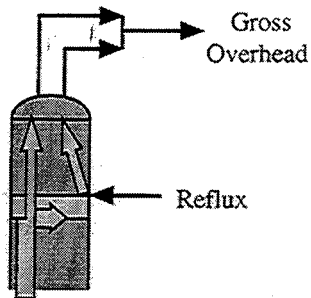
Determine:

- Reboiler Duty at Relief
- Heat Content of Net Streams
- Vaporization Rate of Liquid Inventory
- Overhead Condenser Credit

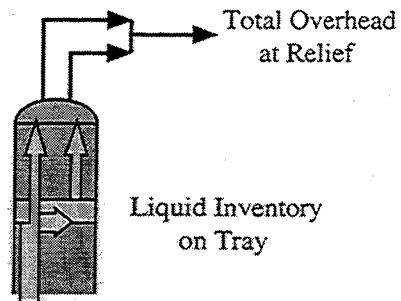
IPD-2004/RS-IV-26

Reflux vs. Liquid Inventory

Normal Operation



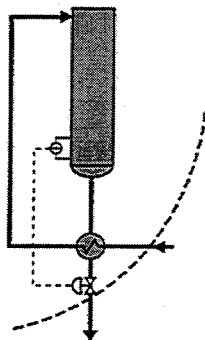
Reflux Stops



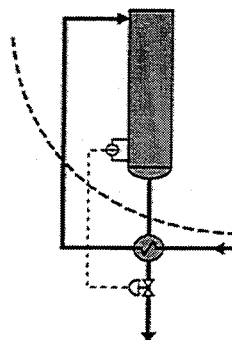
RS 31.6-R00-18
IPD-2004/RS-IV-27

Heat Content of Net Bottoms (leaving the envelop)

Before F/B exchanger



After F/B exchanger



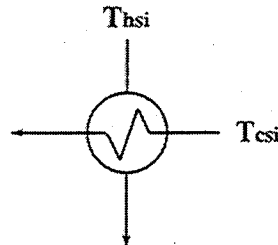
**F/B exchanger
must be rated
regardless of
heat & mass
envelop**

RS 31.6-R00-19

IPD-2004/RS-IV-28

Feed/Bottoms Exchanger Duty

$$Q_{rel} = Q_{nor} \left\{ \frac{(T_{hsi} - T_{csi})_{rel}}{(T_{hsi} - T_{csi})_{nor}} \right\}$$



Assumes:

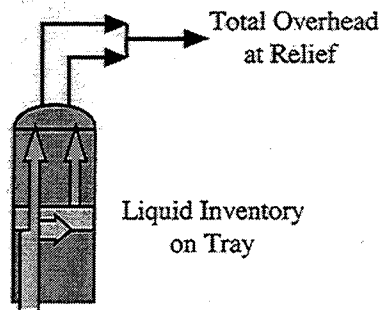
Straight line enthalpy profiles.

Similar thermal effectiveness factor.

Constant overall heat transfer coefficient.

IPD-2004/RS-IV-29

Overhead Vapor Rate



The overhead vapor rate leaving the debutanizer at relief is made up of two parts:

- a) Net overhead vapor**
- b) Vaporized liquid inventory**

RS 31.6-R00-18
IPD-2004/RS-IV-30

Condenser Failure HWB II

	<u>Mass</u>	<u>Heat</u>
Overhead Vapor	$+W_o$	$+W_o h_v$
Relieving Rate	$-(W_o - W_A)$	$-(W_o - W_A)h_v$
Condenser Duty		$-Q_{cond}$
Accum. Liq. Inventory	$-W_A$	$-W_A h_l$
Total	0	0

IPD-2004/RS-IV-31

Condenser Failure HWB II

	<u>Mass</u>	<u>Heat</u>
Overhead Vapor		
Relieving Rate		
Condenser Duty		
Accum. Liq. Inventory		
Total	0	0

IPD-2004/RS-IV-32

Loss of Condenser Duty Relieving Rate

Residual condenser duty (natural convection):

$$W_{cond} = \frac{Q_{cond}}{\Delta H_v}$$

Relieving rate (difference between total vapor rate leaving debutanizer and residual condensation):

$$W_{rel} = W_{total} - W_{cond}$$

IPD-2004/RS-IV-33

Reality Check

- Always check the reasonableness of the calculated relieving rates for each case
 - During a reflux failure, loss of condenser duty, or PRC failing closed, the relieving rate should be similar to the normal gross overhead rate
 - During an electrical power failure, the relieving rate should be similar to the normal reboiler vapor rate
- If the rates differ significantly from that given above, rationalize the difference, i.e., dramatic loss of reboiler duty due to low LMTD

IPD-2004/RS-IV-34

Alternate Calculation (when model assumptions are not valid)

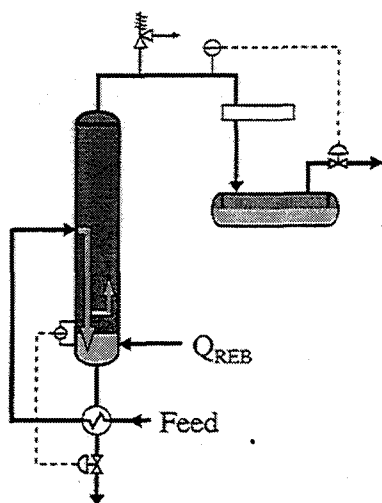
If the "excess" heat is negative, the assumption that fractionation continues as normal is not valid.

A negative value means that there is not sufficient heat to vaporize the net overhead material. Thus, the bubble and dew points used to determine the heat balance are not valid.

Although some fractionation will occur, the problem is bracketed by assuming no fractionation occurs, i.e., only flashing occurs in the column.

IPD-2004/RS-IV-35

No Fractionation Occurs



1. Assume a bottoms temperature.
2. Calculate reboiler duty using assumed temperature.
3. Calculate feed-bottoms exchanger duty using assumed temperature.
4. Flash feed at normal feed enthalpy plus reboiler and feed-bottoms exchanger duty.
5. Compare calculated temperature to assumed temperature.
6. Vapor rate leaving column is vapor rate from the flash.

IPD-2004/RS-IV-36
RS 31.6-R00-21

Receiver PRC Fails Closed

- Net vapor cannot exit overhead system and it blankets the overhead condenser
- Lose overhead condenser duty
- Debutanizer pressure increases
- LLC on overhead receiver stops reflux by resetting FRC
- Heavy material migrates to top section as liquid contacting in top section of debutanizer stops
- Net overhead liquid continues until pump loses positive suction

IPD-2004/RS-IV-37

Separator LLC Fails Opened

- Initially, liquid feed rate to column doubles and floods the bottoms section of debutanizer
- Subsequently, reactor vapors enter debutanizer and blanket the overhead condenser
- Lose overhead condenser duty
- PRC opens wide to attempt pressure control

IPD-2004/RS-IV-38

Separator LLC Fails Opened

(continued)

- Debutanizer pressure increases in spite of controller action
- LLC on overhead receiver stops reflux by resetting FRC
- Heavy material migrates to top section as liquid contacting in top section of debutanizer stops
- Net overhead liquid continues until pump loses positive suction

IPD-2004/RS-IV-39

Debutanizer Condenser Failure

- Determine Accumulated Pressure

$$P_{ACC} = \text{Set Pressure} \times \text{Allowable Accumulation} + 14.7 \text{ psi}$$
$$300 \times 110\% + 14.7 = 344.7 \text{ psia}$$

- Determine Reboiler Duty

$$Q_{\text{relief}} = Q_{\text{normal}} = 10.59 \times 10^6 \text{ Btu/h}$$

IPD-2004/RS-IV-40

Debutanizer Condenser Failure (continued)

- Use Cox Chart to get top and bottom temperatures at accumulated pressures

Top Temperature	Normal	161°F
	@	265 psig
	Relief	181°F
Bottom Temperature	@	345 psia
	Normal	415°F
	@	272 psig
	Relief	435°F
	@	352 psia

IPO-2004/RS-IV-41

Debutanizer Condenser Failure (continued)

