| FOSTER WHEELER | PROCESS SIZING CRITERIA | N.I.O.C I.C.O.F.C |
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# PROCESS SIZING CRITERIA

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| PROJECT | : | SHANUL, VARAVI, AND HOMA GAS REFINERY |
| COMPANY | : | ICOFC |
| SITE | : | SOUTH EAST OF IRAN |

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## CONTENTS

## 1.0 SCOPE

This procedure outlines the basic sizing/ design criteria to be applied as a minimum in the equipment specification for the design of Shanul, Varavi and Homa Project which is located in south east Iran. In the event of a conflict between this specification and any of the Licensers sizing/design criteria, the Licensers shall prevail. Items not covered by this criteria shall be handled as an individual cases, with reference to a relevant industrial code or IPS.

## 2.0 PURPOSE

The purpose of this document is:

- at the BASIC stage to :
  - standardise and group options to be preferred to alternatives.
  - establish and correlate a source of basic information of typically multidiscipline interests for easy access and common reference.
  - agree and record minimum basic sizing criteria.
- at the Detailed Engineering stage to:
  - explain and clarify the options selected for equipment design during the BASIC stage
  - define the minimum requirements for completion of missing items by EP+C contractor(s).

## 3.0 DESIGN BASIS

- At the BASIC stage

  The process design criteria are detailed in the present Project Procedure.

- At detailed engineering stage

  These criteria shall be used to consolidate the equipment process specifications issued by the Licenser and to generate new ones. These criteria must be used only as guidelines. In particular, they will be applied to the pieces of equipment which are not fully specified by Licenser.

No change will be made to the data given by the Licenser without written agreement of the Licenser via ICOFC

The present Project Procedure shall be considered by the EP+C contractor(s) for a better understanding of the criteria used for equipment design during the BED and shall be used by EP+C contractor(s) to complete the design if required.

## 4.0 GENERAL

### 4.1 Applicability

The process design shall apply to new installations and to major modifications or extensions of existing installations. It is also applicable to supplier's packages.

### 4.2 Terminology And Definitions

PROCESS: Discipline(s) in charge to study the process units, but also all associated off sites and utility fluids.

PFD (PROCESS FLOW DIAGRAM): It represents the unit operations required to process the fluid to reach the required products specifications and defines the objective of the plant processing requirements including associated off-sites and utilities.

The controls are indicated by a simplified symbolic representation of control loops.

The equipment reference is given close to the symbolic representation.

In the upper part of the sheet, the equipment reference is given to indicate the function of each item of equipment.

The operating pressure, temperature, flow rate, and heat duty values are indicated in separated sheet by means of fluid number inside a diamond. Operating pressure and temperature in the appropriate square shape are indicated on PFD.

PROCESS DATA SHEET : data sheet for equipment, packages, etc, containing the process information required for sizing and given sometime the sketch with sizes for some equipment such as vessels, columns. It is issued by process.

It is different of the mechanical data sheet which defines additional information for construction and it is issued by specialists and/or vendors.

# 5.0 MECHANICAL DESIGN CONDITIONS

## 5.1 Design pressures

### 5.1.1 General

The design pressure is the value used for the mechanical sizing of equipment.

The design pressure of a piece of equipment (excluding storage tanks, atmospheric tanks and pipelines) shall be taken as the following:

| Maximum continuous operating pressure (bar g) "MOP | Minimum design pressure (bar g) |
|---|---|
| < 1 | 2 or 3.5 (*) minimum |
| 1 - 10 | MOP + 1 bar (**) |
| > 10 | MOP + 10% |

(*) 2.0 barg for PSV discharging to atmosphere, 3.5 barg for PSV discharging to flare networks.

(**) 2 or 3.5 barg as minimum design pressure

**(1)** Unless otherwise noted, the design pressure specified by Process applies to the vapour phase at the top of the vessel. The pressure drop over trays must be accounted for when selecting the design pressure for columns, reboilers etc.

**(2)** Minimum design pressure not applicable for thin wall equipment such as silos, storage tanks. In that case the governing parameter is full of liquid

**(3)** The design pressure shall also account for upset or transient conditions such as start-up, pressure surge, water hammer, settle-out pressure at compressor suction, etc..

**(4)** Vapour pressure at design temperature should be considered as design pressure except when safety relief valves are provided.

**(5)** Equipment subject to operate at pressure below atmospheric pressure will also be designed for full vacuum.

**(6)** Equipment that could face vacuum under abnormal conditions such as :

- vacuum conditions during start-up, shut down and/or regeneration purges.
- normally operated full of liquid but can be blocked in and cooled down
- containing condensable vapor but can be blocked in and cooled down
- could undergo a vacuum condition through the loss of heat input.

will be treated case by case. They will be designed for full vacuum unless fully reliable protective devices are provided (vacuum breaker, pressurisation gas, low pressure switch,....). Note that these devices should only be considered as a alternative if it would not be practical to design the equipment for full vacuum conditions i.e. low pressure storage tanks etc.

**(7)** For equipment in equilibrium with flare, the design pressure of the equipment is the flare design pressure.

**(8)** Hydraulic pressure due to the relative elevation between equipment and also the PSV's location shall be considered.

**(9)** Pressure vessels should normally have a design pressure of not less than 3.5 barg, especially when fitted with relief valves relieving to flare systems Lower design pressures can result in increased sizing of the flare system to limit acceptable backpressures. However, consideration may be given to increasing the design pressure if this can significantly reduce, or eliminate, relief loads. For equipment in equilibrium with Flare, the design pressure of the equipment is the flare system design pressure.

**(10)** In exceptional circumstances the margins between maximum operating pressure and design pressure for any system protected by relief valves can sometimes be reduced by the use of pilot operated relief valves.

**(11)** Conversely, the margins between the maximum operating pressure and the design pressure may be increased, e.g. so that the equipment can tolerate pressure caused by tube rupture or other potential sources of pressure ( see 5.1.5.1).

**(12)** The differential pressure used for tube sheets thickness calculations in feed-effluent heat exchangers (which have no means of isolation between the tube side and shell side streams) may be specified as the pressure difference between the two sides of the unit, rather than maximum system design pressure. This is based on the fact that high system differentials cannot occur over the tube sheet because of the absence of block valves.

**(13)** The design pressure of storage tanks is not usually specified by any general arbitrary rules, i.e. each is individually examined. The vapour pressure, tank venting, purging, and relieving systems described in API 2000 must all be considered before determining the maximum operating pressure of the tank. Tank design pressures may be selected from the recommendations in the appropriate design codes, e.g. those given in API 620 and 650, or BS 2654.

**(14)** Specialist equipment – it should be noted that some items of equipment, e.g. glass lined vessels, carbon block exchangers etc., may have design difference between the two sides of the unit, rather than maximum system design pressure.

**(15)** When the design pressure for an item which is remote from the source of pressure is being determined, it may be necessary to consider the influence of the pressure drop through the circuit, on the design pressures of all items in the circuit.

**(16)** Propane refrigeration systems should have a minimum design pressure of 18 barg, corresponding to propane vapour pressure at 55 °C.

**5.1.2** PSV setting

- The PSV set pressure should comply with the applicable codes of practise. In general for single relief valves this will be the design pressure of the system. The maximum operating pressure will be 90% of the set pressure. This margin between operating pressure and set pressure may be reduced by the use of pilot operated relief valves.

  It is reminded that the following tolerances are generally admitted for conventional instrumentation.

  - Set Pressure tolerance : +/- 3%
  - PSV recommended leak test : 10% below set point
  - In case of absolute necessity (for example in case of high pressure), the use of piloted PSV could help to reduce the design pressure.
  - If two or more PSV are in service, the set pressure will be staggered to avoid chattering. The difference between set points shall be less than 5% of the design pressure see API 520 / IPS-E-PR- 450 for specific details.
  - Refer to ASME Section VIII Appendix M for guidance on set pressures above 69 barg

**5.1.3** Pipelines

For pipelines, the design pressure is function of the Maximum Allowable Operating Pressure (MAOP) and the design factor, which depends on the class location.

Process determines only the MAOP.

MAOP is normally the design pressure of the last equipment upstream the pipeline plus the hydrostatic pressure due to the pipeline profile.

Particular attention shall be paid to the transient conditions such as equilibrium pressure plus hydrostatic pressure, water hammer, etc. ...

**5.1.4** Particular Cases

**5.1.4.1** Compressors

- At reciprocating compressor discharge :
  MOP + 2 bar for MOP ≤ 20 bar g,
  MOP + 10% for MOP > 20 bar g.
  PSV's are required.
- At the discharge of centrifugal compressor :
  MOP + 1 bar for MOP ≤ 10 bar g.
  MOP + 10% for MOP > 10 bar g.

  Generally surge pressure is above design pressure and PSV's are required.

- Consideration to be given to compressor arrangement to determine the settling pressure of the isolated system. The settling pressure is the equilibrium pressure reached between the suction and discharge isolating valves of the compressor system when the compressor is stopped or shut downed. Generally the design pressure of the equipment and piping at compressor suction should be above this settling pressure in order to avoid the use of unnecessary PSV's.

**5.1.4.2** Pumps

**a- Centrifugal pumps**

- Generally no PSV's are provided at the discharge of centrifugal pumps and the design pressure shall be the discharge pressure of the pumps at no flow with the maximum suction pressure and the maximum specific gravity.
- When the discharge pressure of the pumps at no flow is not available, this pressure can be estimated :

$$P_d = P_s \max. + \frac{1.25 \cdot head \cdot d_{\max}}{10.2}$$

| | | |
|---|---|---|
| $P_d$ | = | design pressure at pump discharge (bar g.) |
| $P_s$ max. | = | design pressure of suction drum + static head at $d_{max}$. and at HLA |
| head | = | head of the pump at design point (m) |
| $d_{max}$ | = | maximum specific gravity of liquid pumped under normal operating conditions |

- Note :

  (1) When pump curves are known, this design pressure has to be checked with actual head of the pump at no flow condition with the maximum suction pressure and the maximum specific gravity.

