

GAS SWEETENING PLANT DESIGN

By:

ADITYA SINGH

(1404551002)

ROHIT GOSWAMI

(1404551040)

SHAIYVA ANAND

(1404551044)

SUNIL SAROJ

(1404551049)

Submitted to:

DR. ASWINI SOOD

Supervisor

DR. S. K. GUPTA

H.O.D

DEPARTMENT OF CHEMICAL ENGINEERING

HARCOURT BUTLER TECHNICAL UNIVERSITY

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CERTIFICATE

This is to certify the students listed below are enrolled in the Chemical Engineering Department of HBTU, Kanpur.

- **Mr. Aditya Singh,**
(Roll No. 1404551002)
- **Mr. Rohit Goswami**
(Roll No. 1404551040)
- **Ms. Shaivya Anand**
(Roll No. 1404551044)
- **Mr. Sunil Saroj**
(Roll No. 1404551049)

They are in good standing and have completed their B.Tech project as per the requirements of obtaining their degree to the satisfaction of the undersigned.

Prof. Ashwini Sood,

Project Guide

Prof. S.K. Gupta

HOD, ChemE

ABSTRACT

In this project we have studied and validated the analysis of a gas sweetening unit. We have delved into computational generalizations to extend our analysis and have also used model data and parameters from an existing plant.

The process control aspects have been shown with reference to standard flow sheets. Cost estimates and analysis is covered with the help of standard indexes and known profitability measures. Hazard management and process descriptions have also been indicated.

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Project Overview



Figure 1.1 – Plant Structure

Problem Definition

The main purpose of the gas treatment step is to remove the H₂S from the sour gas. Using the principles of mass transfer and organic chemistry, the project involves suggestion of a suitable solvent, design of absorption and stripping column, process description and a relevant material and energy balance of the overall plant giving details of the major equipment used in the plant as well as cost estimation and economic feasibility of the project. The plant structure is shown in Fig. 1.1.

Project Goals

- Introduction to ammonia based gas sweetening units for refining natural gas
- Description, comparison and justification of the methods used in the process
- Study of kinetics and unit operations involved in the process
- Detailed process description and flow diagram
- Material and Energy balance of the plant
- Design (mechanical and process) design of plant equipment
- Plant Layout and Instrumentation
- Safety Analysis
- Computational modeling and analysis
- Standards and environment oriented design
- Cost estimation and economic optimization

Part I

Process Description

The gas sweetening unit basically consists of two columns viz. absorption column and stripping column. The other nuances include several valves, flash drum and preheaters or plate heat exchangers to maintain the required temperature and pressure in the column.

The overall process is described with the flowsheet in the IPC section and in Fig. 2.1.

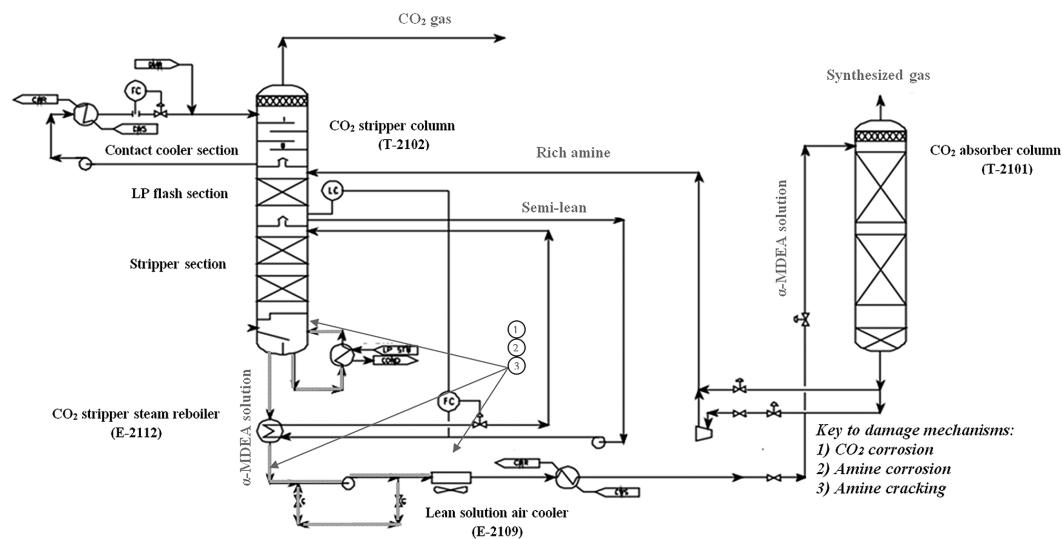


Figure 2.1 – MDEA sample unit

MDEA Process Principles

Methyl-Di-Ethanol-Amine (MDEA) is tertiary amine, which does not react easily with CO₂. The selectivity is so prompted by using the differences in the reaction rates between both H₂S and CO₂ and tertiary amine.

- The selective removal of H₂S is made by washing the sour gas with an aqueous solution of Methyl-Di-Ethanol-Amine (MDEA).
- The process principles are similar to the well known SNPA-DEA process.

The only difference between the aforementioned is the behavior of the ethanol amine used.

We first investigate the case a primary or secondary amine (mono-ethanolamine or di-ethanolamine) whose reactions with the acid components H₂S and CO₂ are similar.

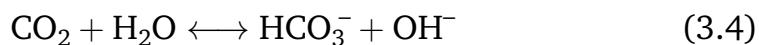
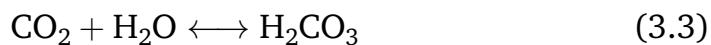
- H₂S reacts to give amine hydrosulfide:



CO₂ can react directly with amine to form an amine carbonate:



But CO₂ can also react with water or hydroxyl ions to form carbonic acid or bicarbonate ions



These acids then react with the amine to form amine bicarbonate (H_2CO_3^- , RNH_2^+) and amine carbonate (CO_2 , R_2NH_2^+).

From kinetic considerations, three types of reactions can be distinguished.

- Reaction (Eq. 3.1) whose rate is infinite
- Reaction (Eq. 3.2) whose rate is moderate, depending on the amine
- Reaction (Eq. 3.3) and (Eq. 3.4), known to be slow.

It is known that using MEA or DEA the absorption rate CO_2 in the absorber may be lower than the absorption rate of H_2S ; however CO_2 removal is regarded as complete. The case of tertiary amine is different. As a matter of fact, the molecular structure of the tertiary amine prevents the direct reaction of CO_2 with carbonate formation (reaction (Eq. 3.2)).

Contact time depends on:

- The gas flow rate
- The liquid height above the active plant area
- The number of plates in the absorber

The first two parameters cannot be acted upon. The third parameter (plate quantity) allows adjusting contact time according to the feeding conditions and required performance.

We show here the average size of a standard unit meant for handling this substance in Fig. 3.1.



Figure 3.1 – Real amine processing plant

Absorption Section

The absorption section operation is as enumerated below by following along the path taken by the raw gas.