- Net Bottoms Enthalpy at Relief

$$Q_R = Q_N + W_B \times C_P (T_R - T_N)_B$$

Calculate C_P of bottoms from exchanger data

$$C_P = \frac{9 \times 10^6 \text{ Btu/h}}{50138 \text{ lbs/h} \times (415^\circ\text{F} - 138^\circ\text{F})} = 0.65 \frac{\text{Btu}}{\text{lb}^\circ\text{F}}$$

$$Q_R = 10.91 \text{ MM} + 50138(0.65)(435^\circ\text{F} - 415^\circ\text{F})$$

$$Q_R = 10.91 + 0.65 = 11.56 \times 10^6 \text{ Btu/h}$$

IPO-2004/RS-IV-42

Debutanizer Condenser Failure ***(continued)***

■ Rerate Feed/Bottoms Exchanger

$$\begin{aligned} Q_R &= Q_N \frac{(T_H - T_C)_{REL}}{(T_H - T_C)_N} \\ &= 9.0 \times 10^6 \frac{(435 - 108)}{(415 - 108)} = 9.0 \times 10^6 \times \frac{327}{307} \\ &= 9.59 \times 10^6 \text{ Btu/h} \end{aligned}$$

IPD-2004/RS-IV-43

Debutanizer Condenser Failure ***(continued)***

■ Enthalpy of Net Bottoms

$$\begin{aligned} H_R &= Q_R(\text{Bottoms}) - Q_R(\text{Exchangers}) \\ &= 11.56 \times 10^6 - 9.59 \times 10^6 \\ &= 1.97 \times 10^6 \text{ Btu/h} \end{aligned}$$

IPD-2004/RS-IV-44

Debutanizer Condenser Failure (continued)

■ Enthalpy of Net Overhead

$$W_O = W_V + W_L = 5169 + 17505 = 22674 \text{ lb/h}$$

$$H_R = H_N + W_O C_P (T_R - T_N)$$

$$C_P = 0.19 \text{ Btu/lb } ^\circ\text{F}$$

$$\text{For } H_N \text{ use } 180 \text{ Btu/lb} \times W_O$$

$$= 180 \times 22675 \text{ lb/h}$$

$$= 4.08 \times 10^6 \text{ Btu/h}$$

$$H_R = 4.08 \times 10^6 + 22674 (19)(181 - 161)$$

$$= 4.17 \times 10^6 \text{ Btu/h}$$

IPD-2004/RS-IV-45

Debutanizer Condenser Failure (continued)

■ Calculate Depletion by Solving Heat Balance

	<u>Mass, lb/h</u>	<u>Heat Balance</u>
Feed	+ 72812	+ 1.96
Net Bottoms	- 50138	- 1.97
Reb. Duty	-	+ 10.59
Gross Overhead Vapor		
Net Vapor	- 22674	- 4.17
Depletion as Vapor	- W_D	- $W_D H_D(\text{Vap})$
Depletion as Liquid	+ W_D	+ $W_D H_D(\text{Liq})$
	<u>0</u>	<u>$6.41 - W_D(H_V - H_L)$</u>
		ΔH_D
		= 0

IPD-2004/RS-IV-46

Debutanizer Condenser Failure (continued)

$$H_D(\text{Vap}) = 184 \text{ Btu/lb}$$

$$H_D(\text{Liq}) = 83 \text{ Btu/lb}$$

$$\Delta H_D = 184 - 83 = 101 \text{ Btu/lb}$$

$$W_D = 6.41 \times 10^6 / 101 = 63465 \text{ lb/h}$$

$$W_D H_D(\text{Liq}) = 83(63465) = 5.27 \text{ mm Btu/h}$$

$$W_D H_D(\text{Vap}) = 184(62773) = 11.68 \text{ mm Btu/h}$$

IPD-2004/RS-IV-47

Debutanizer Condenser Failure (continued)

■ Complete Balance

	<u>Mass, lb/h</u>	<u>Heat, 10⁶ Btu/h</u>
Feed	+ 72812	+ 1.96
Net Bottoms	- 50138	- 1.97
Reboiler	-	+ 10.59
Gross Overhead Vapor		
Net Vapor	- 22674	- 4.17
Depletion as Vapor	- 63465	- 11.68
Depletion as Liquid	<u>+ 63465</u>	<u>+ 5.27</u>
	0	0

IPD-2004/RS-IV-48

Debutanizer Condenser Failure ***(continued)***

- Consider Second Envelope
 - Overhead vapor to envelope

$$W_o = 22674 + 63465 = 86139 \text{ lb/h}$$

$$H_D = 4.17 \times 10^6 + 11.68 \times 10^6 = 15.85 \times 10^6 \text{ Btu/h}$$

- Condenser duty = 20% of normal

$$Q_{REL} = 0.20(9.36 \times 10^6) = 1.87 \times 10^6 \text{ Btu/h}$$

IPD-2004/RS-IV-49

Debutanizer Condenser Failure ***(continued)***

- Determine Condensation Rate

$$W_C = \frac{Q_R}{\Delta H_D} = \frac{1.87 \times 10^6}{101}$$

$$= 18535 \text{ lb/h}$$

$$H_C = (18535)(83) = 1.54 \times 10^6 \text{ Btu/h}$$

IPD-2004/RS-IV-50

Debutanizer Condenser Failure (continued)

■ Calculate Relieving Rate

$$\begin{aligned} W_R &= W_O - W_C \\ &= 86139 - 18535 \\ &= 67604 \text{ lb/h} \end{aligned}$$

$$\begin{aligned} H_R &= (67604)(184) \\ &= 12.44 \times 10^6 \text{ Btu/h} \end{aligned}$$

IPD-2004/RS-IV-51

Debutanizer Condenser Failure (continued)

■ Check Balance

	<u>Mass, lb/h</u>	<u>Heat, 10⁶ Btu/h</u>
Overhead Vapor	+ 86139	+ 15.85
Relieving Rate	- 67604	- 12.44
Accum. Liquid	- 18535	- 1.54
Net Vapor	-	-
Net Liquid	-	-
Reflux	-	-
Condenser	-	-1.87
	<u>0</u>	<u>0</u>

$$\text{TR} = 181\text{F}; \text{MW} = 46.1; \text{WR} = 66942 \text{ lb/h}$$

IPD-2004/RS-IV-52

UOP LLC

25 East Algonquin Road
Des Plaines, Illinois 60017-5017
USA

HEAT AND WEIGHT BALANCES

PAGE NO.: _____

DATE: _____

PROJECT: _____

BY: _____

ENGINEERING DEPARTMENT
DEBUTANIZER

Note:

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STREAM	°API (lb/gal)	UOP K	Mol Weight	Flowing lb/cf* S.G.	Temp., °F	Pressure psig (mmHg)	mm scfs* bpsd	Flowing cfs gpm	Mol/h	Lb/h	Btu/lb	10 ⁶ Btu/h
LP Flash Drum Liquid to Pump	87.1	12.67	67.7	0.624	108	260	7708	233	1075.05	72812	27	1.96
Flash Drum Liquid to Control Valve	87.1	12.67	67.7	0.624	108	300	7708	233	1075.05	72812	27	1.96
Flash Drum Liquid to Fd/Btms Exch	87.1	12.67	67.7	0.624	108	293	7708	233	1075.05	72812	27	1.96
Feed/Bottoms Exchanger Duty												9.00
Feed from Feed/Bottoms Exchanger			67.7		266	283			1075.05	72812	151	10.96
Total Vapor			49.7	2.253*	266	283	3.65*	2*	401.24	19945	236	4.70
Inerts					266	283			0.02	0		
Hydrogen					266	283			2.07	4		
Hydrocarbon					266	283			399.15	19940		
Total Liquid	71.0	12.08	78.5	0.580	266	283	5187	182	673.81	52868	118	6.26
Feed to Debutanizer			67.7		264	269			1075.05	72812	151	10.96
Total Vapor			49.9	2.151*	264	269	3.79*	3*	416.14	20786	235	4.89
Inerts					264	269			0.02	0		
Hydrogen					264	269			2.08	4		
Hydrocarbon					264	269			414.04	20781		
Total Liquid	70.4	12.05		0.584	264	269	5087	178	658.91	52027	117	6.07
Gross Ovhd Vapor to Air Cooled Cond			46.1	2.541*	161	265	13.03*	7*	1430.87	65967	180	11.88
Inerts					161	265			0.08	1		
Hydrogen					161	265			2.44	5		
Hydrocarbon					161	265			1428.36	65960		
Air Cooled Condenser Duty												9.36
Material to Overhead Receiver			46.1		100	260			1430.87	65967	38	2.52
Total Vapor			35.0	1.924*	100	260	1.34*	1*	147.60	5170	152	0.79
Inerts					100	260			0.01	0		
Hydrogen					100	260			2.03	4		
Hydrocarbon					100	260			145.56	5165		
Total Liquid	(4.351)	14.58	47.4	0.489	100	260	7965	249	1283.27	60797	29	1.74
Net Vapor Recycle to Feed Junction			35.0	1.925*	100	260	1.34*	1*	147.58	5169	152	0.79
Inerts					100	260			0.01	0		
Hydrogen					100	260			2.02	4		
Hydrocarbon					100	260			145.55	5165		