(2) Equipment elevation (from TL to grade shown on P+ID and data sheet) shall be considered as typical and shall be minimised when NPSH available / required are known.

**b- Volumetric pumps**

- At discharge of volumetric pumps :

  MOP + 1 bar for MOP ≤ 10 bar g.
  MOP + 10% for MOP > 10 bar g.
  PSV's are required.

- In case of two pumps in series, the maximum differential head will be the sum of the maximum differential head of each pump if there is no pressure relief valve between the pumps.

**5.1.5** Heat exchangers

**5.1.5.1** Heat exchangers design pressure

This concerns TEMA and multitubes heat exchangers.

According to the § 3.18.2 of API 521 : " loss of containment of the low pressure side to atmosphere, is unlikely to result from a tube rupture where the low pressure side ( including upstream and downstream systems) is designed for at least two-thirds of the design pressure of the high pressure side.". This is based the premise of a hydrostatic test pressure of 150% of the equipment design pressure. Hence pressure relief is not required. However, it should be noted that "Where the actual test pressure is less than 150% of the design pressure, this lower pressure should be used to determine whether over pressure protection is needed.". Present test pressure for equipment suppliers to the ASME VIII in 130% of design pressure hence the LP side of an exchanger should be designed with a design pressure of 10/13th of the HP side design pressure.

The recommended practice consists in oversetting, if necessary, the design pressure of low pressure side of heat exchanger:

- in all cases up to the limit of 150 #
- after analysis, case by case, for higher pressure.

This practice applies only to the heat exchanger itself and does not concern relevant piping and valving.

According to the § 3.18.6 of API 521, double pipe type of heat exchangers are not concerned.

As an alternative to the installation of relief valves the capacity of the shell side piping and downstream unit to accept tube rupture case shall be considered. Upstream and downstream piping and equipment must be thoroughly evaluated when this containment approach is taken.

For hot oil heat exchangers in the main process units, the design pressure on the utility side shall be increased to the design pressure of the process side in order to avoid PSV installation for the tube rupture case.

**5.1.6** Columns

- For columns, the same design pressure will be selected for the top of a fractionation tower and associated condenser, reflux drum and inter connecting piping.
- The design pressure at the bottom of a fractionation column (vapour phase) is determined by adding the column pressure drop at the column overhead design pressure.
- Liquid density and maximum liquid height in the bottom will be specified on the process data sheet to allow the vessel designer to calculate the bottom thickness.

**5.1.7** Tanks

Atmospheric tanks shall be designed full of water or full of product if product specific gravity >1 as a minimum. Depending on the type of tank, higher design pressures could be specified. To be treated case by case depending on tank type.

## 5.2 Design temperatures

**5.2.1** Equipment operating above 0 °C

- The design temperature is the value used for the mechanical sizing of equipment.
- The design temperatures shall be specified as follows; with due consideration for the special cases discussed at section 5.2.4 :
  Max. design temperature = max. operating temperature + 15°C or maximum exceptional operating temperature, whichever is the greater.
  Note: Exceptional operating temperature is to be considered for operations duration exceeding a total of 100 hours per year.
- Minimum upper design temperature should be 95°C due to solar radiation. This value should be examined case by case for equipment on which differential expansion can occur (such as doublewall tank, fixed tubesheet, plate heat exchanger) and for insulated high pressure vessels (not to increase wall thickness). For those equipment not influenced by solar radiation the maximum design temperature for all equipment shall be at least 55 °C. This is the maximum estimated temperature that can be achieved in insulated equipment after prolonged shutdown.

**5.2.2** Equipment operating at a temperature below 0 °C

- As a general rule the design temperature will be :
  $T_D = T_{SM} - 5$ °C
  Where:
  $T_D$ : Design temperature (°C).
  $T_{SM}$: Minimum continuous operating temperature (°C)

- Notes :

  1. When applicable, an exceptional "hot" design conditions set, e.g. temperature and related design pressure, may be added to the "cold" design conditions set for equipment operating at low temperature.

  2. When applicable, the exceptional temperature generated by depressurisation of equipment or interconnected items system will be indicated as well as the related residual pressure. Depending upon the depressurisation philosophy of the plant, dynamic simulations of equipment have to be performed using commercial or in-house software so as to determine the relevant pressure and temperature depressuring conditions (see Par. 9, Software).

  3. Temperature associated with a gas blow by from one equipment shall also be considered for the buried drum (belonging to the drainage system) design temperature.

**5.2.3** Discontinuous / Cycled Processes

- Various pressure and temperature conditions sets will be specified for each phase of equipment operation.
- Mixing of extreme conditions of pressure and temperature shall not be considered.
- Typical example is related to molecular sieves vessel design conditions specification. The sets of conditions of each phase of the operating cycle will be specified, e.g. design conditions of the adsorption phase, of the regeneration phase, etc.

**5.2.4** Special Cases

**5.2.4.1** Emergency depressurizing

- The minimum design temperature must take into account any depressurisation and repressurisation (depending of material selection) of the equipment / piping that may occur either during emergency or shutdown situation or gas blow-by from one equipment to another equipment and to the possible consequence of change of material.
-

- The emergency depressurizing shall impact the material selection as follows:

  - Piping material

    Piping material will be selected taking into account the temperature occurred during depressurisation. Piping repressurisation shall be considered as to be performed with the minimum depressurisation temperature.

  - Vessel material

    The minimum temperature due to the blowdown conditions shall be associated with design pressure. Although depressurization of any section of the plant cannot be performed unless the section is isolated and permissive is obtained, repressurization may take place by operator's fault or a valve failure, therefore the minimum temperature shall be associated with design pressure. No special devices are foreseen to ensure that the plant remains isolated and depressurised. In addition the above criteria will ensure safe operation should residual piping stresses be present (in particular for low diameter nozzle/piping).

**5.2.4.2** Heat exchanger and air cooler

- The following conditions mentioned hereafter will be applied generally up to the next process equipment.
- Consideration for design temperature definition should be given to cooling medium failure when coolers are used. Downstream of an air cooler, the design temperature is determined considering that 20% of the duty is provided by natural draft.
- For bypassed air cooler, the design temperature of the downstream equipment, if any, will be the maximum upstream operating temperature of the bypassed exchanger
- Downstream of other coolers, the design temperature will be the upstream maximum operating temperature.

**5.2.4.3** Cleaning and steam out

- Consideration to be given to conditions and fluids used for cleaning (e.g. steam). In this case, both pressure and temperature conditions have to be provided.
- The steam out conditions for vessels are 150°C / ATM. This information is considered by Vessel Department but will be also specified on SPP.
- Steam will be provided by temporary steam boiler(s).
- For equipment subjected to steam out operation full vacuum condition shall be specified @150 °C.

**5.2.4.4** Exceptional cases

- Consideration for upset and transient conditions such as start-up, shutdown, etc. In this case, both pressure and temperature conditions have to be provided.

- The exceptional temperature generated by fire will not be considered for design temperature selection.
- A specific design temperature will be given with the specified vacuum design pressure.
- The following considerations shall not prevail on § 5.2.4.2 regarding cooling medium failure.
  - The accidental temperature which may occur in emergency situations such as loss of utilities, valve closure, air cooler failure or any abnormal operation corresponding to a short duration are not to be taken into account as long as the temperature increase does not exceed the codes limits (Investigation has to be followed with specialists on a case by case basis).
  - However equipment containing parts which can be damaged by abnormal high temperature has to be designed for this temperature. This mainly concerns equipment internals.

## 5.3 Material selection

### 5.3.1 Basis of material selection

The Material Selection Report "RP-20013-171-6300-002" provides a full explanation of for the materials of construction selected for the project. The paragraphs below represent the minimum requirements.

- Material selection used for vessel

| Design temperature °C (1) | Steel type |
| --- | --- |
| -196 ≤ T < -101 | Austentitic SS |
| -101 ≤ T < -45 | 31/2 Ni /Austentitic SS |
| -45 ≤ T < 0 | LTCS |
| 0 ≤ T < 343 | CS |

(1) Design temperature corresponding to operating conditions. For temperature due to depressurisation, LTCS can be used down to a very low temperature (lower than -45°C) provided that vessel shall naturally reheat before repressurisation. This limit should be determined by Vessel Department, depending on Vessel wall thickness.

- Material selection used for piping

2

| Design temperature °C | Steel type |
| --- | --- |
| -196 ≤ T < -101 | Austentitic SS |
| -101 ≤ T < -45 | Austentitic SS |
| *-45 ≤ T < 0* | LTCS |
| *0 ≤ T < 343* | CS |

**5.3.2** Corrosion due to $H_2S$ and $CO_2$ services

**5.3.2.1** General considerations

$H_2S$ and $CO_2$ corrosion results in metal loss and in metal embrittlement (cracking). Regarding the stress cracking corrosion, this occurrence can be limited by using material listed in the NACE MR0175/ISO 15156.

Regarding the hydrogen induced cracking corrosion, its occurrence is directly linked to the material quality that is to say the possible presence of elongated inclusions, micro-segregations. Therefore, the HIC control is a matter of material specification.

**5.3.2.2** Sour service definition

Sour service definition is based on NACE MR0175/ISO 15156 criteria considered as a minimum requirement. The $H_2S$ service definition is to base the maximum pressure criteria, mentioned in the NACE, either on equipment design pressure when no PSHH protection is implemented on the concerned system, or on PSHH level when existing. Note that the gas associated with the Shanul, Varavi and Homa Gas Refinery as defined as 'sweet' hence sour conditions are not expected.

**5.3.2.3** Corrosion allowance

| | **Corrosion allowance on Carbon Steel and low alloy steel (mm)** | **Corrosion allowance on Stanless Steel or high alloy steel or cladded steel mm)** |
|---|---|---|
| Wet $H_2S$ service | 3 mm (1) | 0 |
| Corrosive process service (except wet $H_2S$ service) (2) | 3 min. | 0 |
| Non corrosive process service | 3 min. | 0 |
| Utility | 3 min. | 0 |

Notes

(1) Minimum corrosion allowance depending on corrosion control philosophy.