1. Slug-catcher derived raw gas is heated in steam heater and made to pass through a pressure regulating control valve.
2. Its temperature is maintained above 25°C by controlling the heater steam flow.
3. This raw gas is then mixed with the condensate treating unit derived raw gas, which is made up of light hydrocarbon fraction in the raw condensate.
4. The raw gas mixture then passes through a knock out drum where liquid carryovers (water, hydrocarbon) are retained to be pumped to the condensate treating unit.
5. Upon leaving separator raw gas enters an amine absorption column where it comes into contact counter currently with the aqueous solution of MDEA.
6. The absorber column includes 14 valve trays. The lean MDEA solution at 45°C at the column upper part.
7. Different liquid feeds allow the possibility of selecting an adequate number of the trays to obtain the required results taking into account about feed gas quality and operating conditions.
8. The selected feed position could be to the trays 1, 3, 5, 7, or 9 and the switching over from one to another is done manually.
9. The MDEA feed nozzles are arranged such that the liquid feed is fed to the downcomers of trays (1, 3, 5, 7 and 9) and not the trays themselves.
10. The upper tray is dry and retains the major part of carryovers.
11. The total lean MDEA solution flows to the column under flow control valve.
12. At the top of absorber, treated gas contains less than 4 ppm volume of H_2S . Temperature ranges from 40°C to 44°C . The gas is cooled to 38°C by passing through a water cooler.

Gas Feedstock Specifications

The following mixed gas composition (mixture of slug catcher gas and CFU off gas) has been considered for the design of GSU and GDU.

Pressure at battery limits : $96 - - 54 \text{ kg/cm}^2 \text{ abs}$

Temperature at battery limits : $20 - 33^\circ\text{C}$

Gas sweetening train will have a maximum operating pressure of 77 kg/cm^2 . A pressure control valve and heating device upstream of the pressure valve will prevent hydrate formation. H_2S and CO_2 content will vary depending on conditions.

Product Specifications

The product from each Gas Sweetening train will meet the following specifications :

Sweet and Dry Gas

H_2S content : 4 ppm volume maximum

Pressure at battery limits : $74.9 - 51.9 \text{ kg/cm}^2 \text{ abs}$

Temperature at battery limits : 40°C

Rich MDEA Flash Drum:

H_2S content : 4 ppm volume maximum

Pressure at battery limits : $4.5 \text{ kg/cm}^2 \text{ abs}$

Temperature at battery limits : $48 - 55^\circ\text{C}$

Acid Gas:

Hydrocarbon content will be minimized.

Pressure at Battery Limits : 1.9 kg/cm^2

Temperature at Battery Limits : $50^{\circ}C$

Turndown Ratio

Overall turn down ratio to 40% of the design capacity can be obtained without any special additions to the standard features of the equipment. With this turndown ratio of 40%, design sweetening and drying performances will be maintained, with an increase in the specific energy consumption.

Major process equipment for the gas sweetening unit are:

- Absorption column.
- Stripping column.

The main process variables in the absorption and stripping columns are:

- Temperature.
- Pressure.
- Flow Rate.

Absorption Tower Calculations

- Tray Tower
- Valve Tray
- MDEA solvent
- Low Temperature and High Pressure in absorption column
- High Temperature and Low Pressure in stripping column
- Countercurrent operations

The above operations are elaborated.

Part II

Literature Review

Tall Tower Design Theory

The design of tall towers, here taken to be the mechanical design of the absorption and stripping towers is crucial to the stable operation of the plant. Hence we must first determine the concept of construction behind towers of large scales.

Theory of Tall Vertical Vessels

Self-supporting tall vessels are widely used in chemical process industries. Tall vessels may or may not be designed to be self-supporting. Distillation column, fractionating columns, absorption tower, multi-stage reactor, stacks, chimneys etc. comes under the category of tall vertical vessels. In earlier times high structure (i.e. tall vessels) were supported or stabilized by the use of guy wires.

Design of self-supporting vertical vessels is a relatively recent concept in equipment design. and it has been widely accepted in the chemical industries because it is uneconomical to allocate valuable space for the wires of guyed towers. In these units ratio of height to diameter is considerably large due to that these units are often erected in the open space, rendering them to wind action. Many of the units are provided with insulation, number of attachments, piping system etc.

For example, distillation and absorption towers are associated with a set of auxiliary equipment i.e. reboiler, condenser, feed preheater, cooler and also consists of a series of internal accessories such as plates or trays or variety of packings.

Often the vertical vessels/columns are operated under severe conditions, and type of the material these columns handles during operation may be toxic, inflammable or hazardous in other ways. Structural failure is a serious concern with this type of columns. As a result, the prediction of membrane stresses due to internal or external pressure will not be sufficient to design such vessels. Therefore, special considerations are

necessary to take into account and predict the stresses induced due to dead weight, action of wind and seismic forces.

Shell Stresses

The primary stresses are basically due to:

- (a) Circumferential stress, radial stress and axial stress due to internal pressure or vacuum in the vessel
- (b) Compressive stress caused by dead load such as self weight of the vessel including insulation, attached equipment and weight of the contents

Loads

Dead loads This is the weight of a structure itself, including the weight of fixtures or equipment permanently attached to it.

Live loads This is moving or movable external load on a structure and includes the weight of furnishing of building, of the people, of equipment etc. but doesn't include wind load.

Wind Loads

If the vessels are located in open, it is important to note that wind load also act over the vessel. Under wind load, the column acts as cantilever beam as shown in figure. Therefore while designing the vessel stresses induced due to different parameters have to be considered such as:

- (i) compressive and tensile stress induced due to bending moment caused by wind load acting on the vessel and its attachments;
- (ii) stress induced due to eccentric and irregular load from piping, platforms etc.
- (iii) stress induced due to torque about longitudinal axis resulting from offset piping and wind loads and

(iv) stress resulting from seismic forces. Apart from that, always there are some residual stresses resulting due to methods of fabrication used like cold forming, bending, cutting, welding etc

Under the loads discussed, the generalized bending moment diagram is shown in Fig. 6.1.

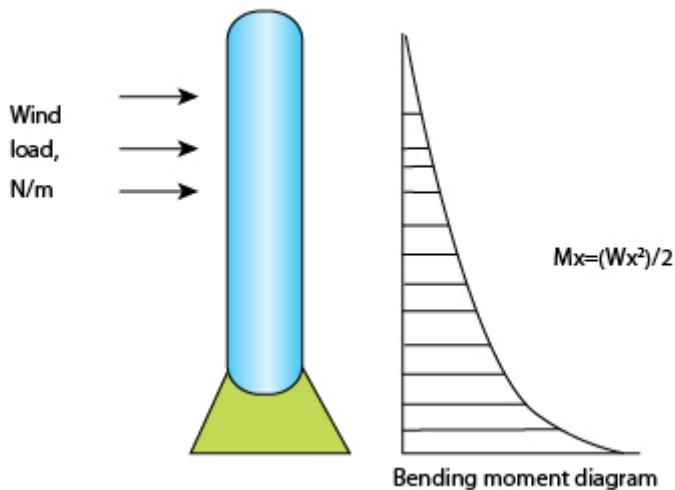


Figure 6.1 – Bending Moment

Pressure Stresses

We shall develop the general form of the equations for axial and circumferential pressure stresses resulting from internal and external pressures.

Tensile stresses resulting from internal pressure

The simple equation may be derived to determine the axial and circumferential stresses due to internal pressure in the shell of a closed vessel. Fig. 6.2 shows a diagram representing a thin walled cylindrical vessel in which a unit form stress, f , may be assumed to occur in the wall as a result of internal pressure.