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USA

HEAT AND WEIGHT BALANCES

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STREAM	°API (lb/gal)	UOP K	Mol Weight	Flowing lb/cf* S.G.	Temp., °F	Pressure psig (mmHg)	mm scfs* bpsd	Flowing cfs* gpm	Mol/h	Lb/h	Btu/lb	10 ⁶ Btu/h
Net Vapor to Battery Limit			35.0	0.418*	70	50	1.34*	3*	147.58	5169	152	0.79
Inerts					70	50			0.01	0		
Hydrogen					70	50			2.02	4		
Hydrocarbon					70	50			145.55	5165		
Overhead Receiver Liquid to Pump	(4.351)	14.58	47.4	0.489	100	260	7965	249	1283.29	60797	29	1.74
Receiver Liquid from Pump	(4.351)	14.58	47.4	0.490	100	300	7965	248	1283.29	60797	29	1.74
Net Overhead Liquid to Control Valve	(4.351)	14.58	47.4	0.490	100	300	2293	71	369.50	17505	29	0.50
Net Overhead Liquid to Battery Limit	(4.351)	14.58	47.4	0.489	100	260	2293	72	369.50	17505	29	0.50
Reflux to Control Valve	(4.351)	14.58	47.4	0.490	100	300	5672	177	913.79	43292	29	1.24
Reflux from Control Valve	(4.351)	14.58	47.4	0.489	100	285	5672	177	913.79	43292	29	1.24
Reflux to Column	(4.351)	14.58	47.4	0.489	100	265	5672	177	913.79	43292	29	1.24
Total Bottoms from Column	58.9	11.65	89.9	0.521	415	272	21061	877	2542.11	228424	218	49.69
Net Bottoms to Feed/Bottoms Exch	58.9	11.65	89.9	0.521	415	272	4623	192	558.00	50140	218	10.91
Bottoms to Pump	58.9	11.65	89.9	0.521	415	272	16438	684	1984.11	178285	218	38.79
Bottoms to Control Valve	58.9	11.65	89.9	0.524	415	317	16438	680	1984.11	178285	218	38.79
Bottoms to Convection Section of Htr	58.9	11.65	89.9	0.522	415	292	16438	682	1984.11	178285	218	38.79
Debutanizer Reboiler Duty												10.59
Reboiler Return to Column			89.9		446	272			1984.11	178285	277	49.38
Total Vapor			85.6	3.515*	446	272	9.48*	7*	1040.53	89089	322	28.72
Total Liquid	51.9	11.36	94.5	0.534	446	272	7922	334	943.58	89196	232	20.66
Net Bottoms to Feed/Bottoms Exch	58.9	11.65	89.9	0.521	415	272	4623	192	558.00	50138	218	10.91
Feed/Bottoms Exchanger Duty												9.00
Net Bottoms to Product Cooler	58.9	11.65	89.9	0.708	138	252	4623	142	558.00	50138	38	1.91
Bottoms Cooler Duty												0.96
Net Bottoms to Control Valve	58.9	11.65	89.9	0.728	100	247	4623	138	558.00	50138	19	0.95
Net Bottoms to Battery Limit	58.9	11.65	89.9	0.726	100	50	4623	138	558.00	50138	19	0.95

INTRODUCTION TO PROCESS SAFETY

Process safety in a modern refinery is the concern of every employee. Managers, process engineers, operators, and maintenance personnel should be aware of minimizing the risks involved in the operation of a refinery or petrochemical unit processing combustible materials. This section will deal with process safety, mainly from the viewpoint of the process designer. The responsibilities of the process designer are:

- Design the process unit to minimize risk
- Know the regulations and standards for process safety
- Recognize causes of overpressure
- Determine relieving rates based upon possible events
- Design the relieving device and relief system

POTENTIAL HAZARDS

The potential hazards in a refinery can be stated simply as **explosion**, **fire**, and **toxic** release. These hazards are life-threatening, and the resultant damage is expensive to repair. Everyone can certainly relate to the tragedies of Bhopal and Chernobyl. Why did these events occur? There is usually not one event, but a combination of events leading to these disasters. The likely causes fall into the following classifications:

- Human error
- Equipment failure
- Loss of utilities
- Uncontrolled reactions
- External fire

DESIGN AND PLANNING FOR SAFETY

For a given well-designed process unit, there are a number of protective layers. Each of these layers is designed to minimize risk.

Process Design

The inherent safety aspects of the process unit start with an accurate representation of the heat, mass, and mol balance of the unit at normal operation. Additionally, the process configuration of equipment aids in minimizing risk.

Basic Process Controls

The basic process controls should be designed to maintain the desired safe operating conditions and return the unit to normal operating conditions after a process upset.

Alarms and Operator Supervision

Any deviations from normal conditions, or deviations which could signal unsafe conditions, are brought to the attention of the operation personnel.

Shutdowns

Any deviations which signal a potential risk and are not corrected or answered will activate a system of shutdowns in a fail-safe action.

Pressure Relief Systems

Pressuring relieving devices are designed and installed to vent the system pressure when it exceeds the design pressure of equipment by a finite amount. The discharge from the relieving devices go to a safe location.

Emergency Response

In case an explosion, fire, or toxic release does occur, the refinery must provide contingency plans for minimizing risk to personnel and property within the refinery. The community adjacent to the refinery must provide equal contingency plans to protect the community.

SAFETY MANAGEMENT

Process safety management continues after the unit is started up and safely operating. Engineers trained in safety management are required to oversee and manage a program of safety training, oversee the routine inspection of process equipment, determine maintenance requirement of equipment, generate the safety procedures for work required, oversee the testing required to insure safe work conditions, and inform personnel of changes necessary to maintain a risk free environment.

REGULATIONS AND STANDARDS

In the U.S., the following items comprise the publications that define the requirements and recommended practices for process safety.

- OSHA 1910.119
- SARA Title III
- Clean Air Act
- API Recommended Practices
- Local and state codes and regulations
- AIChE "Center for Process Safety" publication

OSHA 1910.119

In the oil and gas industry, OSHA regulations dictates the requirements to insure personnel safety and health.

This regulation requires a process hazard analysis, publication of process safety information, documentation of operating procedures, and training of operations personnel. Other provisions involve maintenance, planning, and changes.

There is a schedule for compliance to the regulations of OSHA by the owners of the process units. In 1993, 100% of all new plants shall comply and, for existing plants, 100% shall comply by 1997.

The major provision, which affects the refining industry, is the requirement for conducting and documenting a process hazard analysis. There are several methodologies that meet this requirement, and the characteristics of each method vary and give somewhat different results.

- **Checklist Analysis**

This method uses a specific written list of known hazards, design deficiencies and potential risks to evaluate materials, equipment, and procedures. This method is most often used in industries, which have significant experience. Used in past as part of the P&ID check.

- **"What If" Analysis**

This method uses a team which has expertise in design, operation, and safety. A leader generates a list of "what if" questions and the team identifies potential hazards, potential consequences, and recommends safeguards and solutions. Answers to all questions are documented.

- **Failure Modes and Effects**

In this method, an analyst lists all the failure modes (conceivable malfunctions) for each separate piece of equipment. A team then evaluates the ways equipment can fail or misoperate. This method is equipment oriented.

- **HAZOP**

This method divides the process unit into manageable sections or nodes and evaluates the significance of process deviations from design intent. A facilitator supplies key guidewords and a team determines the causes, consequences, safeguards, and a recommendation for each action. A qualitative risk is assigned to each action based upon frequency and consequence of the action. The HAZOP study is fast becoming the preferred methodology of the refining

industry. A comprehensive HAZOP can become very tedious and time consuming. The documentation is often as lengthy as the process specifications.

- **Fault Tree Analysis**

This method goes one step further and statistically assigns the quantitative risk for each action. The team now becomes involved in determining the probability of many actions occurring over some specified time period. These actions could include unrelated events. This type of analysis started in the nuclear industry.

ROLE OF PROCESS DESIGN ENGINEER

The process design engineer is most knowledgeable in all major aspects of the design including equipment configuration, heat balance, reactor kinetics, instrumentation, exchanger network configuration, relationship to upstream or downstream units, utility system, and applicable codes and standards. With this range of knowledge, the process design engineer is best qualified to recognize the causes of overpressure, process deviations, and the relationship between events. It is important to consider all these steps when creating a scenario to simulate an overpressure situation.

Cause of Overpressure

The major causes of overpressure are equipment failure or malfunction, loss of utilities (power, steam, cooling water, instrument air), human errors, external fire, and uncontrolled reactions.