(2) Including amine systems, sulphur recovery units

(3) The corrosion allowance applies for pressure vessels, shell and tube type, air fin type heat exchangers.

(4) For piping, refer to piping classes.

(5) For storage tanks, corrosion allowance shall be:

- for fixed roof tank: 1.5mm for shell and roof except tank bottom where wall thickness is generally imposed by other constraints.

- for floating roof type tank: 0mm

Protective painting is applied on roof and bottom part of shell when water may be accumulated in the tank.

Licensers recommendations to be followed for concerned tanks.

## 5.4 Lethal service classification for pressure vessels/rotating equipment

The pressure vessel code requires to classify the equipment in accordance with lethal service.

In order to carry out this classification, the note "lethal service" will be specified on equipment data sheets handling a fluid with $H_2S$ content higher than 1000 ppm wt or where mercaptan level is exceeding 100 ppm in release gas.

For rotating machinery lethal service is noted on the process data sheet. The same criteria are used for pressure vessels. The purpose of the note is to assist seal selection and indicate items of equipment, which may require toxic gas detection.

It is noted that the gas from Shanul, Varavi and Homa is not expected to contain $H_2S$ or mercaptans.

# 6.0 EQUIPMENT DESIGN CRITERIA

## 6.1 Pumps

The minimum margin between the normal and rated flow for a pump will be as below :

| | | |
| --- | --- | --- |
| • reflux pump | = | +20% of normal flow |
| • other process pump | = | +10% of normal flow |
| • utility pump | = | +10% of normal flow |
| • loading pump | = | 0% of normal flow |
| • export pump from storage (continuous operation ) | = | +15% of normal flow |
| • Boiler feed water pump | = | See applicable codes but not less than 10% of normal flow |

To be noted that:

(1) When a non automatically controlled minimum flow protection has been installed, the permanent re-circulation flow (if required) must be added to the net process flow.

(2) Normal and rated flows will be identical in such instance as:
- intermittent service pumps : sump pump
- when the pump has been overrated to allow for a centrifugal type and if overrating is >10%
- re-circulation flow such as for product loading lines or through amine filtration system.

(3) Automatic start
- As base case, pump automatic start will be done generally through FSLL (if flow transmitter already exists).
- Automatic start is determined considering the following rules :
- personnel safety : for example flare KO drum pump will be started in order to avoid liquid in flare tips. In that case, considering the non continuous operation of flare drum pumps, the start of the spared pump can be performed by LSH or by DCS logic.
- equipment safety : for example BFW pump will be started in order to protect the steam drum and the steam coil.
- severe process upset : pump generating by its shutdown one process unit trip or generating an off spec product shall be spared automatically.
- flaring: automatic start shall NOT be considered to minimize the flaring. For example, reflux pumps unless a severe process upset is faced.

### 6.1.1 Pump Calculations

Pump calculations are simplified and standardised by using the Foster Wheeler "Pump Calculation Sheet" produced by the FW Baled Program for pump hydraulic calculation.

**6.1.2** Physical Properties

Physical characteristics of the fluid being pumped are based on the heat material balance and also established by reference to Foster Wheeler data, handbooks, information supplied by the customer, or other standard and reliable sources of data.

**6.1.3** Capacity

Design margins as set out in Section 6.1 are to be applied when setting pump design capacities.

For flows less than about $3.4m^3/h$, a larger pump is usually purchased and a fixed recycle is added to provide for recirculation of excess pumped fluid to the suction supply. An orifice plate is frequently included in the bypass line and this can be supplied by the pump vendor, who will size the orifice for a minimum flow requirement. The pump requisition sheet should specify a capacity equal to the desired flow (including extra capacity) plus the amount bypassed.

**6.1.4** Suction Calculation

This calculation yields the system pressure available at the pump centreline of horizontal pumps or at the centreline of the suction inlet nozzle for vertical shaft pumps. It involves the summation of the feed vessel's normal operating pressure and the static head loss the pressure drop in the suction piping resulting from friction, inlet-exit, and other losses.

The static head for vertical vessels is calculated from the bottom tangent line while for horizontal vessels, the bottom invert line is used. Usually, no credit is taken for the head contributed by liquid operating levels in a vessel. This should be reviewed on a case by case basis.

a) Suction line equivalent length (Le)

Equivalent length may be calculated using FW Baled Program in two ways for the suction lines, either the user inputs straight line length and fitting factor and the Le is calculated by multiplying the two, or the Le is automatically estimated from the pipe diameter d (inches) as follows:

| | |
|---|---|
| Pumping temperature < 150° C | Le=8d+30 m |
| Pumping temperature ≥ 150° C | Le=12d+30 m |

NB: the estimation excludes an allowance for suction strainer. This should be included as a additional loss see section 6.1.8. for DP.

b) Centreline elevations of horizontal pumps

The pump centreline elevation is automatically selected from the table below. If the flow exceeds 4540 $m^3/h$, the user is prompted to input an elevation.

| $m^3/hr$ | metres |
|---|---|
| Up to 45.4 | 0.76 |
| 45.4-227.1 | 0.91 |
| 227.1-2271 | 1.07 |
| 2271-4542 | 1.37 |

The suction pipe for fluids at or near their bubble point will be adequately sized if the pressure drop is in the range of 0.034 to 0.068 bar/100m. This is automatically checked by the 'baled programe' but must also be checked against the sizing criteria Section 7.3

**6.1.5** Net Positive Suction Head Available (NPSHA)

NPSHA is calculated by deducting the Vapour Pressure of the fluid at pumping conditions from the Suction Pressure and converting it to pressure head in terms of liquid column.

In reporting the vapour pressure, use the notation "Bubble Point Fluid" for boiling point liquids instead of numerical values.

Process engineers are to include a Safety Factor of 1.0 m in the NPSH calculated for:-

a) All boiling point fluids either single or multi-component.

b) Fluids that contain dissolved gas.

c) Foaming fluids.

In the case of boiler feedwater pumps, a margin of 2.0 m is to be used.

The static head used in calculating the NPSH shall be taken from either the tangent line in the suction vessel to one of the following:

- The centre line of a horizontal or rotary pump.
- The suction impeller on a vertical centrifugal pump.

The design of suction lines from storage tanks shall be based on the a NPSH taken from the lowest specified level in the tank at which rated pump capacity is required.

Suction line sizing for reciprocating pumps shall take into account acceleration head.

**6.1.6** Discharge Calculations

For pump discharge lines when fittings and valve count are not available, a reasonable estimate of the total equivalent length can be made by multiplying the approximate run of actual pipe by the multiplying factor. Details of applicable factors are given in section 7.3.6.

**6.1.7** Shutoff Pressure

The shutoff pressure of a typical centrifugal pump is approximately equal to the sum of the maximum suction pressure and 125% of the net differential pressure generated by the pump, based on the maximum anticipated fluid density. Other pumps with steep H-Q curves such as turbine, multistage and mixed flow pumps, however, will have higher shutoff pressures. The process engineer specifying these types of pumps shall consult with the Rotating Equipment Group to determine this value since it may influence the design pressure of downstream equipment.

The maximum discharge pressure sets the design pressure of a pump casing. This is the sum of the maximum suction pressure and maximum differential pressure, which usually occurs at zero flow. In cases where the feed vessel is protected by a safety relief valve, the maximum suction pressure will be equal to the sum of the safety valve set pressure and the maximum static head.

**6.1.8** Equipment Pressure Drops

The following typical pressure drops may be used in line size calculation when the actual pressure drop data are not available:

TYPICAL EQUIPMENT PRESSURE DROPS

| | Bar |
|---|---|
| Coalescer | 0.7 |
| Dessicant Drier | 1.0 |
| Desalter | 1.7-2.7 |
| Exchangers: | |
| S&T, Double-pipe & Air Coolers | 0.7 |
| Box Coolers | 3.5 |
| Fixed Bed Reactors | 1.4-3.5 |
| Flow Orifice | 0.14 |
| Orifice Mixer | 0.35/plate |
| Pump Suction Strainer | 0.07 |
| Rotary & Turbine Flow Meters | 0.4 |

For systems involving multiple heat exchangers in series, consult with the Heat Exchanger Group for pressure drop estimation.

**6.1.9** Control Valve Pressure Drop

The control valve normal pressure drop is calculated in three ways,

33% of frictional pressure drop or,

10% of operating pressure or

the value corresponding to a control valve pressure drop of 0.7 bar at maximum flow.

The maximum of these three values is automatically input in the calculation for the net design discharge pressure.

For systems operating above 69 barg, the control valve may take less than 10% of the operating pressure, depending on process and control considerations.

**6.1.10** Pump Cooling Water Requirement

Cooling water, preferably fresh, is used to cool bearings, stuffing boxes, pedestals and glands to safe temperature levels.

The coolant rate varies with temperature and to some extent with pump size. For design purposes, the following rates may be used:

| Casing Design Temp °C | m3/h |
|---|---|
| 120 | 0.25-0.75 |
| 120-250 | 0.75-1.5 |
| 250+ | 1.5-2.5 |

**6.1.11** Pump Efficiency

The efficiency of centrifugal pumps varies from about 20% for low capacity pumps (less than 6.3 $m^3/h$) to a high of almost 90% for certain large capacity pumps. Low head pumps using open type impellers are less efficient than closed impellers. The efficiency factor used in calculating power requirements of centrifugal pumps may be estimated from Fig. 1

**6.1.12** Motor-driven Pumps

For estimation of power consumption and motor rating of various size pumps, refer to tables in Figures 2a and 2b.

## 6.2 Compressors

- 10% margin should be taken on capacity.
- The variations of gas compositions, molecular weight, Cp/Cv,etc., and the operating conditions(mainly suction and discharge temperature and pressure, and, flowrate ) shall be taken into account to determine the sizing case.