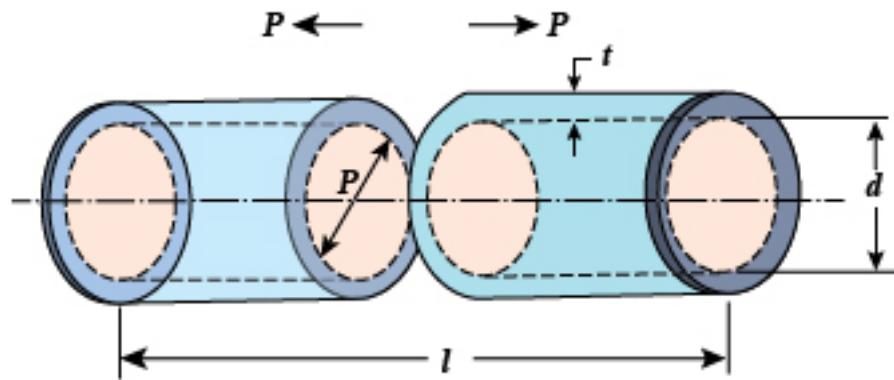


Figure 6.2 – Longitudinal forces acting on thin cylinder (internal pressure)

We adopt the following notation:

- l is the length (inches)
- d is the inside diameter (inches)
- t is the thickness of the shell (inches)
- p is the internal pressure (pounds per square inch gage)

Longitudinal Stresses

Limiting ourselves to analyzing only the pressure stresses, the longitudinal force, P resulting from an internal pressure, p acting on a thin cylinder of thickness t , length l and diameter d is:

$$P = \frac{p\pi d}{4} \quad (6.1)$$

$$a = t\pi d \quad (6.2)$$

Where we note that:

- P is the force tending to rupture the vessel longitudinally
- a is the area of metal resisting longitudinal rupture

Therefore we note that

$$f = \text{stress} = \frac{P}{a} = \frac{pd}{4t}$$

$$t = \frac{pd}{4f} \quad (6.3)$$

Circumferential Stresses

Considering Fig. 6.3 we shall consider on the circumferential stresses induced by internal pressure.

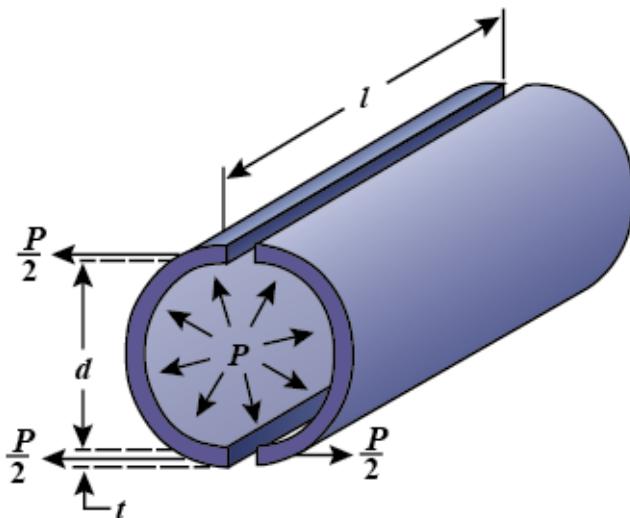


Figure 6.3 – Circumferential forces acting on thin cylinders (internal pressure)

We note that:

- $P = p \times d \times l$ is the force tending to rupture the vessel circumferentially
- $a = 2 \times t \times l$ is the area of the metal resisting the forces

Therefore we obtain

$$f = \text{stress} = \frac{P}{a} = \frac{pd}{2t} \quad (6.4)$$

$$t = \frac{pd}{2f} \quad (6.5)$$

Analysis

Equation 6.5 and Equation 6.3 indicate that for a specific allowable stress, fixed diameter and given pressure, the thickness required to restrain the pressure for the condition of Equation 6.5 is double than that of the Equation 6.3. Therefore, the thickness as determined by Equation 6.5 is controlling and is the commonly used thin walled equation referred to in the various codes for vessels.

The above equation development makes no allowances for corrosion and does not recognize the fact that welded seams or joints may cause weakness. Experience has shown that an allowance may be made for such weakness by introducing a joint efficiency factor j in the equations and this factor is always less than unity and is specified for a given type of welded construction in the various codes. The thickness of metal, c , allowed for any anticipated corrosion is then added to the calculated required thickness, and the final thickness value rounded off to the nearest nominal plate size of equal or greater thickness.

Based on this we may now rewrite Equation 6.5 and Equation 6.3 to be in the form:

$$t = \frac{pd}{4fj} + c \quad (6.6)$$

$$t = \frac{pd}{2fj} + c \quad (6.7)$$

Where:

- l is the length (inches)
- d is the inside diameter (inches)
- t is the thickness of the shell (inches)
- p is the internal pressure (pounds per square inch)
- f is the allowable working stress (pounds per square inch)
- E is the joint efficiency (dimensionless)
- c is the corrosion allowance (inches)

Compressive stresses due to dead loads

The major sources of the load acting over tall vertical vessels are the weight of the vessel shell and weight of the vessel fittings which includes the internal, external and auxiliary attachments.

Internal fittings trays, packing, heating and cooling coils etc.

External fittings platforms, piping, insulation, ladders etc.

Auxiliary attachments instruments, condensers etc.

Therefore, stresses caused by dead loads may be considered in three groups for convenience:

- (a) stress induced by shell and insulation
- (b) stress induced by liquid in vessel
- (c) stress induced by the attached equipment

Stress induced by shell and insulation

We assert that the stress due the weight of shell and insulation at any distance X from the top of a vessel having a constant shell thickness is given by Eq. 6.8.

$$W_s = \frac{\pi}{4} \times (D_o^2 - D_i^2) \times \rho \times X \quad (6.8)$$

Where:

- W_s is the weight of shell above point X from the top
- D_i and D_o denote the inner and outer shell diameters
- X is the distance measured from the top of the vessel
- ρ is the density of the shell material

Also for the insulation as height X we have Eq. 6.9

$$W_i = \pi \times D_m \times \rho \times X \times t \quad (6.9)$$

Where:

- W_i is the weight of the insulation
- D_m is the mean diameter of insulation
- X is the height measured from the top of the column
- t is the thickness of the insulation
- ρ is the density of the insulation

As we define the compressive stress to be the force per unit area we may now write Eq. 6.10 and Eq. 6.11.

$$f_{d_{wtshell}} = \frac{\frac{\pi}{4}(D_o^2 - D_i^2) \times X \times \rho_s}{\frac{\pi}{4}(D_o^2 - D_i^2)} = X\rho_s \quad (6.10)$$

$$f_{d_{wtins}} = \frac{\pi \times (D_{mpt})_{ins} X}{\pi D_m t_s} = \frac{\rho_{ins} t_{ins} X}{t_s} \quad (6.11)$$

Where:

- D_m is the mean diameter of the shell given by $D_m = \frac{D_s + D_i}{2}$
- t is the thickness of shell without corrosion allowance

Stress induced due to liquid retained in column

This depends on the internal configuration of the column as the major hurdle is in determining accurately the amount of liquid. Hence we state Eq. 6.12 recognizing that there is a dependency on the tray column diameter, number of plates, holdup over each plate, liquid in the downcomer etc.

$$f_{d_{liquid}} = \frac{\sum W_{liquid}}{\pi D_m t_s} \quad (6.12)$$

Where:

- D_m is the mean vessel diameter (feet)
- t is the thickness of shell without corrosion allowance

Stress induced by attachments

Similar to the liquid stresses, the stresses due to attachments are calculated on the basis of the weight of attachments like trays, overhead condensers, platforms, ladders etc.