Process Deviations

Process deviations occur when the equipment or unit balance has an unbalance in mass, heat, or molal composition.

Relationship Between Events

- **Isolated Occurrences**

There is always a remote chance several unrelated events occur which cause a mishap that cannot be predicted with logic. Bhopal is an example. Instead, fault tree analysis is required

to determine a distinct mishap possibility with a small probability of happening. Earthquakes, sabotage, or war could be factored into this scenario.

- **Simultaneous Events**

One event could simultaneously effect considerable equipment in a process unit. For example, a power failure - this could shutdown compressors, pumps, and air cooler fans. Each of these secondary events must be considered for a cause of overpressure.

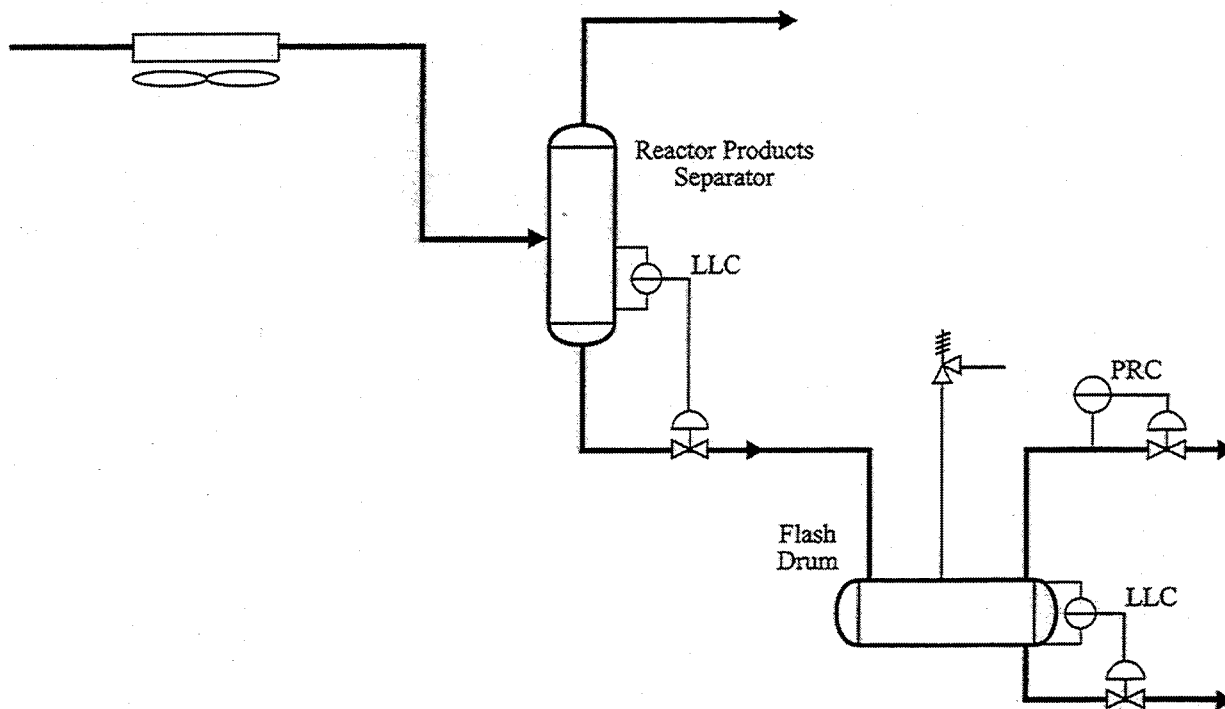
- **Sequential Events**

Sometimes one event creates a scenario which triggers a domino effect. For example, the loss of a recycle gas compressor: the quench gas to an exothermic reactor stops, the reactor temperature rises, and an uncontrolled reaction takes place. The time framework for this series of sequential events would have to be calculated based upon multiple assumptions involving heat transfer and reactor kinetics.

HYDROGENATION UNIT FLASH DRUM

For the flash drum shown in Figure XV-1, determine the events which would cause a change in the mass, heat, or mol balance. The flash drum is assumed operating at design conditions in a steady state condition.

Figure XV-1



IPD-R00-01

DETERMINATION OF RELIEVING RATES

Guidelines

There are several excellent sources of help to calculate a relieving rate. The ASME Code lists the equations to be used for relieving rates, the requirements for use of rupture discs, the instrumentation required for system integrity, and the requirements for derating a system. There are two applicable sections: Section I (Steam Systems) and Section VIII (Pressure Vessels). There is a non-mandatory section concerning items such as maximum pressure loss from a relieving device and use of block valves under a relieving device.

Every engineer required to design a relief valve should have API Recommended Practice 520, 521, and 2000 booklets. These cover guidelines and design practice for pressure relieving devices and systems.

The AIChE Center for Chemical Process Safety has many publications related to all aspects of process safety management.

Heat and Material Balance

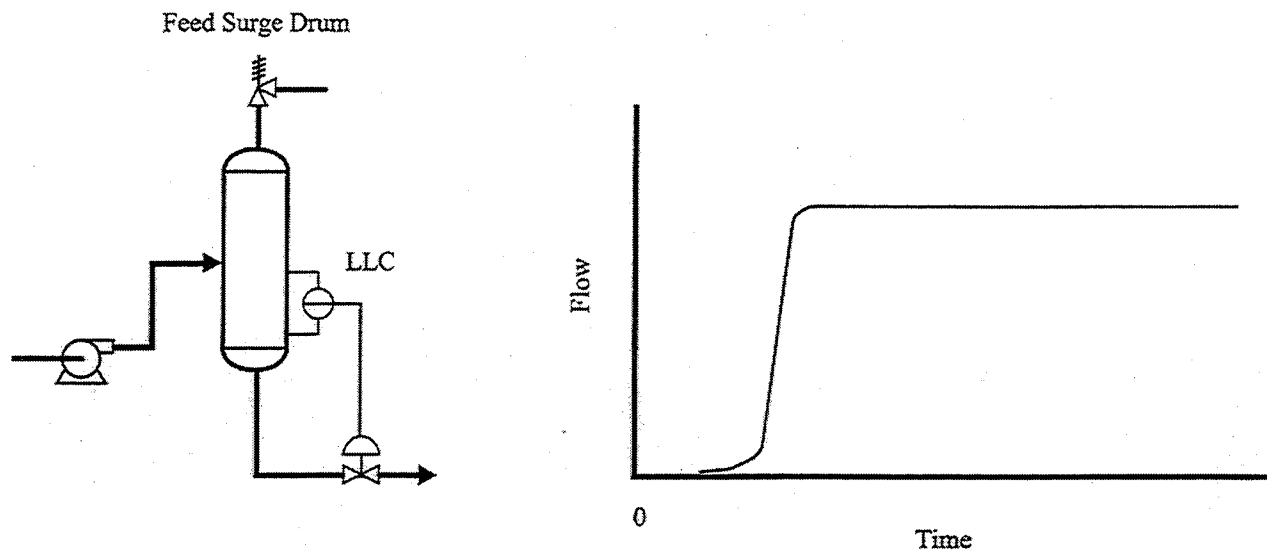
The process specifications which UOP issues for a process unit have the heat and material balance shown at operating conditions at steady state conditions. The in "streams" must exactly equal the out "streams". For the transient conditions required to model a unit for emergency or overpressure conditions, some type of numerical integration of variables versus time is required. Mathematically, the use of differential equations is necessary for this analysis. Fortunately, the calculations can be simplified by looking at the model to obtain a snapshot in time. This requires only the use of algebra to do an instantaneous balance. The following unsteady state equation at a point in time is used:

$$\text{In} + \text{Depletion} = \text{Out} + \text{Accumulation}$$

Where does one take the snapshot for Figures XV-2, 3, and 4? The selected time for the snapshot is essential to determining the maximum rate.