### 6.2.1 Compressor Process Specifications

#### 6.2.1.1 Operating case

If more than one case exists, all these alternative cases shall be included in the specification so that the compressor vendor is able to evaluate the most stringent case for design.

#### 6.2.1.2 Capacity

The volume flowrate capacity shall be determined by the compressor manufacturer from the data sheet provided. The process engineer shall determine the mass flowrate based on minimum, normal and maximum flow conditions. Design margins are as set out in Section 6.2.

#### 6.2.1.3 Suction Temperature

Suction temperature is to be accurately specified since it is directly related to the volume of gas at suction conditions, the discharge temperature, and the horsepower requirements. It is important for the vendor to know the minimum and maximum temperatures for proper compressor design and selection of correct driver rating.

#### 6.2.1.4 Suction pressure

Suction pressure is the pressure at the suction flange of the compressor and not before filters, pulsation dampers, etc. The suction pressure must be accurately specified.

#### 6.2.1.5 Molecular weight

Molecular weight is an important consideration in the design of a centrifugal compressor. When this or any type of compressor is to be used in multiple services, the vendor is to be supplied with data on the molecular weight of the gases.

#### 6.2.1.6 Specific Heat Ratio

The specific heat ratio is also an important consideration in the design of centrifugal and reciprocating compressor as it affects both power and efficiency of the machines. It should be clearly documented what the basis for the stated Cp/Cv ratio e.g. ideal or polytropic etc.

**6.2.1.7** Compressor power estimation

Compressor power estimates shall include gear losses. When a compressor is to be used in vacuum or refrigeration service, peak driver load may be required during start-up and a footnote to this effect is to be added to the specification form. The final determination of compressor power requirements and discharge gas temperatures is part of the vendor's responsibility.

**6.2.1.8** Gas composition

This is to be supplied by the process engineer and is to be expressed on a wet basis if the gas contains moisture.

**6.2.1.9** Discharge Temperature (maximum allowable)

This is to be supplied by the process engineer when a known process limitation exists. Discharge temperatures are limited by gas reactions, e.g. polymerisation or in the case of air compressors with the lube oil, safe lubrication temperatures. Some compressors are limited by mechanical considerations and these will be defined by the Mechanical Equipment Group and the compressor vendor.

**6.2.1.10** Corrosive compounds

Corrosive compounds in the gas (such as sulphur oxides, hydrogen sulphides, acidic compounds, chlorides, etc.), are to be specified by the process engineer as these may determine the selection of materials by Foster Wheeler or the compressor manufacturer.

**6.2.1.11** Start-up considerations

Start-up methods are to be considered by the process engineer since items such as anti-surge control systems, bypass lines, valve lifters and pockets on reciprocators, etc., are involved. In addition, compressors generally require a running-in period during which time an alternative feed gas may be used. If air is to be used for running-in, then suitable vents, etc. may be an additional requirement.

**6.2.1.12** Compressor Selection and Comparison

Centrifugal compressors are the preferred type for the majority of applications.

Reciprocating compressors are to be considered for conditions of low flow, high differential pressures, intermittent loads, varying gas densities, and varying discharge pressures, combined with moderate temperatures.

Screw compressors may be employed for applications involving relatively low flows and differential pressures. Their selection should be referred to the rotating equipment specialists.

**6.2.1.13** Safety Considerations

The following potential hazards are to be considered for compressor installations.

a) At high pressures, many reactions proceed at higher rates, e.g. the reaction between a hydrocarbon lube oil and oxygen or air. The discharge temperature of air from reciprocating compressors is generally limited to about 149°-166°C. Compressor circuits frequently have automatic shutdown instrumentation, which operates on high gas discharge temperature.

b) Excessive discharge pressures from positive displacement machines can be attained if a discharge valve is inadvertently closed. Therefore, safety valves are mandatory for this class of compressors.

c) Adequate ventilation of the compressor house should be provided when compressing toxic or flammable gases. This is frequently accomplished by omitting the siding from a portion of the compressor house.

d) Adequate inlet K.O drums should be provided where necessary to prevent liquid slugs from damaging compressors. Providing demisters in the K.O drum can reduce entrainment.

e) Rotating compressors and their drivers have speed limitations. Trip-outs are indicated and these are usually supplied by the vendor and specified by the Mechanical Equipment Section.

**6.2.1.14** Bearing and Seal Losses

The polytropic horsepower absorbed by the gas compression phase does not include additional power, which is required for bearing and seal losses.

The combined losses may be estimated from the table below and are to be added to the polytropic power requirement.

| **Polytropic Power, kW** | **Power Loss, kW** |
|---|---|
| Up to 4500 | 19 |
| Above 4500 | 38 |

**6.2.1.15** Gear Losses

The mechanical efficiency of gears used to transmit power from a driver to a compressor varies as follows:

| Type of Gear | Mech. Efficiency % | Gear Loss |
|---|---|---|
| Single Reduction | 98-98.5 | 2-1.5% |
| Double Reduction | 97-97.5 | 3-2.5% |
| Triple Reduction | 96-96.5 | 4-3.5% |

## 6.3 Heater and boiler

The design margins to be applied are as follows:

- Fired heaters and furnaces : 10% on duty.
- Boilers : 10% on flow rate.

## 6.4 Vessels

**6.4.1** Over design factor

- 1st separation equipment (plant inlet): 10% on inlet gas flow rate.
- Other drums : 0% unless specific requirements
- Fractionation column : 0% unless specific requirements

**6.4.2** Vapour Area sizing

- The following excludes the flare/vent drums, desalters and electrostatic dehydrators.
- If internals are installed, the common vapour internal shall be a wire mesh but for some services a vane pack can be used after discussion with the Company
- The use of others vapour internals such as cyclones, etc. requires Company approval.
- The basis of sizing is the critical velocity Vc (m/s)

$$V_C = 0.048\left(\frac{\rho l - \rho g}{\rho g}\right)^{0.5}$$

with $\rho l$ = liquid density in $kg/m^3$
$\rho g$ = vapour density in $kg/m^3$

The maximum gas velocity is K*Vc.

K is a coefficient depending on the service, and the use or the absence of wire mesh.

- Recommended K values are given hereafter for different services.

| Service | Without wire mesh | With wiremesh |
|---|---|---|
| Production separator | 1.7 | 2.2 |
| | | |
| Fuel gas drum | 0.8 | 1.7 |
| | | |
| Compressor suction drum | 0.8 | 1.7 |
| | | |
| Glycol or amine contactor inlet drum | 0.8 | 1.7 |
| Reflux drum | 1.7 | 2.2 |
| | | |
| Steam drum | N/A | 1.3. |

- If a vane pack internal is used, the recommended K value is 3.3. This has to be confirmed with the supplier.
- For horizontal vessels without vapour internal ( wire mesh, vane pack, etc. ), the minimum distance between the top of the vessel and the LSHH is the largest of 300 mm or 0.2 internal diameter.
- Vessels handling paraffin oil shall not be equipped with gas internals.

A high efficiency inlet distributor can be considered to improve gas / liquid separation provided that EPC contractor verify pressure drop through distributor and dimensions between inlet distributor/mesh and inlet distributor/HLA

**6.4.3** Hold up residence times of liquids

- If the vessel is sized to receive a slug, that slug volume shall be taken between NLL and HLA.
- The residence time corresponds to half of the hold-up time, the Normal Liquid Level (NLL) being set at 50% of the HLL-LLL range. Exceptions will be specified on data sheet.

- The minimum liquid hold up time between LLA and HLA are as follows :

| SERVICES | TIME ( MINUTES ) |
| --- | --- |
| Feed Surge Drum | |
| A. to heater | 5 |
| B. to others | 3 without pump |
| | 5 with pump |
| Reflux Drum | 5 |
| Fractionation tower bottom : the largest of | |
| A. product to next process | 5 |
| B. product to other column | 5 |
| C. product to storage tank | 3 without pump |
| | 5 with pump |
| Steam flash drum (process units) | 5 |
| Steam drum (utility generation) | 10 |
| Desalter | 15 |
| Deaerator (note1) | 15 |
| Atmospheric degassing drum | 15 |
| Others Drums | 3 without pump |
| | 5 with pump |

Note 1: Liquid hold up time is based on one deaerator in shutdown associated with the normal liquid flowrate

- Two phase vessels

| ISA SYMBOL | DATA SHEET SYMBOL | VERTICAL DRUM | HORIZONTAL DRUM |
|---|---|---|---|
| LAHH/LSHH | HHLA/HHLS (HLL) | | |
| | | at least 1 to 2 min. with 150 mm min.<br>to verify : min. 10% of control range<br><br>IF only HLL : HLA-HLL : 10% of control range | at least 1 to 2 min. with 100 mm min.<br>to verify : min. 10% of control range<br><br>IF only HLL : HLA-HLL : 10% of control range |
| LAH | HLA | | |
| | | liquid hold up time to be considered with 300 mm min. | liquid hold up time to be considered with 300 mm min. |
| LAL | LLA | | |
| | | at least 1 to 2 min. with 200 mm min.<br>to verify : min. 10% of control range<br><br>IF only LLL : LLA-LLL : 10% of control range | at least 1 to 2 min. with 100 mm min.<br>to verify : min. 10% of control range<br><br>IF only LLL : LLA-LLL : 10% of control range |
| LALL/LSLL | LLLA/LLLS (LLL) | | |
| | | 300 mm min., but to be compatible with time required to close a SDV | 150 mm min., but to becompatible with time required to close a SDV |
| | Tangent line (1) | | |