Hence we have Eq. 6.13 in a manner similar to Eq. 6.12.

$$f_{d_{\text{attachments}}} = \frac{\sum W_{\text{attachments}}}{\pi D_m t_s} \quad (6.13)$$

Analysis

From the forgoing analysis we may now write the total dead loss stress, f_{total} acting along the longitudinal axis of the shell to be a simple sum as shown in Eq. 6.14.

$$f_{\text{total}} = \sum_{i=\text{liq,shell,etc}} f_i \quad (6.14)$$

Axial stresses due to wind loads

The stress due to wind load may be calculated by treating the vessel as uniformly loaded cantilever beam. The wind loading is a function of wind velocity, air density and shape of tower and is given by Eq. 6.15

$$P_w = \frac{1}{2} \times C_D \times \rho \times V_w^2 \times A \quad (6.15)$$

Where:

- C_D is the drag coefficient
- ρ is the density of the surrounding fluid
- V_w is the wind velocity
- A is the projected area normal to the direction of wind

If wind velocity is known approximate wind pressure can be computed from the simplified relationship given in Eq. 6.16.

$$P_w = 0.05V_w^2 \quad (6.16)$$

Where:

- P_w is the minimum wind pressure to be used for the moment calculation (Newton per meter square)
- V_w is the maximum wind velocity experienced by the region under the worst weather condition (kilometers per hour)

Wind velocity varies with height. This can be seen in Fig. 6.4. The velocity of wind near the ground is less than that away from it. Therefore, to take into account this factor a variable wind force may be taken.

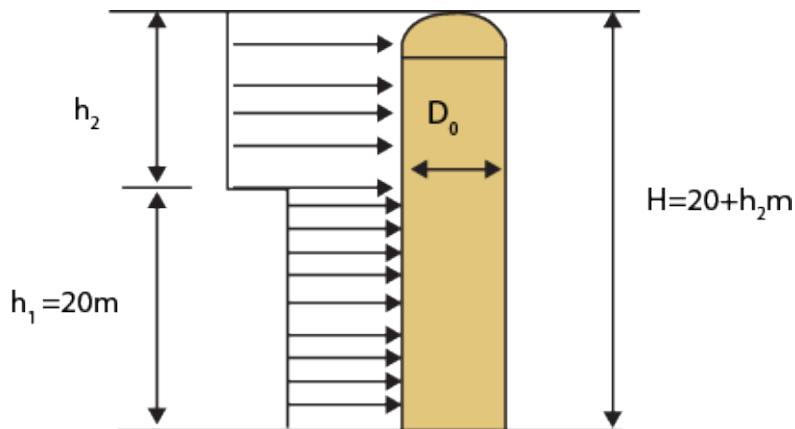


Figure 6.4 – Tall column subjected to wind pressure

It is recommended to calculate the wind load in two parts, because the wind pressure does not remain constant through the height of the tall vessel. Say for example in case of vessel taller than 20m height, it is suggested that the wind load may be determined separately for the bottom part of the vessel having height equal to 20m, and then for rest of the upper part.

Load due to wind acting on the bottom portion of the vessel

$$P_{bw} = K_1 K_2 p_1 h_1 D_o$$

Where:

- P_{bw} is the total force due to wind load acting on the bottom part of the vessel with height equal to or less than 20m
- D_o is the outer diameter of the vessel including the insulation thickness
- h_1 is the height of the bottom part of the vessel equal to or less than 20m
- K_1 is the coefficient depending upon the shape factor (i.e. 1.4 for flat plate; 0.7 for cylindrical surface)

Load due to wind acting on the upper portion of the vessel

$$P_{uw} = K_1 K_2 p_2 h_2 D_o$$

Where:

- P_{uw} is the total force due to wind load acting on the upper part of the vessel, above 20m
- D_o is the outer diameter of the vessel including the insulation thickness
- h_2 is the height of the upper part of the vessel, above 20m
- K_1 is the coefficient depending upon the shape factor (i.e. 1.4 for flat plate; 0.7 for cylindrical surface)
- K_2 coefficient depending upon the period of one cycle of vibration of the vessel

We note that $K_2 = 1$, if period of vibration is 0.5 seconds or less; $K_2 = 2$, if period exceeds 0.5 seconds.

Stress due to bending moment

For the stress induced by the wind due to the bending moment in the axial direction we shall note that we may write Eq. 6.17 and Eq. 6.18.

$$M_w = P_{bw} \frac{h_1}{2} \quad (6.17)$$

$$M_w = P_{bw} \frac{h_1}{2} + P_{uw} \frac{h_1 + h_2}{2} \quad (6.18)$$

From which we can assert that the bending stress is given by Eq. 6.19

$$f_w = \frac{4M_w}{\pi t(D_i + t)D_i} \quad (6.19)$$

Where:

- f_w is the longitudinal stress due to wind moment
- M_w is the bending moment due to wind loads
- D_i is the inner diameter of shell
- t is the corroded shell thickness

Tall Tower Height Estimation

Height of the tall vessel X can be estimated by combining all the stresses acting in the axial direction may be added and equated to the allowable tensile stress, excluding the stresses due to eccentricity of load and seismic load.

Hence we have:

$$\frac{\sum W_{attachments}(X)}{\pi D_m t_s} = f_{t_{max}}$$

$$\frac{\sum W_{attachments}(X)}{\pi D_m t_s} - f_{tall} J = 0$$

Where:

- t_s is the shell thickness

From the above equations, on replacing f_{tmax} by f_{tall} we get the quadratic form. $aX^2 + bX + c = 0$

The solution to this is well known to be simply:

$$X = \frac{-b \pm \sqrt{b^2 - 4ac}}{2a} \quad (6.20)$$

Once the value of X is estimated, it is used to adjust the plate thickness, t , for the top portion of the column, so that the height of portion X will be multiple of the plate width used. The plate thickness which is originally selected is satisfactory up to a considerable height. Trays below the distance X of the column must have an increased thickness. If the above condition does not satisfy then calculation of the axial stress with an increase in the thickness according to Eq. 6.8 and Eq. 6.20 are repeated, and this repetitive steps in calculation helps to estimate subsequent height ranges to corresponds with increased thickness. The procedure is repeated till the entire height of the vessel is covered.

Column Internals

Design and construction features of plates and trays

Plate or trays can be constructed either as one piece trays or as sectional trays. Several factors control the design and construction features of plates or trays. These factors include:

- down coming liquid impact, liquid weight, load on the tray due to dead weight
- expansion due to rise in temperature
- fabrication and installation ease
- support type

- material of construction and safety

One piece trays may be made of materials such as cast iron, copper or steel including the risers and downcomers, with a thickness of 2 to 6 mm depending on the diameter and the material.

The sectional tray is made from section in the form of floor plates cut from sheets, which are laid on the supporting beams and peripheral ring. A clearance is provided between adjacent sections and clamping devices are used for fixing.

The cast iron tray is able to withstand compressive forces created due to thermal expansion within reasonable limits and their diameters are also limited to small sizes. Whereas the one piece shaped tray made of ductile material is comparatively thin and has a limited ability to absorb forces due to thermal expansion. Therefore, in order to prevent the distortion of the tray floor, provision should be made for a packing seal between the edge of the tray and column wall.