Figure XV-2

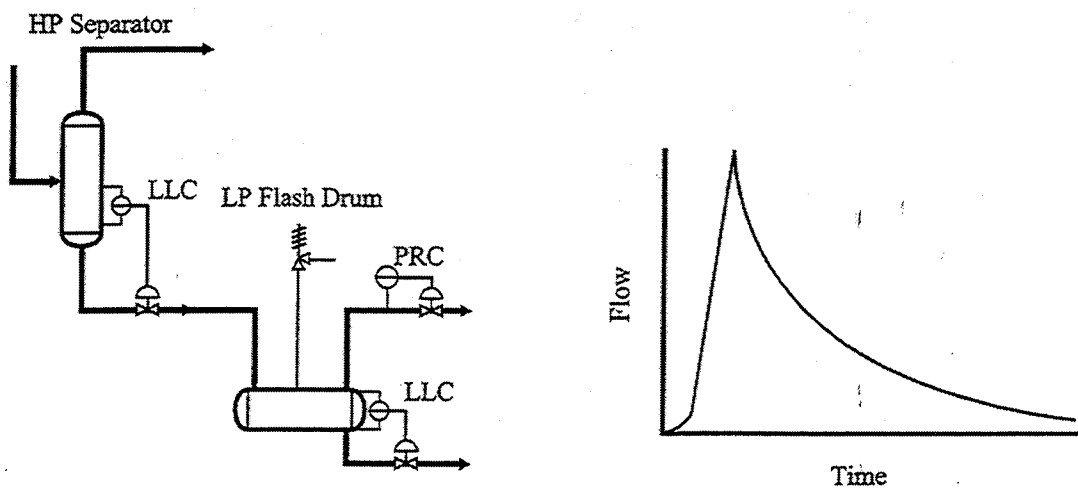
Blocked Outlet Transient



IPD-R00-02

Figure XV-3

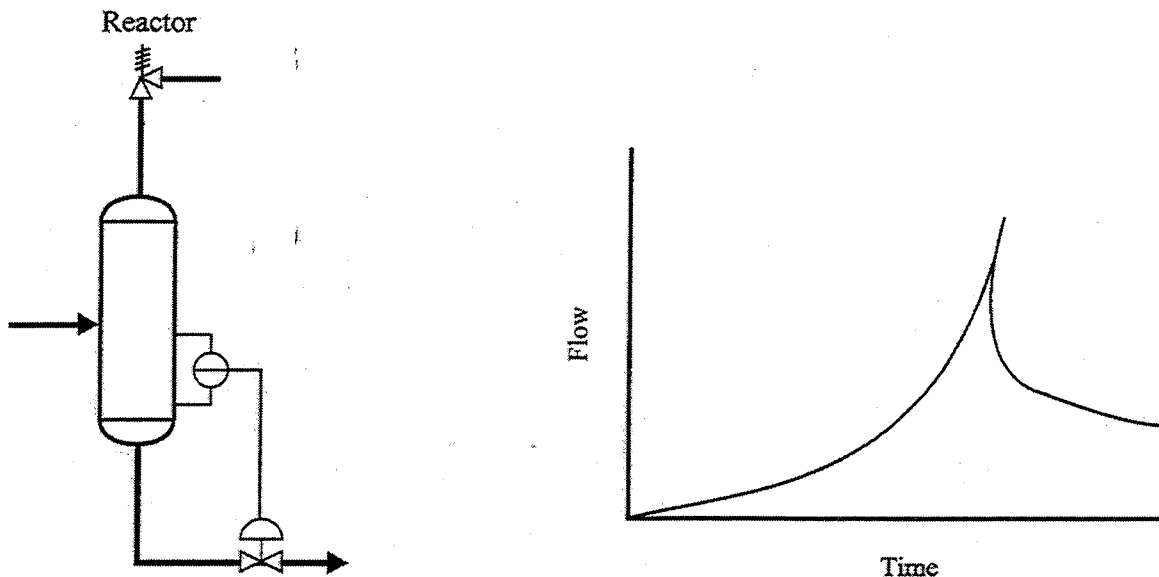
Control Valve Failed Open



IPD-R00-03

Figure XV-4

Runaway Chemical Reaction



IPD-R00-04

Creation of a Model

A thorough knowledge of the system is necessary as a starting point to create a model. One needs to question what equipment is involved within the system. For example, in order to model a fractionator relief valve, the column, overhead condenser, reboiler, reflux receiver, and pumps are all part of the system to be investigated. For all this equipment and for the instrumentation involved, the engineer must be knowledgeable to determine the response or performance when process variables make a delta change. If chemical reactions are involved, this knowledge must include the kinetics of the reactants and products of the reaction.

Once the basis is firmly established, the next step is to develop a scenario of sequential events. In this regard, engineering judgment is required, equipment performance for each event estimated, and operator intervention at specific periods in the sequence of events anticipated. The real world is often complex but, when possible, the engineer uses judgement to make assumptions to simplify the system. The guidelines for any simplifying assumptions should:

- Incrementally increase the relieving rate
- Have minimal economic impact
- Consider the risk involved
- Use available information

A specific guideline to use for instrumentation response is to take no credit for instrumentation response if the instrumentation tends to reduce the relieving load. The other side of the coin assumes that instrumentation responds as designed if the instrumentation tends to increase the relieving load. These guidelines are the recommended approach of API.

API does allow credit in a model scenario for operator response. Again, judgment must be factored into the scenario based upon the instrumentation design to make the operator aware of events to which he must react. At this point, one must consider if the operator has time to correct the problem or problems. A power failure could trigger most alarms, but the operator must follow a predetermined sequence to minimize risk to the entire complex. A majority of catastrophic accidents involve human error to some degree. Therefore, in all cases needing human response, one must consider the possibility of failing to act, taking the wrong corrective action, or making the wrong diagnosis.

Examples

- **Fire Case Problem**

Determine the relieving load generated during a fire case for a vertical feed surge drum. The following data are given:

VESSEL DATA

design pressure	= 200 psig
diameter	= 5 ft ID
tangent length	= 15 ft
liquid level	= 6 ft
vessel elevation	= 18 ft
average mol weight	= 100 lb/lb-mol

Take no insulation credit. Use figure D-3 in API RP-520 to determine the temperature and latent heat at relieving conditions.

- **Debutanizer**

Determine the relieving load generated during a condenser failure. Use the heat and material balance and flow sketch in the handout as a basis.

DESIGNING THE RELIEF DEVICE

Type of Relief Device

There are many types of relief devices. The design principles, construction and maintenance needs vary for each type. The three main categories of relieving devices are spring-loaded, pilot-operated and rupture discs. The design expectations for these devices are:

- Open automatically and quickly during a rise in system pressure
- Close with minimum leakage when normal pressure is reached (except for rupture discs)
- It must be highly reliable

- **Spring-Loaded**

The spring-loaded type is utilized in a refinery for the majority of services. One must be aware of two types – the conventional and the bellows-type relief valves.

The conventional gas or vapor relief valve has the outlet or discharge pressure always on the downstream side of the disc sealing the outlet nozzle of the relief valve. Normally, the spring force is the differential between the set pressure and atmospheric pressure. The set pressure will be increased by superimposed back-pressure unless the spring force is adjusted accordingly. The impact of downstream pressure can, therefore, affect the performance of the relief valve.

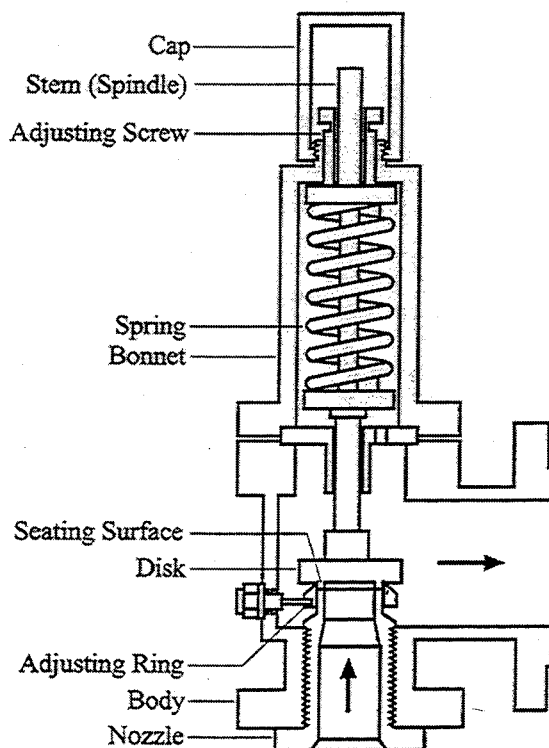
The bellows-type gas or vapor relief valve has a bellows, which isolates the outlet or discharge pressure from the major portions of the disk sealing the outlet nozzle. The spring and major portion of the outlet side of the disk are at atmospheric pressure. The resultant design does not allow any unbalanced forces on the disc due to downstream back-pressure

variations. The balanced pressure relief valve makes possible higher pressures in the relief discharge manifolds.

The opening characteristics of the spring loaded relief takes place in three distinct steps. First, the valve will "simmer" or pass minute quantities of gas that can be heard at about 95% of set pressure. Second, with the seating disc off the seat, the increased underside pressure causes an unbalanced force on the seating disc, rapidly opening the disc. Third, once the disc is up, the reactive forces of discharge vapors upward on the seating disc cause a secondary lift to provide maximum discharge area at 105% of lift. When closing, the valve throttles to about 50-70% of its capacity and then closes at 5-7% below its set point.