- three phase separator

| ISA SYMBOL | DATA SHEET SYMBOL | VERTICAL DRUM | HORIZONTAL DRUM |
|---|---|---|---|
| LAHH/LSHH | HHLA/HHLS (HLL) | | |
| | | at least 1 to 2 min. with 150 mm min.<br>to verify : min. 10% ofcontrol range<br>IF only HLL : HLA-HLL : 10% of control range | at least 1 to 2 min. with 100 mm min.<br>to verify : min. 10% of control range<br>IF only HLL : HLA-HLL : 10% of control range |
| LAH | HLA | | |
| | | lightest density liquid hold up time to be considered with200 mm min. | lightest density liquid hold up time to be considered with 200 mm min. |
| LAL | LLA | | |
| | | at least 1 to 2 min. with 200 mm min.<br>to verify : min. 10% of control range<br>IF only LLL : LLA-LLL : 10% of control rang | at least 1 to 2 min. with 100 mm min.<br>to verify : min. 10% of control range<br>IF only LLL : LLA-LLL : 10% of control range |
| LALL/LSLL | LLLA/LLLS (LLL) | 450 mm min. | 450 mm min. |
| LDAHH | HHIA (HIL) | | |
| | | at least 1 to 2 min. with 150 mm min.<br>to verify : min. 10% of control range<br>IF only HIL : HIA-HIL : 10% of control range | at least 1 to 2 min. with 100 mm min.<br>to verify : min. 10% of control range<br>IF only HIL : HIA-HIL : 10% of control range |
| LDAH | HIA | | |
| | | highest density liquid hold up time to be considered with 200 mm min. | highest density liquid hold up time to be considered with 200 mm min. |

| ISA SYMBOL | DATA SHEET SYMBOL | VERTICAL DRUM | HORIZONTAL DRUM |
|---|---|---|---|
| LDAL | LIA | | |
| | | at least 1 to 2 min. with 200 mm min.<br>to verify : min. 10% of control range<br>IF only LIL : LIA-LIL : 10% of control range | at least 1 to 2 min. with 100 mm min.<br>to verify : min. 10% of control range<br>IF only LIL : LIA-LIL : 10% of control range |
| LDALL | LLIA (LLIS) | | |
| | | 300 mm min., but to be compatible with time required to close a SDV | 150 mm min., but to be compatible with time required to close a SDV |
| | Tangent line (1) | | |

To be noted that:

Exception is made for vertical vessels with negligible liquid on clean service with manual or on/off liquid outlet valve ; in that case volume of the hemi-spherical head can be used : LLL (or LLLA/LLLS) location to be still compatible with SDV or control valve closing time.

**(1)** The distance between HLL and LLL will be in the typical range of 1 to 3.5 meters.

**(2)** When applicable, the hold up time below the very low liquid level ( LLLA or ILLLA) has to be compatible with the time required to close a SDV.

**(3)** For three-phase separators, the retention time for the two liquid phases shall be considered.

**(4)** The effective retention volume in a vessel is that portion of the vessel in which the two liquid phases remain in contact with one another. As far as the two liquid phases separation is concerned, once either substance leaves the primary liquid section, although it may remain in the vessel in a separate compartment, it cannot be considered as a part of the retention volume.

The highest density liquid retention volume is taken between the bottom and the normal interface level ( INLL).

**(5)** The lightest density liquid retention volume is taken between the INLL and the normal liquid level (NLL).

**(6)** Stand pipe shall be installed on clean service when at least 3 level instruments have to installed (independently from level instrument required for safety actions) e.g. : one level transmitter with two level gauges

**(7)** Minimum size for stand pipe: 3”

**(8)** Particular case : slug catcher: stand pipe shall be installed

Gauge glasses and level controller will cover the full range of level transmitters and alarm switches.

**(9)** Connections for level instruments generating a trip function shall be independent from control function.

**6.4.4** Diameter

- As a general rule, inside diameter will be specified on process data sheets ( in mm )
- If the required diameter for a vessel is lower than 800 mm, a note will be added specifying that a piping element is acceptable.
- For vessels less than 1000 mm ID, flanged heads may be specified.
- Recommended L/D ratio for horizontal vessel :

| PRESSURE barg | L/D |
|---|---|
| Lower than 17 barg | 3 |
| 17 barg up to 34 barg | 4 |
| Higher than 34 barg | 5 |

**6.4.5** Manholes

- size of manholes

- for vessel diameter < 1000 mm
  - Flanged vessel will be considered if equipment contains internals
  - otherwise, size of manhole = 18"

- for vessel diameter ≥ 1000 mm
  - toxic service size of manhole = 24"
  - non toxic service size of manhole = 20"

- location of manholes
  at the opposite side of the utility connection for horizontal vessel

- number of manholes

Vessel

For vessel length/height less than 6 m a single manhole will be provided. For other vessel (length/height > 6m), two manholes to be provided at least ; one manhole each 6 m for longer/higher vessel. If vessel is equipped with internals (baffle, etc.), one manhole to be provided on each compartment.

Except when provided with full bolted full diameter end closure all vessels 800 mm O.D. and larger shall have at least one manway for inspection maintenance requirements.

Trayed column

Manhole will be provided at the top, below the bottom tray, at the feed tray, at any other tray at which removable internals are located, and at intermediate points so that the maximum spacing of manholes does not exceed 15 trays. Tray spacing with manhole in the internal will be at least 900 mm.

**6.4.6** Handhole

Handhole size = 8". Handhole to be installed on vessel with diameter lower than 800 mm or on vessel where severe fouling of internals is expected.

**6.4.7** Vortex breaker

Vortex breaker to be installed for the following services:

- pump suction
- thermosiphon or kettle inlet.
- letdown to a low pressure capacity

Vortex breaker on fouling service to be at 150mm from vessel wall.

**6.4.8** Drains

- Location

  The drain of the vessel shall be connected

  - to the outlet line at low point for vertical vessel
  - and directly on the capacity for horizontal vessel or for vertical vessel with outlet line entering inside vessel.

- Vent and drain diameter shall be defined as follows:

| **Volume or diameter of vessel ($m^3$ or mm)** | **Vent diameter** | **Drain diameter** |
|---|---|---|
| $V \leq 15$ OR $D \leq 2500$ | 2" | 2" |
| $15 < V \leq 75$ or $2500 < D \leq 4500$ | 2" | 3" |
| $75 < V \leq 220$ or $4500 < D \leq 6000$ | 3" | 4" |
| $220 < V \leq 420$ or $D > 6000$ | 4" | 4" |
| $V > 420$ | 6" | 4" |

- Drain number :

  For horizontal drums having a length greater than 6 m TL to TL, additional drain connections are required . Additional drain is also required on each compartment of the vessel.

  On toxic service, a open drain connection (washing out) to be provided with blind flange (not connected to outside).Size of open drain will be in the same diameter as drain connection.

**6.4.9** Utilities connections (steam out, purging)

- for vertical vessel

  The steam out connection will be of 2". For large vessel, two 2" connections will be provided for diameters greater or equal to 4.5 m.

  If vessel is equipped with internals (baffle), one steam out connection will be provided on each compartment.

- for horizontal vessel - non toxic

  Advantage shall be taken to use connection on drain to steam out the vessel. Specific steam out connection is not required.

- for horizontal vessel - toxic

  The steam out connection will be of 2". For large vessel, two 2" connections will be provided for length greater or equal to 6 m.

  If vessel is equipped with internals (baffle), one steam out connection will be provided on each compartment.

  No hard piping connection for steam out will be provided.

**6.4.10** Elevation of equipment

As a conventional rule for a vessel containing a product at its boiling temperature, a minimum elevation of 5000 mm will be specified when a recovery bottom centrifugal pump is provided. The elevation will be updated when NPSH requirements are defined with rotating equipment specialist. If there is no process requirement regarding the elevation, a note on P+ID will be indicated "minimum for piping".

## 6.5 Heat exchangers and air coolers

### 6.5.1 Oversizing

- Shell and tubes heat exchangers: 10 % flow and duty
- Air coolers : 17 % flow and duty

### 6.5.2 Fouling factors

The following gives some fouling factors for process and utility fluids which can be reviewed case by case.

- Process fluids m².°C/W

| | |
|---|---|
| Acid gas | 0.00020 |
| Sour natural gas from slug catcher | 0.00035 |
| Sour natural gas downstream unit 100 | 0.00017 |
| Sweet gas | 0.00017 |
| Liquid LPG | 0.00020 |
| Raw Feed Condensate from slug catcher | 0.00052 |
| Raw Feed Condensate | 0.00035 |
| Stabilised Condensate | 0.00020 |
| Process water | 0.00035 |
| Stripped water | 0.00030 |
| Glycol | 0.00040 |
| Refrigerant (propane) | 0.00015 |

- Licensed units fluids by Licenser

- Utilities

| | |
|---|---|
| Sea water | 0.00050 |
| Chilled water | 0.00020 |
| Potable water | 0.00020 |
| Saturated steam / LP condensate | 0.00017 |
| BFW/Demineralized water | 0.00017 |
| Nitrogen | 0.00017 |
| Instrument air | 0.00017 |
| Fuel gas | 0.00017 |
| Diesel | light: 0.00030 |
| heavy: 0.00035 | |

### 6.5.3 Fouling factors for plate exchangers

For Plate Frame Heat Exchangers, a general fouling factor of 0.00005 m²°C/W shall be taken for all fluids (or Process Licenser recommendation).

For Plate Fin heat Exchangers, no fouling factor shall be applied but an extra surface of 15% to be added on calculated area.

**6.5.4 Temperature approach for heat exchangers**

The temperature approach shall be optimised for heat exchangers but it shall not be smaller than:

- 5°C for TEMA type heat exchangers (shell and tube)
- 10°C for air coolers
- 3°C for plate type heat exchangers.
- 3°C for kettle type

**6.5.5** Specific requirements for heat exchangers

TEMA R will be generally used for all shell and tubes and air fin type heat exchangers. Fixed tube sheet exchangers are acceptable for non fouling service on the shell side. In this case, all exceptional operating conditions ( start-up, shutdown, etc. ) to check the necessity to provide an expansion bellows on the shell shall be defined.

**6.5.6** Air cooler type

Air cooler to be induced type when air cooler is installed on pipe rack. Forced draft type shall be considered

- if air cooler is installed at grade
- if inlet process temperature is above 175°C or air outlet over 93°C or for multiple purpose exchangers, several sections stacked.