On the other hand one of the main advantages of the sectional tray is its ability to cope with thermal expansion. The individual sections of the tray are placed on the supporting structures, an asbestos jointing material inserted between the section and the support member. Each section is finally held by frictional clamping devices. Sectional trays are also necessary when these are to be taken inside through the limited size of column man holes in parts and assembled inside.

Loading conditions of trays and plates

Plates and trays used in the tall column have to be maintained flat in order to provide a uniform seal of the liquid on their surfaces. During operation various loads acts on the plates and trays, and due to that plates and trays are likely to deflects greatly, unless they are provided with sufficient supporting systems or and made adequately thick.

Deflections caused by the different loads are:

- (a) tray weight with contacting devices and downcomers
- (b) liquid weight

- (c) impact load of the down coming liquid
- (d) weight of maintenance personnel and tools
- (e) expansion due to a rise in temperature (if prevented).. Usually these have provision for free expansion. Load due to this may be ignored.

Quantitatively we estimate the load via Eq. 6.21.

$$\text{Load} = \frac{w\nu}{g} \quad (6.21)$$

Where:

- w is the weight of liquid per second
- ν is the velocity per second
- g is the gravitational constant

Loads denoted by item (a) and item (c) need to be considered for realistic estimations of the deflection. In general, a deflection of 3mm permissible and in special cases this is may be limited to 2mm.

Similarly, for cleaning and assembly or inspection operations, loads in item (a) and item (e) should be considered. The design is based not on the permissible deflection but on the permissible stresses. # Column Internal Design Theory

Having determined the exterior of the tower and it's mechanical properties we shall now move towards an understanding of the mechanical properties of the column internals, which involve plates, closures (heads) and their choices and performance reviews.

Sectional Construction

The trays, downcomer segments and other tray components are usually constructed in sections for column diameter of 1m and above. The plate sections are designed in such that it can be removed through the column manholes, preferably detachable from both above tray and tray

below. The plate sections are installed usually starting from one side on a peripheral supported ring welded to inside shell wall.

The support ring is not normally extended into the downcomer area and tray perforations needs to be avoided over the support ring area. Support ring width is usually between 40mm to 90mm. Trays are either clamped or bolted to the support ring.

The support beams prevent tray deflection under load for larger diameter column. One or more major beams are required for large diameter column (usually < 3m). Trays can be supported by support rings and minor beams for small diameter column (usually < 3m) i.e. omitting the major beams.

The beams are commonly placed at about 0.6m apart and should able to hold the necessary weight during column operation as well as installation. The maximum plate deflection under the operating conditions should be less than 3mm for tower diameter more than 2.5m. The number of plate sections installed under an apron and across the accumulator trays is minimized. Clearance should be provided between the bottom of the beams and the tray below for the perpendicular installation of the beams to the liquid flow direction on the tray below Fig. 6.5 i.e. in case of cross flow.

Cartridge Construction

The cartridge or stacked type plate construction is used for column diameters smaller than 1m as in such small diameter columns man entry to the column for the installation and necessary maintenance work is difficult. Cartridge constructions are employed for sieve, valve and bubble cap trays. Cartridge assemblies of 10 plates or so, are formed with the help of screwed rods and plate spacer bars.

The tall columns are divided into multiple flanged sections and the prefabricated cartridge tray assemblies are installed in each flanged sections with suitable downcomer clearances. The plates are not fixed to the shell wall and leakages may occur. A vertical metal seal is

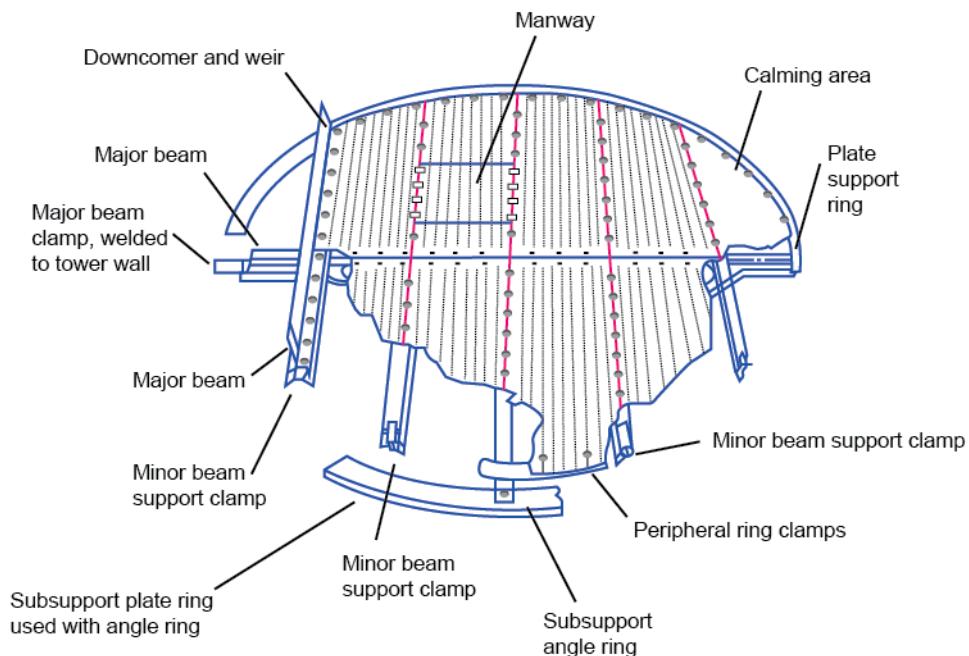


Figure 6.5 – Common sectional construction of plate

frequently used around each tray to decrease leakage around the tray edge.

Classical cartridge construction is shown in Fig. 6.6.

Downcomer Details

The liquid from the top tray is directed to the tray below through a downcomer or downspout. The downcomer liquid flow passage is normally formed by a vertical plate starting from the weir outlet, called an apron. The adequate residence time needs to be ensured in the downcomers for the disengagement of vapor/gas from liquid. The downcomer residence time of more than 3s is desirable to get only clear liquid on the tray. The different types of segmental downcomers constructions are shown in Fig. 6.7. The bottom downcomer area is normally more than 50% of the top downcomer area for sloped downcomers.