Figure XV-5

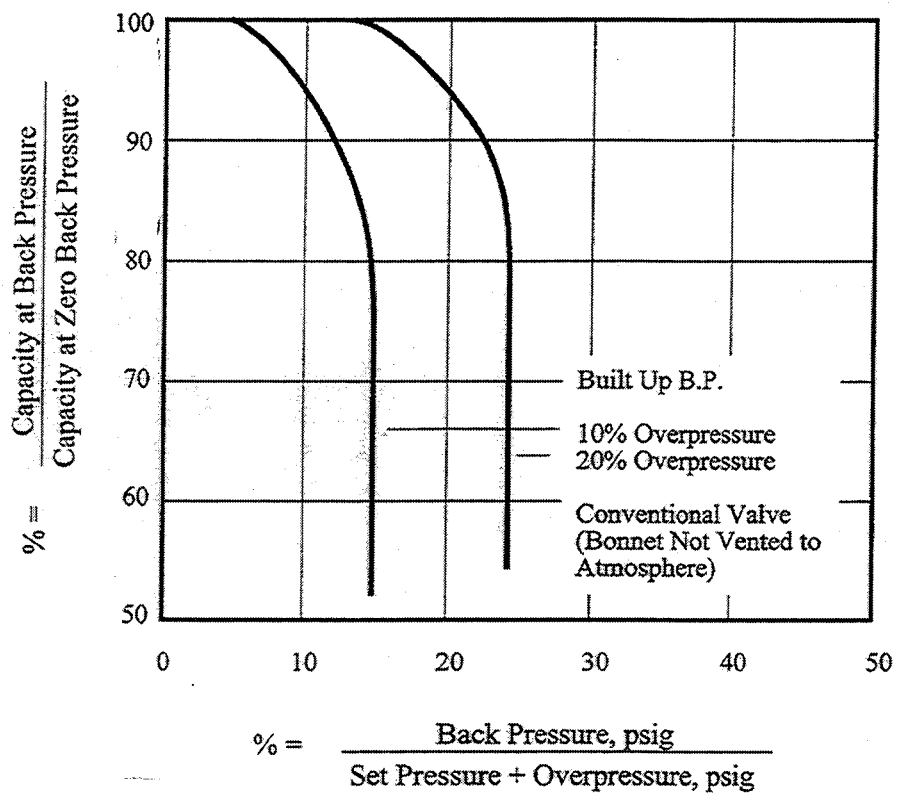
***Conventional Safety Relief Valve with a Single
Adjusting Ring for Blowdown Control***



IPD-R00-11

Figure XV-6

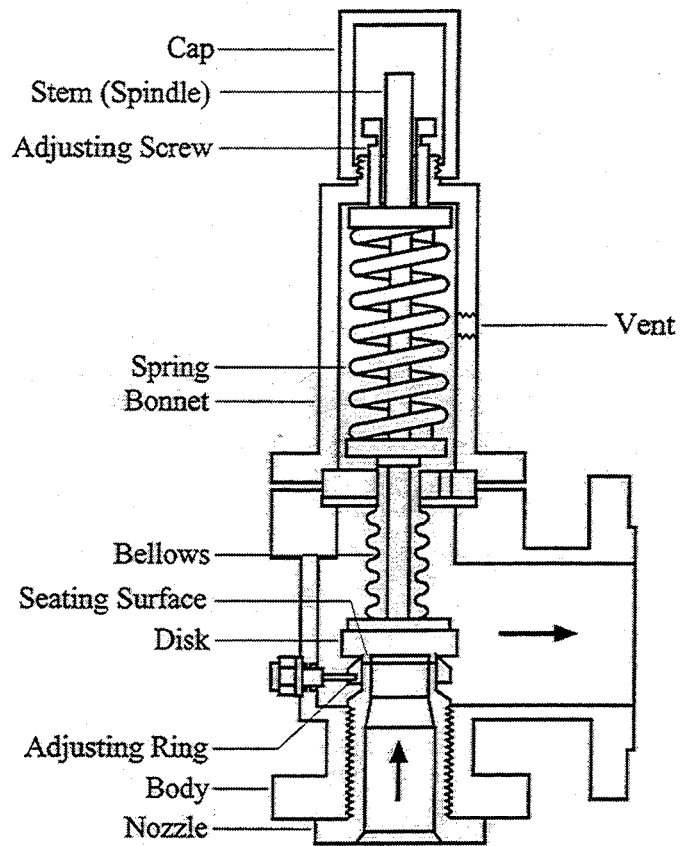
Back Pressure Response



IPD-R00-05

Figure XV-7

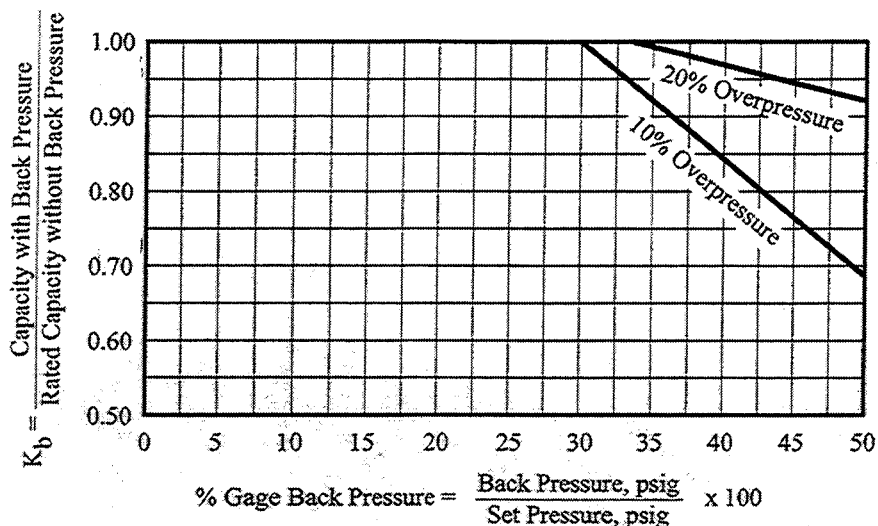
Balanced-Bellows Safety Relief Valve



IPD-R01-12

Figure XV-8

Back Pressure Response



IPD-R00-06

- **Pilot-Operated Relief Valves**

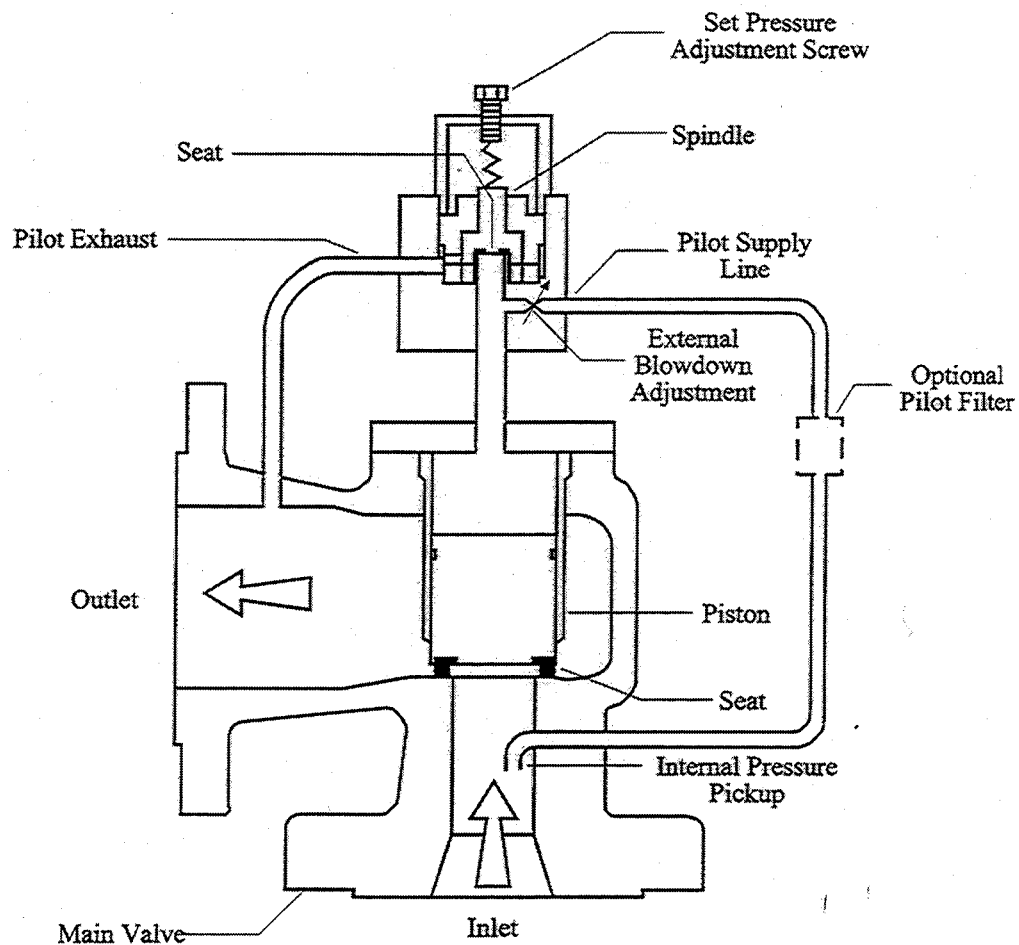
The pilot-operated relief valve has many advantages over the spring-operated relief valve. The pilot-operated relief valve consists of a main seating piston and a small external pilot. The piston is designed to have a larger effective area on the top than on the bottom. The system pressure is communicated to the top side of the piston through the pilot. As the system pressure increases, the seating force tends to make the valve tighter. When the set pressure is realized, the pilot opens discharging the gas above the piston and seals off the communication port that ties into the system pressure. The sealing piston immediately goes wide open. Simmering and leakage common with spring-type relief valves is not the case with pilot-operated relief valves. For a typical installation, the pilot-operated relief valve allows operating pressure to be at 95% of set pressure.