**6.5.7** Allowable Pressure Drop – Shell and Tubes

Typical allowable pressure drops are given below:

**6.5.7.1** LIQUIDS

| | Total Pressure Drop (bar) -Shells in series- | | |
|---|---|---|---|
| Viscosity, Cp. (at average temperature) | One | Two | Three |
| Less than 1.0 | 0.35 to 0.7 | 0.35 to 0.7 | 0.7 to 1.0 |
| 1.0 –5.0 | 0.7 | 1.0 | 1.0 to 1.4 |
| 5 – 10 | 1.0 | 1.0 to 1.4 | 1.4 |
| Above 10 | 1.4 | 1.4 to 2.0 | 2.0 |

Notes:

1. Under the following circumstances, $\Delta P$'s approaching the higher recommended values should be employed: when the $\Delta T$ is small (28°C or less) or when the temperature range is large, (above 111°C).
2. Calculated tube side pressure drop values are subject to greater variation than shell side values, because of the nature of tube bundle construction.
3. It must be realised that little can be gained by specifying increased pressure drop for one fluid in an exchanger when the other fluid has a significantly lower film coefficient.
4. For gravity flow, the pressure drop is usually limited from 0.07 to 0.14 bar.

**6.5.7.2** GASES

| Operating Pressure (barg) | Pressure Drop (bar) |
|---|---|
| 0 – 0.7 | 0.035 – 0.07 |
| Above 0.7 | 0.14 – 0.35 |

**6.5.7.3** CONDENSERS

| Types | Pressure Drop (bar) |
|---|---|
| Partial | 0.14 – 0.35 |
| Total | Negligible |

**6.5.7.4** REBOILERS

| Types | Pressure Drop (bar) |
|---|---|
| Kettle | Negligible (shell side) |
| Thermosyphon | |
| Horizontal | 0.02 – 0.035 |
| Vertical | Equivalent to approx. 3 – 5m tube length |

**6.5.8** Allowable pressure drop – Air Fin Exchangers

Suggested pressure drop for various services are given below. However, care should be taken to ensure that the selected pressure drop results in the most economic overall installation.

The allowable pressure drop for product cooling and non-critical services should not control the size of the exchanger, as this may result in an uneconomic design which could be avoided by reconsidering the hydraulics of the process circuit. Special consideration is required for wide temperature range cooling of viscous liquids, low pressure gases or condensation of vapours at very low pressures. In these services, pressure drop is a critical requirement, which greatly influences the size of the heat transfer surface.

ALLOWABLE PRESSURE DROP

| Service | Allowable pressure drop, (bar) | |
|---|---|---|
| Liquid Cooling | 0.7 | Note 1 |
| Gas Cooling: | | |
| Operating pressure 1.0 to 3.5 barg | 0.2 | |
| Operating pressure 3.5 to 17.5 barg | 0.35 to 0.7 | |
| Condensing: (atmospheric pressure and above) | | Note 2 |
| Total condensation | 0.035 min | Note 3 |
| Partial condensation | 0.14 to 0.35 | Note 3 |

Notes:

1. Not valid for viscous fluids.

2. For vacuum service the selection of an allowable pressure drop should be from the results of an economic study. Pressure drops are usually in the range of 3-5 mm Hg.

3. For multi-pass air coolers high pressure drops assure proper flow distribution. The higher pressure drop will also assure proper distribution at lower than design throughput.

# 7.0 PIPING

## 7.1 General design and hydraulics

7.1.1 Line velocity and friction loss for liquid line and gas line.

Line size of each line shall be firstly selected based on the mass flow rate and in accordance with the velocity range criteria and then be checked in accordance with the friction loss range criteria as given in paragraph 7.3 Line sizing criteria

7.1.2 Minimum piping sizes

Except for instrument piping, connections to equipment or piping in which minimum flow velocity requirements govern, the minimum size shall be:

- ¾" for pipe when located above ground
- 2" for process line on pipe rack
- 2" for utility line on main pipe rack
- 2" for pipe on pipe sleeper
- 2" for underground steel pipe
- 2" for underground non metallic piping

7.1.3 Pump suction line

Pump suction lines shall be sized to provide available net positive suction head (NPSH) 1 meter larger than that required by the pump selected.

Pump suction lines shall not be smaller than suction nozzle. Therefore, reducer at pump suction is acceptable provided that the available calculated NPSH (with pressure loss in the reducer) remains acceptable regarding the required NPSH.

Pump suction valves shall be in the same diameter as the line.

As a first estimate, the static head to be used in calculating the available NPSH shall be taken from, for vertical vessels, the bottom tangent line while for horizontal vessels, the bottom invert line, to one of the following:

- The centerline of a horizontal pump or rotary pump
- The suction impeller on a vertical centrifugal pump.

Usually, no credit is taken for the head contributed by liquid operating levels in a vessel. This should be reviewed on a case by case basis.

For the value of NPSH specified on process data sheet, the reference elevation shall be indicated.

The design of suction lines from storage tanks shall be based on a NPSH taken from the lowest specified liquid level in the tank at which rated pump capacity is required.
In sizing suction lines for reciprocating pumps, acceleration head shall be considered.

**7.1.4** Control valve

- Control valve pressure drops are defined in section 6.1.9

- In case of mis-operation, the gas blow-by shall consider the flow rate through the control valve full open and its by-pass also fully open when it is installed. If that flow rate sizes the flare, the manual by-pass could be deleted or a mechanical interlock between the upstream control valve manual block valve and upstream by-pass valve manual block valve could be installed.
- For control valve arrangement, refer to P+ID development DB 20013-999-P312-203.

## 7.2 Insulation and tracing

Thermal insulation for hot or cold services may be required for :

- heat or cold conservation of equipment and piping,
- personnel protection of equipment for operating temperatures above 70°C. A physical barrier with warning signs attached to hot surface is preferred to thermal insulation if it is not required for process reasons,
- to avoid external water condensation or ice
- steam or electrical heat tracing

In all cases, insulation shall be minimised in order to limit CUI (Corrosion Under Insulation).

## 7.3 Line Sizing Criteria

This paragraph shall not be applied to the flare lines. The pressure drop and velocity guidelines provided may be used for the preliminary sizing of lines. However, the final sizing should also take into account other factors, such as pump NPSH requirements, pressure drops available, and specific client or process requirements. Where specific maximum velocity limits are given these should not be exceeded.

### 7.3.1 Line sizing criteria for gases and steam

| Vapour and steam lines | $\rho v^2$ max. (kg.m.s $^{-2}$) | Max. Velocity (m/s) | DP (bar/km) | |
|---|---|---|---|---|
| | | | Normal | Maxi |
| - Continuous operation | | | | |
| P <= 20 bar g | 6000 | | ) | |
| 20 < P <= 50 bar g | 7500 | | ) | |
| 50 < P < =80 bar g | 10000 | | ) Pressure drop must be | |
| 80 < P <= 120 bar g | 15000 | | ) considered compatible | |
| P > 120 bar g | 20000 | | ) with corresponding service | |
| - Compressor suction | Compatible with above | | 0.2 | 0.7 |
| - Compressor discharge | Compatible with above | | 0.45 | 1.15 |
| - Discontinuous operation | | | | |
| P <= 50 bar g | 10000 | | ) | |
| 50 < P <= 80 bar g | 15000 | | ) Pressure drop must be | |
| P > 80 bar g | 25000 | | ) considered compatible | |
| - Column overhead | 15000 (high pressure columns) | | ) with corresponding service | |
| - Stripper vapor return | | | 0.2 | 0.45 |
| - Kettle vapor return | | | 0.2 | 0.4 |
| Steam lines | | | | |
| - P <= 10 bar g | | | | |
| Short line (L <= 200 m) | 15000 | | 0.5 | 1.0 |
| Long line (L > 200 m) | 15000 | | 0.15 | 0.25 |
| - 10 < P <= 30 bar g | | | | |
| Short line (L <= 200 m) | 15000 | 42 | 1.2 | 2.3 |
| Long line (L > 200 m) | 15000 | 42 | 0.25 | 1.0 |
| - P > 30 bar g | | | | |
| Short line (L <= 200 m) | 15000 | 30 | 1.2 | 2.3 |
| Long line (L > 200 m) | 15000 | 30 | 0.35 | 1.0 |
| Vacuum (<0.2 bara) | | | 0.001 | 0.002 |

7.3.2 Line Sizing Criteria For Liquids

| Liquid line type | ΔP (bar/km) | | Max. Velocity. (m/s) (2) | | | |
|---|---|---|---|---|---|---|
| | Norm. | Max. | To 2" | 3" to 6" | 8" to 18" | from 20" |
| <u>Pump suction</u> | | | | | | |
| - Liquid at bubble point with dissolved gas | 0.6 | 0.9 | 0.6 | 0.9 | 1.2 | 1.5 |
| - Non boiling liquid | 2.3 | 3.5 | 0.9 | 1.2 | 1.5 | 1.8 |
| <u>Unit lines</u> | | | | | | |
| - Liquid at bubble point with dissolved gas | 0.6 | 1.0 | 0.6 | 1.0 | 1.4 | 1.8 |
| - Non boiling liquid | 2.3 | 3.5 | 0.9 | 1.2 | 1.8 | 2.4 |
| <u>Pump discharge (1)</u> | | | | | | |
| - Disch. pres. <= 50 bar g | 3.5 | 4.5 | 1.5 to 4.5 m/s | | | 6.0 |
| - Disch. pres. > 50 bar g | 7.0 | 9.0 | 1.5 to 4.5 m/s | | | 6.0 |
| **COLUMN OUTLET** | 0.6 | 0.9 | 0.6 | 0.9 | 0.9 | 0.9 |
| Gravity flow | 0.25 | 0.45 | 0.6 | 0.6 | 0.6 | 0.6 |
| Water lines (CS)(3) | | | | | | |
| - Cooling water & service water (4) | | | | | | |
| Large feeders between pumps and units | 1.5 | | 1.5 to 3.0 m/s | | | |
| Unit lines (long) | | 1.5 | 1.5 | 2.5 | 3.0 | 3.0 |
| Unit lines (short) | | 3.5 | 1.5 | 2.5 | 3.0 | 3.0 |
| - Boiler feed | | | | | | |
| Pres. <= 50 bar g | 3.5 | 4.5 | 1.5 to 4.5 m/s | | | 6.0 |
| Pres. > 50 bar g | 7.0 | 9.0 | 1.5 to 4.5 m/s | | | 6.0 |
| -Sea water lines | | | 2.5 to 3.5 m/s (2 m/s mini) | | | |
| - Steam cond. return | | | 1 to 1.5 m/s | | | |
| - Reboiler feed (for indication) | 0.2 | 0.4 | | | | |

Notes:

**(1)** 3.0 m/s maximum ( 2 m/s average ) at storage tank inlet or in loading.

**(2)** Vendor and/or Licenser requirements could supersede maximum velocity values upon Company approval.

**(3)** Special considerations can be applied for copper-nickel or glass reinforced plastic piping upon Company approval.