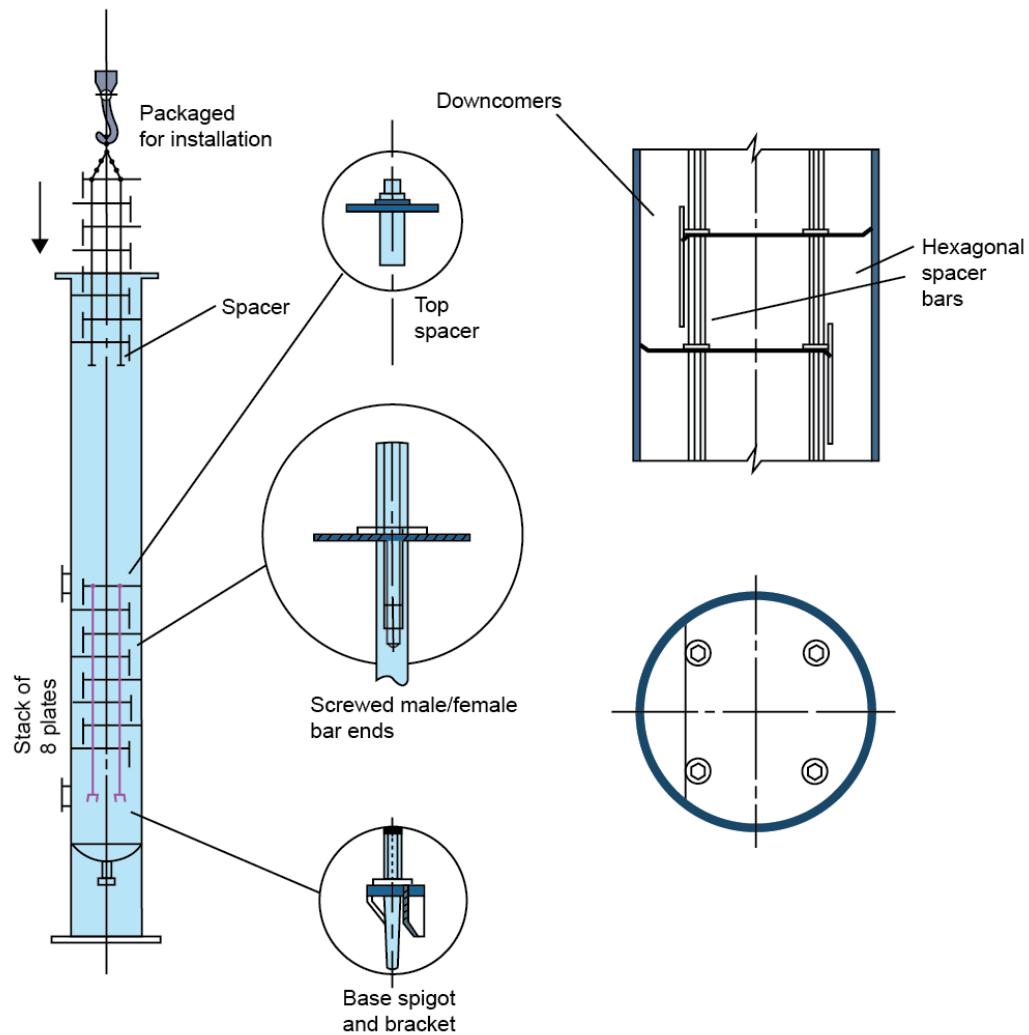


Figure 6.6 – Common plate cartridge construction

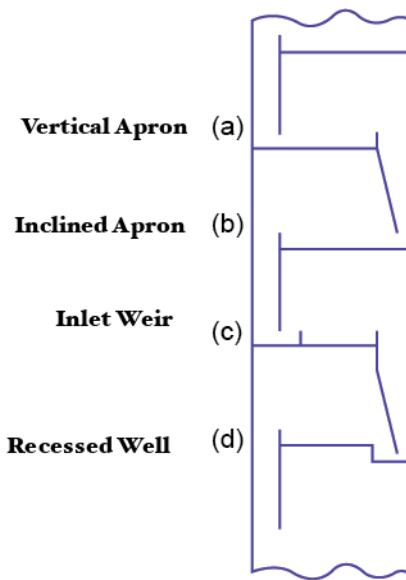


Figure 6.7 – Downcomer segmental design

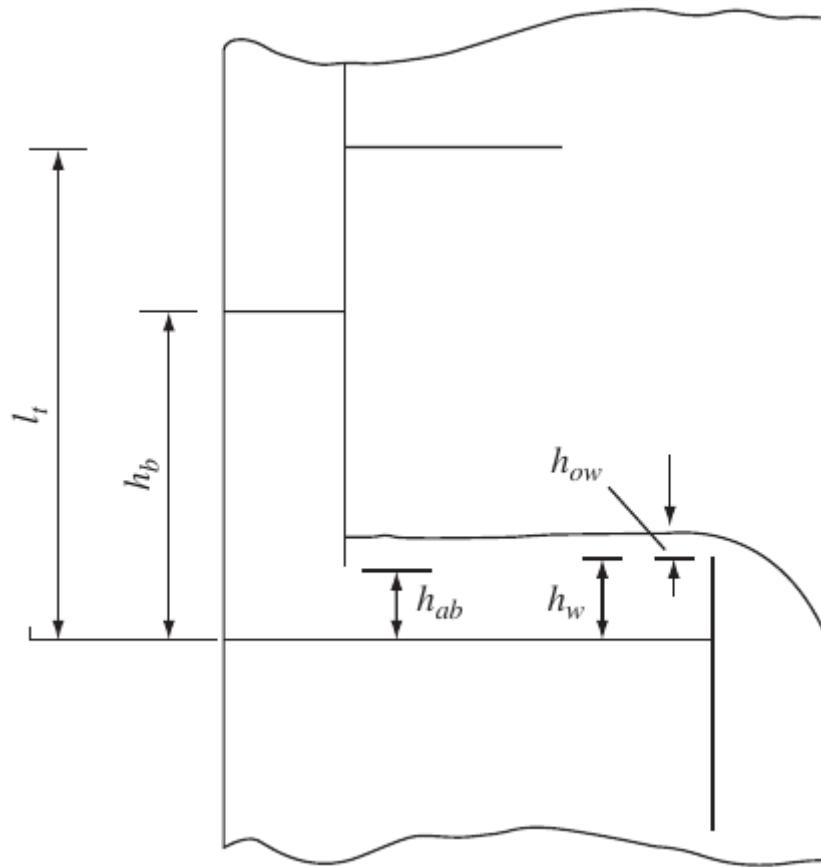


Figure 6.8 – Downcomer backup

Design

The downcomer area and plate spacing must be such that the level of the liquid and froth in the downcomer is well below the top of the outlet weir on the plate above. If the level rises above the outlet weir the column will flood. The backup of liquid in the downcomer is caused by the pressure drop over the plate (the downcomer in effect forms one leg of a U-tube) and the resistance to flow in the downcomer itself; see Fig. 6.8

In terms of clear liquid the downcomer backup is given by:

$$h_b = (h_w + h_{ow}) + h_t + h_{dc} \quad (6.22)$$

Where:

- h_b is the downcomer backup from plate surface in mm
- h_{dc} is the head loss in the downcomer in mm

The main resistance to flow will be caused by the constriction at the downcomer outlets and the head loss in the downcomer can be estimated using the equation given by Cicalese et al. (1947):

$$h_{dc} = 166 \left[\frac{L_{wd}}{\rho_L A_m} \right]^2 \quad (6.23)$$

Where:

- L_{wd} is the liquid flow rate in the downcomer in kg/s
- A_m is either the downcomer area A_d or the clearance area under the downcomer A_{ap} , whichever is smaller in m^2

$$A_{ap} = h_{ap} l_w \quad (6.24)$$

Where:

- h_{ap} is the height of the bottom edge of the apron above the plate.

$$h_{ap} = h_w - 5 \text{ to } 10 \text{ mm}$$

Froth Height To predict the height of „aerated“ liquid on the plate, and the height of froth in the downcomer, some means of estimating the froth density is required. The density of the „aerated“ liquid will normally be between 0.4 to 0.7 times that of the clear liquid. For design purposes it is usually satisfactory to assume an average value of 0.5 of the liquid density. This value is also taken as the mean density of the fluid in the downcomer, which means that for safe design the clear liquid backup, calculated from equation Eq. 6.22, should not exceed half the plate spacing, to avoid flooding. Allowing for the weir height:

$$h_b \leq \frac{1}{2}(l_t + h_w)$$

This criterion is typically considered to be over safe, and where close plate spacing is desired a better estimate of the froth density in the downcomer should be made.