The vent of gas from the pilot can go to atmosphere if non-toxic or non-inflammable. In many cases, there is no good home for this vent gas, and manufacturers have a design to provide a non-flowing pilot when required.

The pilot valve can be designed to provide a pop-action or a modulating-action pilot. The pop-action open completely and the modulating-action pilot opens the main valve only enough to satisfy the required relieving capacity. The user should consult the manufacturer to determine the characteristics and mechanical construction of these special types.

Figure XV-9

Pop-Action Pilot-Operated Valve (Flowing Type)



IPD-R00-13

In summary, the advantages and disadvantages of pilot-operated relief valves are following:

ADVANTAGES	DISADVANTAGES
Seat tightness	More difficult to install
Remote pressure sensing	More delicate
Less sensitive to back pressure	Lower temperature constraint
Field testing of set point	Limited metallurgy selection
Remote depressuring	
Ability to modulate flow	
More flexible technology	

At UOP, about 10% of all relief valves are specified as pilot-operated types. In part, this is to satisfy the 3% inlet pressure loss constraint of API 520, Part II. In addition, the remote sensing allows delinking of the dynamic performance of the valve (chattering) from the inlet pressure loss.

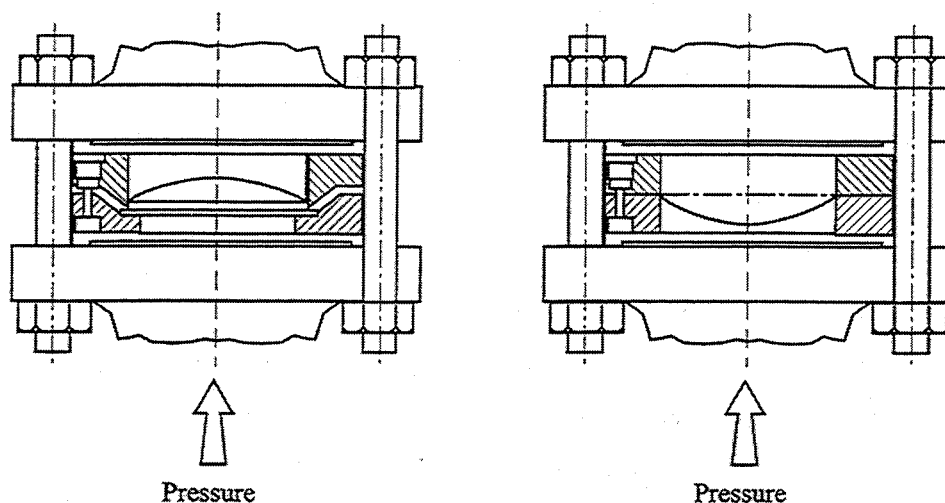
- **Rupture Disc Device**

The rupture disc is a pressure-containing element designed to rupture at a predetermined set pressure. Some designs are domed with the system pressure on the concave side, others are designed to open along prescored lines, and others are reverse buckling and, when overpressured on the convex side, the dome is forced into knife blades when open the element. The ratio of operating pressure to burst pressure varies from 80-90%, depending on the specific design.

The big disadvantage to a rupture disc device is that once open or burst, it will not close or reseal as other types of pressure relieving devices. In addition, it is back pressure sensitive, temperature sensitive, and subject to corrosion. One of the biggest uses today is for installation under a pressure relief valve to eliminate leakage or to isolate the pressure relief valve from a corrosive environment. Consult API recommended practice and the codes to determine the restrictions for this configuration.

Figure XV-10

Rupture Disc



IPD-R00-07

SIZING EQUATIONS

From first principles, the orifice area required for a relieving device is dependent upon desired capacity, characteristics of the fluid (compressible or non-compressible) at discharge temperature and pressure, pressure differential across the orifice area, and characteristic of the orifice. All sizing equations consider these elements to determine the required area to satisfy the relieving loads.

For vapor service, the equations allow the orifice to be sized at the accumulated pressure, which is a percentage of the set pressure. The code standards for various contingencies follow:

Single Valve	Non-Fire Case:	110%
Multiple Valves	Non-Fire Case:	116%
Single Valve	Fire Case:	121%

If the back-pressure caused by the sum of static and frictional pressure losses exceed the allowable limit for a given relieving device, a back-pressure correction must be applied to increase the orifice area. See API or manufacturer's literature for this correction factor. For liquid service, the equations have terms for both the set pressure and the destination pressure.

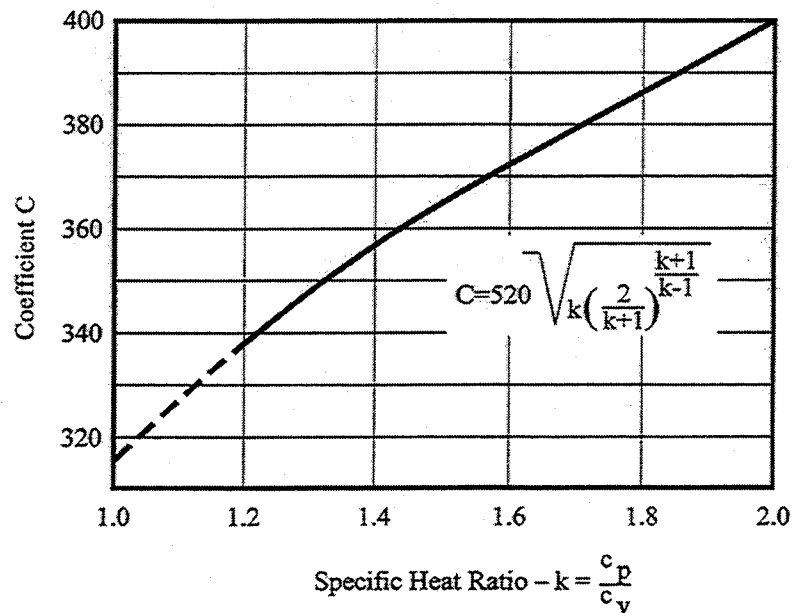
Vapor Flow Equation

$$RA = \frac{W}{CKP_{acc}K_b} \sqrt{\frac{TZ}{MW}}$$

where: RA = required orifice area, in²
 W = vapor flow rate, lb/h
 CK = flow coefficient
 P_{acc} = accumulated pressure, psia
 K_b = back pressure correction factor
 T = temperature, °R
 Z = compressibility factor
 MW = molecular weight