(4) Velocities below 1 m/s should not be used for cooling water service to avoid solids deposition.
(5) For amine service velocity should not exceed 1 m/s to avoid corrosion/erosion.
(6) For lines containing mixtures of hydrocarbon and water, velocity should be limited to 1 m/s to avoid generation of static.
(7) 60 to 98% sulphuric acid lines velocity should not exceed 1.2 m/s to avoid cororsion.

**7.3.3** Line Sizing Criteria For Two Phase Flow

For preliminary mixed phase fluid line size calculations, the average density method will be used while considering the following criteria:

- $V_m$ : 10 to 23 m/s
- $\rho_m V^2$ : 5000 to 10 000 Pa
- $\rho_m V^3$ : 100 000 to 150 000 kg/s$^3$

where:

- $\rho_m = W / ((W_l/\rho_l) + (W_v/\rho_v))$ in kg/m$^3$

$W = W_l + W_v$ = total flow rate in kg/h
$W_l$ = liquid flow rate in kg/h
$W_v$ = vapour flow rate in kg/h

$\rho_l$ = liquid density in kg/m$^3$
$\rho_v$ = vapour density in kg/m$^3$

and the apparent fluid velocity $V_m$ expressed as :

- $V_m = W / (3600\rho m\ \pi D^2/4)$ in m/s

D = internal diameter of the line in m.

As a general, continuous flow patterns should be ensured such as :

- Stratified, annular, bubble, wavy flow patterns, etc. for horizontal lines or slightly sloped.
- Annular or bubble flow, etc. for the vertical lines
- For horizontal lines in slug and plug flow regimes and for vertical line in slug flow regimes reinforced anchoring shall be specified.

**7.3.4** Line Sizing Criteria For Offsite Line

The following criteria are typical and may have to be supported by economical appraisal.

| LINE TYPE | ΔP (bar / km) | | Max. velocity (m/s) |
|---|---|---|---|
| | Normal | Maximum | |
| Long Carbon steel water line | 0.58 | 1.16 | - |
| Bonna line | - | - | 2.5 to 3 |
| Steam condensate (mixture) | 0.2 to 0.3 | - | 10 to 20. |

### 7.3.5 Corrosion / erosion criteria

Corrosion

For corrosion resistant material ( SS, Special alloys, etc. ), no limitation of flowing velocity up to 100 m/s and no requirement for corrosion allowance.

For non corrosion resistant material, in corrosive fluid service, corrosion allowance for a design service life and corrosion inhibitor injection are required. The flowing velocity is limited by the inhibitor film integrity.

Erosion

For Duplex, SS or alloy material, the flowing velocity must be limited to:

- 100 m/s in single phase vapour lines and multiphase lines in stratified flow regimes ( 65 m/s for 13% Cr material ),
- 20 m/s in single phase liquid lines and multiphase lines in annular, bubble or hydrodynamic slug flow regime,
- 70 m/s in multiphase lines in mist flow regimes.

For Carbon Steel material:

- In case of continuous injection of corrosion inhibitor, the inhibitor film ensures a lubricating effect, which increases the erosion velocity limit. The corrosion inhibitor erosion velocity limit will be calculated taking into account the inhibitor film wall shear stress.
- The API RP 14 E recommendation should apply: the flowing velocity must be maintained below the erosional limit :

$$Ve = C / (\rho_m)^{0.5}$$

With: Ve erosional velocity in ft/s

$\rho_m$ gas / liquid mixture density at flowing conditions in $lb/ft^3$

C empirical constant equal to 100 to 125. For solids free where corrosion is not is not anticipated or when corrosion is controlled by inhibition or by employing corrosion resistant alloys the value of C =150 - 200 can be considered with the peak flow rate.

When solid and/or corrosive contaminants are present C value shall not be higher than 100.

**7.3.6** Equivalent Line Length Calculation

The total equivalent length (Le) can be calculated using a factor multiplied by the straight length of pipe or by adding up the equivalent length of pipe fittings and the straight length of pipe. This method should be used when pipe routing has not been finalised / defined.

For the Fitting Factor calculation, input the straight line length and the calculation will be performed automatically from the default fitting factors (ensure all the number of fittings = zero), within the FW Baled Programs. The Fitting Factor values may be overwritten if required.

**TABLE OF FITTING FACTORS**

Default values used by Liquid and Vapour Line Sizing Programs.

| Line sizes, diameter | Approximate line length , ft | | |
|---|---|---|---|
| | 100 | 200 | 500 |
| 3in or less | 1.9 | 1.6 | 1.2 |
| 4in | 2.2 | 1.8 | 1.3 |
| 6in | 2.7 | 2.1 | 1.4 |
| 8in or over | 3.4 | 2.4 | 1.6 |

| Fitting types | Fitting sizes, in | | | L/D ratio |
|---|---|---|---|---|
| 90 deg bend | 3.0 | 4.6 | 6.1 | 30 |
| 45 deg bend | 1.6 | 2.4 | 3.3 | 16 |
| Gate valve | 0.8 | 1.2 | 1.6 | 8 |
| Thru Tee | 2.0 | 3.0 | 4.1 | 20 |
| Branch Tee | 6.1 | 9.1 | 12.2 | 60 |

## 7.4 Absolute Roughness

The values for absolute roughness used by the FW Baled Program for liquid and vapour line sizing calculation is as follows:

| ABSOLUTE ROUGHNESS, mm | |
|---|---|
| Commercial steel | 0.046 (default) |

## 8.0 FLARE AND COLD VENT SYSTEMS

### 8.1 Type of Flare Tip

For flares and cold vents, the tip can be normally conventional or sonic depending on the required back pressure and noise limitation.

When possible a sonic tip will be preferred. Sonic tip with Coanda effect and/or with variables slots are prohibited.

The flares shall be generally smokeless.

The analysis of the causes of relief is required and an occurrence flaring loads balance including each individual relieving rate for each possible cause shall be performed.

### 8.2 Radiation Levels Criteria

The radiation levels criteria shall follow the Basic Engineering Design Data. The minimum relative humidity stated on the basis of design shall be applied.

### 8.3 Emissivity Coefficient

When the radiation calculations are performed by a flare vendor it is necessary to check carefully the emissivity coefficient used. The emissivity coefficient used by vendors does not take into account the liquid carry over, they consider an ideal gas/liquid separation. The droplets size for the flare drum sizing and the expected liquid carry over shall be clearly indicated in the flare tip process data sheet.

**RECOMMENDED EMISSIVITY COEFFICIENT**

For pipe flare :

- natural gas molecular weight of 18 : 0.21
- natural gas molecular weight of 21 : 0.23
- ethane : 0.25
- propane : 0.30
- see also API RP 521.

For sonic flare:

The emissivity coefficient = 0.13 for all gasses without liquid carry over, and 0.15 with liquid carry over not exceeding 5% weight.

## 8.4 Flare and cold vent lines sizing criteria

### 8.4.1 Lines upstream relieving devices

PSV's:

For the line sizing, the maximum capacity of the PSV, even if this figure exceeds the actual maximum flow rate due to process limitations shall be considered.

- DP between the protected equipment and the PSV $\leq$ 3% of PSV set pressure ( API RP 520 Part II )
- $\varnothing \geq \varnothing$ PSV inlet
- $\rho V^2 \leq$ 25 000 kg/m/s² for $\varnothing \leq$ 2"
- $\rho V^2 \leq$ 30 000 kg/m/s² for P $\leq$ 50 bar g.
- $\rho V^2 \leq$ 50 000 kg/m/s² for P > 50 bar g.

DEPRESSURISATION DEVICE

- minimum line size 2"
- $\rho V^2$ criteria are the same as for PSV's

### 8.4.2 Lines downstream relieving devices

flare and cold vent headers and sub-headers:

- minimum line size 2".
- back pressure to be compatible with the protected equipment.
- velocity and $\rho V^2$ :
  - SINGLE PHASE (GAS AT THE RELIEF DEVICE INLET):
    - intermittent flow :
      - lines downstream relieving devices and sub-headers : 0.7 Mach maximum and $\rho V^2$ < 150 000 kg/m/s² considering the maximum capacity of the relieving devices even if this figure exceeds the actual maximum flow rate due to process limitation and the relevant occurrence.
      - headers: 0.7 Mach maximum and $\rho V^2$ < 150 000 kg/m/s² considering the maximum flow rate due to process limitations and for the relevant occurrence, however a velocity of 0.8 Mach could be accepted for a long straight line without elbows and connections ( e.g. stack, line on bridge ).
      - For a $\rho V^2$ > 100 000 kg/m/s² vibration and line support studies are required.
    - continuous flow :
      - velocity < 0.35 Mach and $\rho V^2 \leq$ 50 000 kg/m/s².