Downcomer Residence Time Sufficient residence time must be allowed in the downcomer for the entrained vapor to disengage from the liquid stream to prevent heavily „aerated“ liquid being carried under the downcomer. A time of at least 3 seconds is recommended. The downcomer residence time is given by

$$t_r = \frac{A_d h_{bc} \rho_L}{L_{wd}} \quad (6.25)$$

Where:

- t_r is the residence time, (s)
- h_{bc} is the clear liquid back-up (m)

Feed Inlets and Side Stream Draws

The feed liquid is commonly introduced vertically downward into the downcomer of the tray above the feed point through an open pipe. A slotted branch distributor is used for of flashing feed (liquid) to avoid early flooding as shown in Fig. 6.9.

The vapor generally enters the column tangentially through an open inlet nozzle placed almost at right angles to the liquid flow on the above tray. The clearance between the feed pipe end and feed plate below is normally kept more than 0.4 m when the feed liquid is flashed in active tray area. Wider tray spacing is needed where the feed pipe and the side stream takeoff pipe to provide a liquid seal are installed.

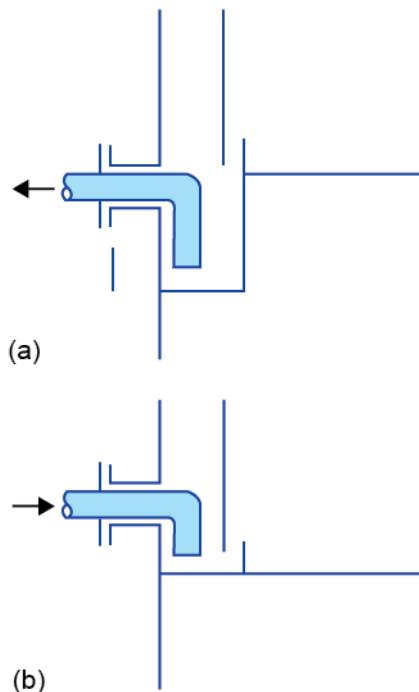


Figure 6.9 – Nozzles and inlets

Tray Miscellany

We also briefly discuss here two important practical aspects of trays, namely the drainage and manways.

Tray drainage

Some amount of liquid is usually remains trapped on a bubble cap (positive seal tray) and on certain valve trays when the column is shut

down. This liquid is normally drains through weep holes located near the outlet weirs. Weep hole diameter usually ranges from 3 to 15mm. The recommended weep hole area is 4in^2 per 100ft^2 of tray area with 50 bubble caps and 4in weir to drain in about 8hr .

Tray manways

The maintenance workers travel from one tray to another through tray manways . The size of the manways should be such that a person can safely travel through a tray manway segment and also it is removable through the column manholes. Manways are usually rectangular and the recommended size is $40\text{cm} \times 50\text{cm}$. Tray manways should be removable from top and as well as from the tray below.

Ends

We assert that under the design considerations already enumerated in the project our tray column is considered to be an unfired pressure vessel as defined by (*IS 2825: Code for Unfired Pressure Vessels n.d.*) and we shall follow the sizing accordingly.

Some common heads used for domed vessels, like that of our tray column are:

1. Hemispherical
2. Semi-ellipsoidal
3. Dished and flanged

For reasons of performance and ease of fabrication we have opted for a torispherical head which is referred to as a dished and flanged end in the relevant code. Note the structure as shown in Fig. 6.10

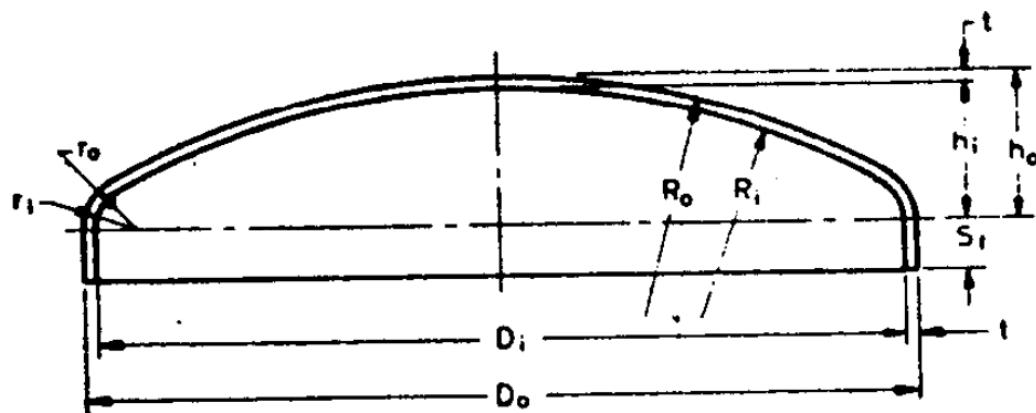


Figure 6.10 – Dished and Flanged End

We shall follow here a brief overview of heat exchangers in the manner of Shah and Sekulic (2003) for the classification schemes.

Heat Exchanger A heat exchanger is a device that is used to transfer thermal energy (enthalpy) between two or more fluids, between a solid surface and a fluid, or between solid particulates and a fluid, at different temperatures and in thermal contact. In heat exchangers, there are usually no external heat and work interactions.

Classification

Though the variance of heat exchanger forms are almost as large as that of the applications of chemical engineering itself due to their ubiquitous nature and synergistic relation to industries, we seek here to represent the vast number of possible options and discuss briefly our choice of equipment.

On The Basis Of Construction

More relevant to our discussion on the probable equipment options for sweetening plants are the classifications based on the blueprint scheme as shown in Fig. 7.2.