Figure XV-11

Flow Coefficient



IPD-R00-08

Liquid Sizing Equations

$$RA = \frac{gpm}{38K_d K_p K_v} \sqrt{\frac{SpGr}{1.25P_{set} - P_b}} \quad \text{or} \quad \frac{gpm}{38K_d K_v} \sqrt{\frac{SpGr}{P_{acc} - P_b}}$$

where: RA = required orifice area, in²

gpm = liquid flow rate, gpm

K_d = coefficient of discharge

K_p = overpressure correction factor

K_v = viscosity correction factor

$SpGr$ = specific gravity

P_{set} = set pressure, psig

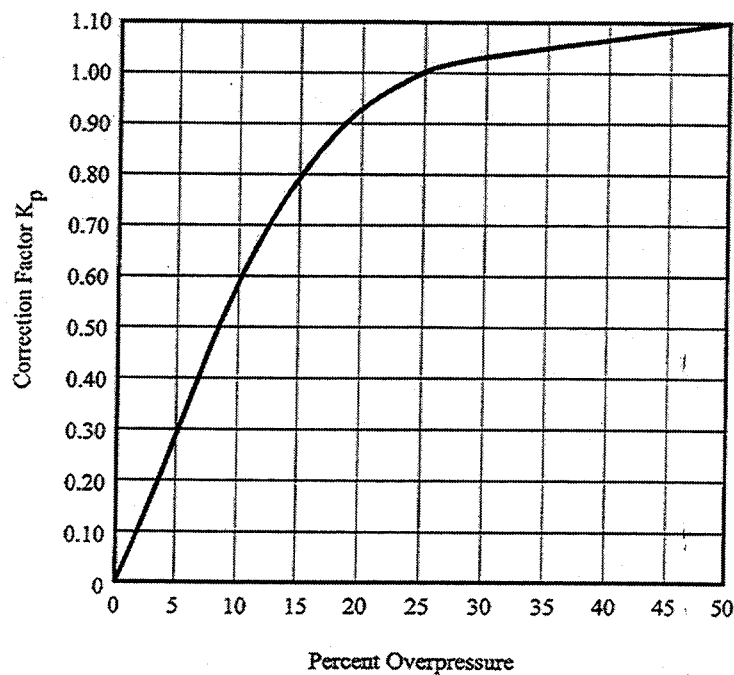
P_b = back pressure, psig

P_{acc} = accumulated pressure, psig

Note: Second equation is used when code certification is required.

Figure XV-12

Overpressure Correction



IPD-R00-09

DESIGN CONSIDERATIONS

There are a number of factors which must be considered in selecting the correct relieving device. Some of these factors are:

- Liquid or vapor service
- Balanced or non-balanced
- Temperature suitability
- Metallurgy
- Seat tightness

Once the orifice size is calculated, the proper combination of valves can be selected to fit the requirements. The manufacturer's catalogs has a series of tables for standard orifice areas and options. If multiple valves are required, similar size valves should be selected.

INSTALLATION

Once the relieving device has been sized, it is the responsibility of the mechanical designer to specify, configure, and orientate the piping going into and out of the relieving device. Again, the mechanical constraints and code regulations must be followed. Some of these factors are:

- Inlet line pressure loss
- Back pressure constraint
- Free draining tailpipe
- Reactive forces
- Piping forces from thermal expansion
- Liquid surge collection vessels
- Special handling for corrosive material
- Winterizing
- Staggering set points for multiple valves

RELIEF SYSTEM DESIGN

The selection of a disposal method is subject to many factors that may be specific to a particular location or an individual unit. The basic purpose of a relief system design is to safely transport the relieving fluids to a location where it can be discharged or recovered. In order to safely discharge refinery fluids to a safe location, the engineer must be aware of toxicity, flash point, explosive limits, and ignition temperature of the relieving fluids. Other important concerns are excessive noise level and air pollution limits.

Discharge Destination

- **Atmosphere**

In some situations (nontoxic and non-flammable streams), pressure relief vapor streams may be safely discharged directly to the atmosphere with recognition of appropriate design parameters. Atmospheric discharge reduces the significant problems associated with the analysis of system loads, proper sizing of piping, mechanical design criteria, and back pressure buildup from closed systems. If the air quality codes allow emergency discharge to the atmosphere, the remaining potential hazards can often be overcome.

- **Flare or Incinerator**

Flares are used in most refineries for disposal of continuous flows of excess gases and surges of gases in an emergency situation. The flare should be designed to prevent liquid discharge, operate smokelessly, without excessive noise, and have a positive ignition system. The flare system design, including safe location and safe height, is covered in API recommended practices. If toxic chemicals require disposal, the complete combustion to safe stack products in an incinerator is a common solution.

- **Process Location**

Discharge of material to the same or another process system at lower pressure may be a safe economical design alternative (e.g., pump discharge relief valves frequently discharge back to the suction of the same pump or expensive solvents discharge to alternative fractionators).

- **Vent Tank or Quench Tower**

For materials which require isolation from the normal flare system, such as liquids which economically justify recovery, corrosive or toxic fluid which must be isolated, highly viscous fluids which could solidify, or hot fluids which require cooling require special handling. Quench drums, neutralization facilities, scrubbers, or separate closed relief header with recovery sump are special designs found in a refinery.

Piping Design

Some considerations for piping to and from a relief device are following:

- Lengthen inlet piping to improve access, remove from extreme hot or cold service, dampen pressure fluctuations, locate outside insulation
- Provide free draining to avoid liquid seal
- Size inlet equal or greater than psv inlet (ASME code requirement)
- Inlet friction drop to be less than 3% of psv set (except for a remote sensing device)
- Thermal and reactive load at relieving conditions

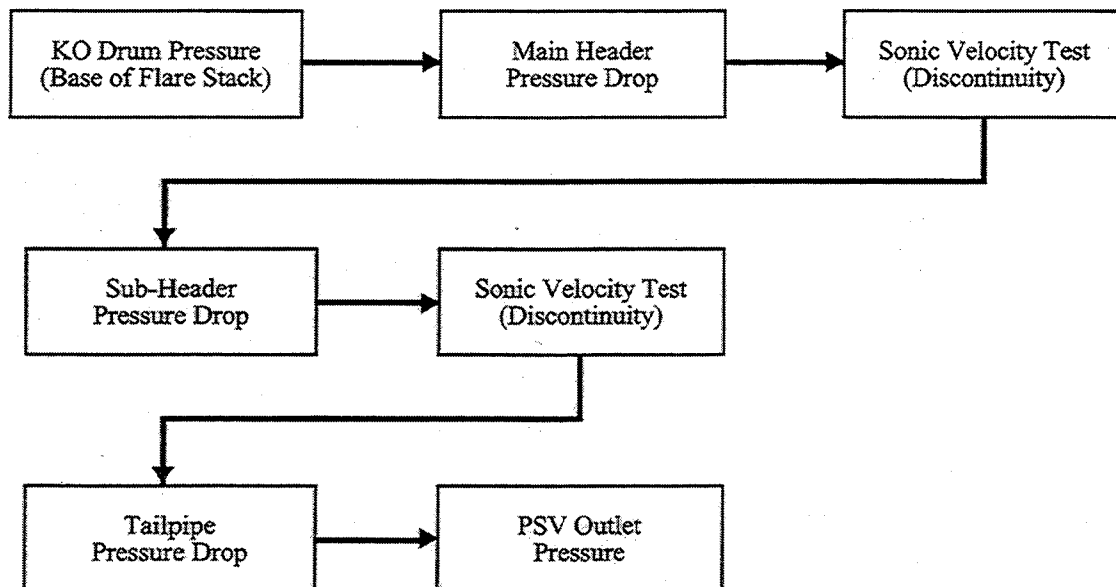
The basic criterion for sizing relief device discharge piping and manifolds is that the back-pressure not reduce the capacity of the relieving devices to protect the equipment from overpressure. Back-pressure cannot be calculated until the maximum flow rates are established. To this end, beside large individual relief rates, the simultaneous relieving rates must be determined for power failure, cooling water failure, instrument air failure, and fire case. When the maximum vapor relieving requirement has been determined and the maximum allowable header back pressure has been defined (as determined by type of valves and code requirements), the selection of line size is then reduced to a fluid flow calculation. The allocation of pressure drop is typically 70% for the main flare header and 30% for the sub-headers and tailpipe. Typical pressure drops are 15 psi for main header, 7 psi for sub-header and tailpipe, and 3 psi for flare stack.

The cost of the flare piping can be substantial and many times there is incentive to utilize existing facilities for new services. Therefore, means to reduce the maximum vapor relieving

loads are given special attention. UOP reduces the maximum calculated load by 25% to account for the improbability of simultaneous discharge at maximum rate, but there is no industry-wide standard. In some cases, shutdown devices may be used to reduce flare load.

Figure XV-13

Delta P Calculations



IPD-R00-10

Other considerations for relief device discharge piping and manifolds are critical flow in header system, which introduce pressure discontinuities, auto-refrigeration due to flashing of fluids, removal of liquid in knockout drums, and rapid acceleration due to pressure changes in system. To minimize reaction force, increased erosion, excessive noise, and pressure drop, the manifold line size can be increased until velocity of relieving fluids are subsonic.