- MULTIPHASE ( 2 phase flow at the inlet of relieving device ) :
  - velocity ≤ 0.25 Mach and $\rho_m V_m^2 \leq 50\,000$ kg/m/s².
    For $\rho_m$ and $v_m$ definitions see § 7.3.3

The sizing shall be done for the line downstream each device with the built-up back pressure for the corresponding occurrence and not with the maximum built-up pressure for the maximum flow rate to the flare or cold vent. The same shall be applied for the headers and sub-headers.

## 8.5 Flare Drum Sizing

For flare drum and cold vent drum, the sizing shall follow API RP 521 method with the following droplets size in microns and a relief to flare time of 20 minutes with no liquid disposal:

- vertical flare or cold vent onshore : 600 μm

## 8.6 Purge Gas

The purge gas is provided to avoid:

- an explosive mixture in the stack or header for air intaking into the flare or cold vent stack.
- the risk of burn back which induces a quickest deterioration of the flare tip.

The purge gas flowrate shall not be lower than the value given by the following equation:

- without gas seal : Purge gas flow = $24000 \text{x } D^3 \text{x} MW^{-0.565}$
- with gas seal : Purge gas flow = $12000 \text{x } D^3 \text{x} MW^{-0.565}$

Where: Purge gas flow in $Sm^3/h$
D is the tip internal diameter (1) in m
MW is the purge gas molecular weight in kg/kmole

(1) for sonic flare, the tip internal diameter is taken as the equivalent diameter corresponding to the exit gas area.

For flare if fuel gas is used for purge gas, the source of purge gas shall be common to the fuel source to the pilots in order to avoid a loss of purging while pilots remain in service.

The heaviest available gas should be preferably used as the normal source of purge gas in order to minimise the vacuum pressure in the flare header for an elevated flare or cold vent.

In some cases, nitrogen could be used as purge gas.

## 9.0 COMPUTER PROGRAMS

The computer programs approved and authorised for use on this Project are defined in the Process Coordination Procedure.

## 10.0 APPENDIX:

**Fig. 1 Pump Efficiencies vs GPM**

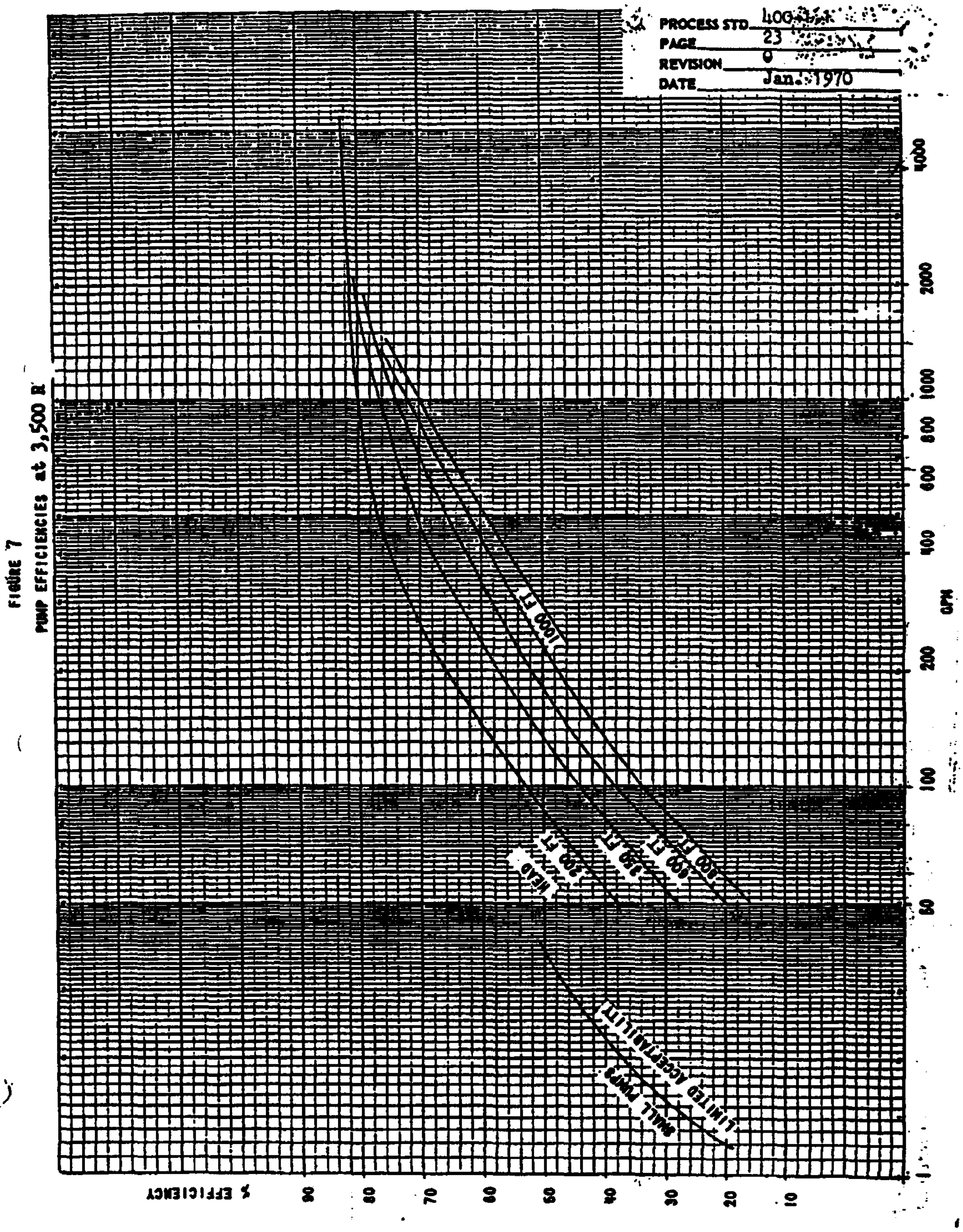

Fig. 2a Motor Efficiency for Pumps (up to 250 HP)

**TABLE 1A**

**HIGH EFFICIENCY ELECTRIC MOTORS (1)**
**250 AND LESS HORSEPOWER**

| Motor Rating HP | Motor Efficiency @ % of Full Load Capacity 50 | 75 | 100 | Conn Load KW |
|---|---|---|---|---|
| 1 | 78.5 | 81.5 | 81.5 | 0.9 |
| 1.5 | 80.0 | 82.5 | 84.0 | 1.3 |
| 2 | 81.5 | 84.0 | 84.0 | 1.8 |
| 3 | 86.0 | 88.0 | 88.0 | 2.5 |
| 5 | 87.5 | 88.5 | 88.0 | 4.2 |
| 7.5 | 91.0 | 90.6 | 90.6 | 6.2 |
| 10 | 91.2 | 91.2 | 91.0 | 8.2 |
| 15 | 91.4 | 91.3 | 91.4 | 12.2 |
| 20 | 92.8 | 92.5 | 92.3 | 16.2 |
| 25 | 93.0 | 93.0 | 92.3 | 20.2 |
| 30 | 93.2 | 93.2 | 92.8 | 24.1 |
| 40 | 93.6 | 93.7 | 93.5 | 31.9 |
| 50 | 93.3 | 93.9 | 93.5 | 39.9 |
| 60 | 93.7 | 94.0 | 93.7 | 47.8 |
| 75 | 94.0 | 94.4 | 94.4 | 59.3 |
| 100 | 94.0 | 94.4 | 94.4 | 79.0 |
| 125 | 94.0 | 94.5 | 94.2 | 99.0 |
| 150 | 94.5 | 94.8 | 94.8 | 118.0 |
| 200 | 95.1 | 95.2 | 95.1 | 156.9 |
| 250 | 95.2 | 95.4 | 95.4 | 195.5 |

Notes: (1) Based upon G.E. low voltage (460V) high efficiency motors. This type of motor shall be used only on client's request.

Fig. 2b Motor Efficiency for Pumps (over 250 HP)

TABLE 1B

ELECTRIC MOTORS, RECOMMENDED SIZE & EFFICIENCY
OVER 250 HORSEPOWER

| Motor Rating HP | Motor Efficiency (1) @ % of Full Load Capacity 50 | 75 | 100 | Conn Load KW |
|---|---|---|---|---|
| 300 | 90.6 | 91.7 | 91.8 | 243.8 |
| 350 | 90.9 | 91.9 | 92.1 | 283.5 |
| 400 | 91.3 | 92.3 | 92.4 | 322.9 |
| 450 | 91.5 | 92.5 | 92.6 | 362.5 |
| 500 | 91.7 | 92.6 | 92.8 | 401.9 |
| 600 | 92.2 | 93.0 | 93.2 | 480.3 |
| 700 | 92.4 | 93.2 | 93.4 | 559.1 |
| 800 | 92.6 | 93.2 | 93.6 | 637.6 |
| 900 | 92.7 | 93.5 | 93.8 | 715.8 |
| 1000 | 92.7 | 93.5 | 93.7 | 796.2 |
| 1250 | 93.0 | 93.9 | 94.1 | 991.0 |
| 1500 | 93.2 | 93.9 | 93.9 | 1191.7 |
| 1750 | 93.3 | 94.1 | 94.4 | 1382.9 |
| 2000 | 93.5 | 94.2 | 94.5 | 1578.8 |
| 2250 | 93.6 | 94.4 | 94.6 | 1774.3 |
| 2500 | 93.7 | 94.5 | 94.7 | 1969.4 |
| 3000 | 93.8 | 94.5 | 94.8 | 2360.8 |
| 3500 | 94.0 | 94.6 | 94.8 | 2754.2 |
| 4000 | 94.0 | 94.7 | 95.0 | 3141.1 |
| 4500 | 94.1 | 94.8 | 95.0 | 3533.7 |
| 5000 | 94.2 | 94.8 | 95.0 | 3926.3 |

Notes: (1) Based upon G.E. high voltage (4160V) standard efficiency motors.
(2) Refer to Page 25 for motor HP selection.