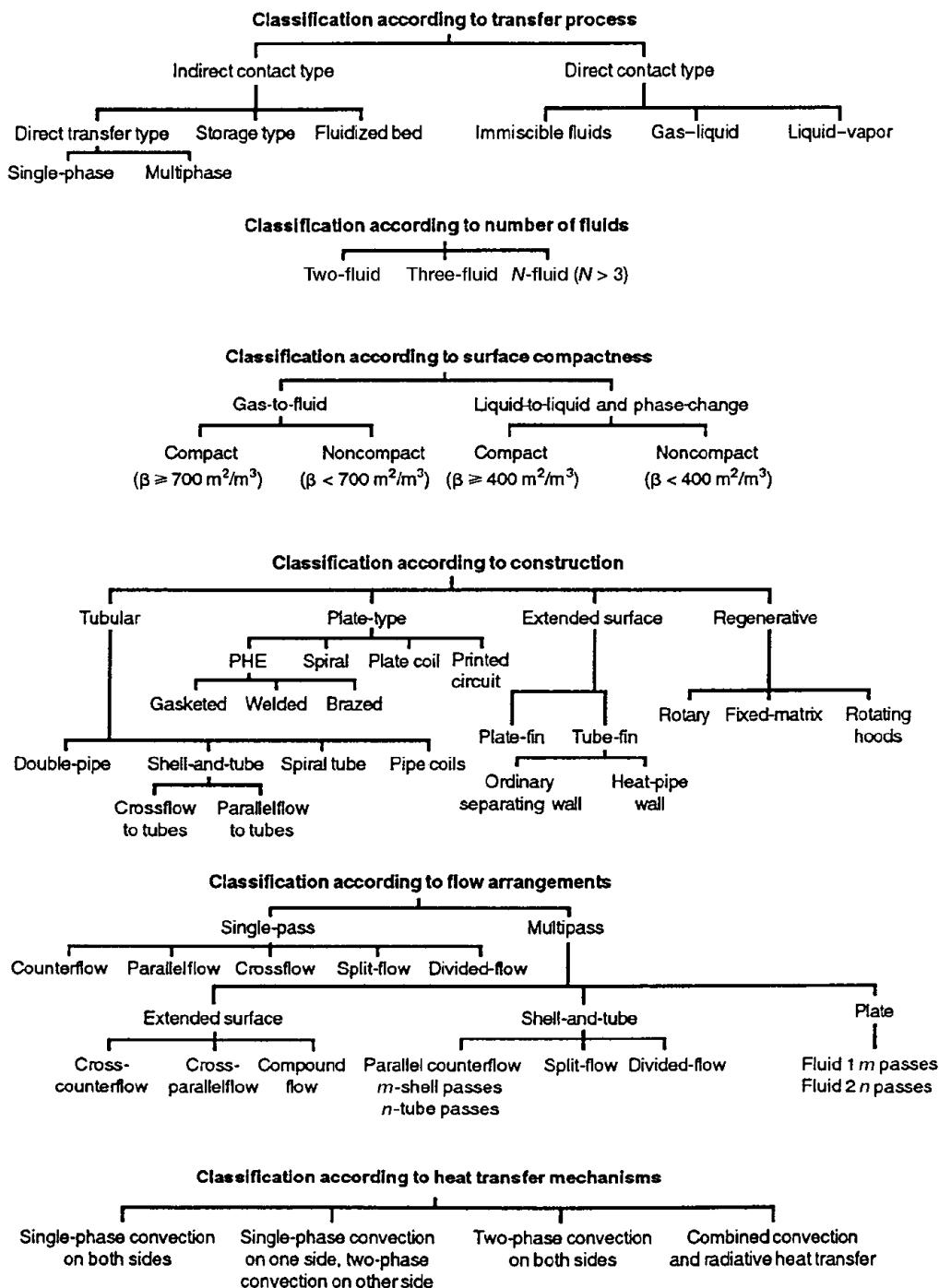


Figure 7.1 – A general classification scheme (Shah and Sekulic 2003)

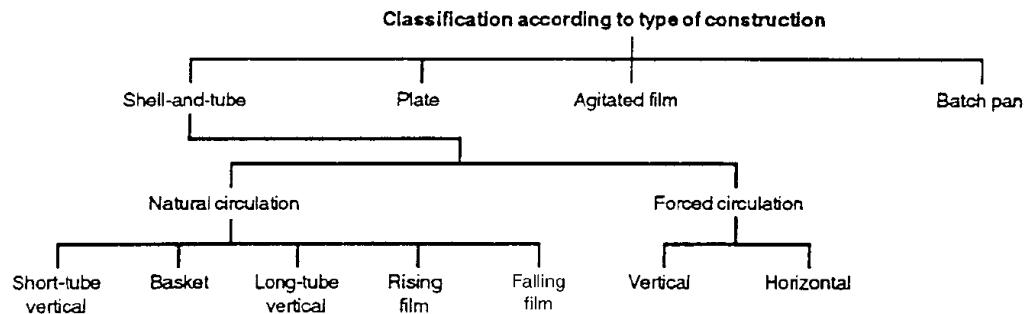


Figure 7.2 – Construction based classification (Shah and Sekulic 2003)

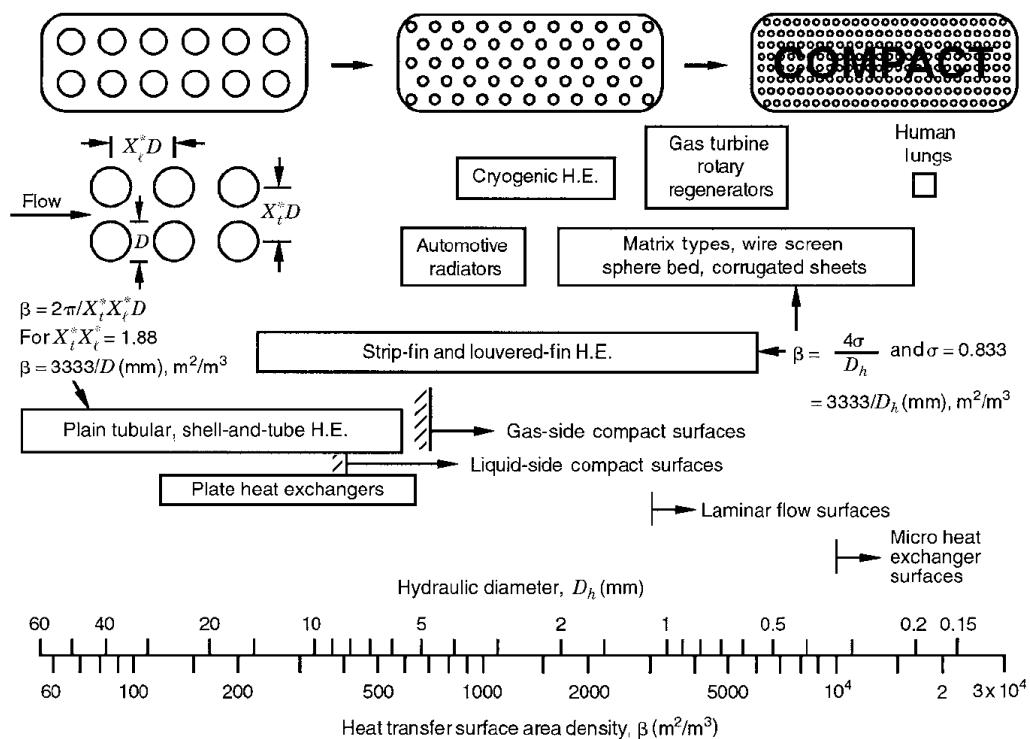


Figure 7.3 – Heat transfer surface area density spectrum of exchanger surfaces (Shah and Sekulic 2003)

Design Theory for Plate Heat Exchangers

We can design a plate heat exchanger (PHE) using the concepts of Shell and Tube Heat Exchangers by scaling up the data and calculating complex flow pressure drop using the methods assigned for Shell and Tube Heat Exchangers.

On the other hand shell and tube heat exchangers of simple construction are used as a preheaters and kettle reboilers. The process calculations of preheaters are performed previously in the Energy Balance section of the project. Mechanical construction is discussed henceforth.

Construction

The basics of plate heat exchanger design is covered masterfully by Cao (2009).

A thin planar metallic plate separating two circulating fluids at different temperatures can act as a heat transfer surface and to complete the device, two more plates must be added at both sides to form the channels where the fluids circulate.

The plates are kept separated by means of elastomeric gaskets and the assembly is clamped together in a frame by clamping bolts that must exert the necessary pressure to maintain the plates in position. A schematic drawing of a plate heat exchanger is shown in Fig. 8.2. The gaskets are such that the separation between plates is only a few millimeters to achieve high fluid velocities and hence high heat transfer coefficients. This assembly is the basis of an elemental plate heat exchanger.

The main objection to this type of device is that it has important limitations regarding the maximum flow rates that it can handle, and it would not be practical to indefinitely increase the size of the plates when higher flow rates and heat duty are required. In this case it is necessary to increase the number of plates, making both fluids circulate

through alternate channels separated by the heat transfer plates, as shown in Fig. 8.1. This is basically the arrangement of a plate heat exchanger.

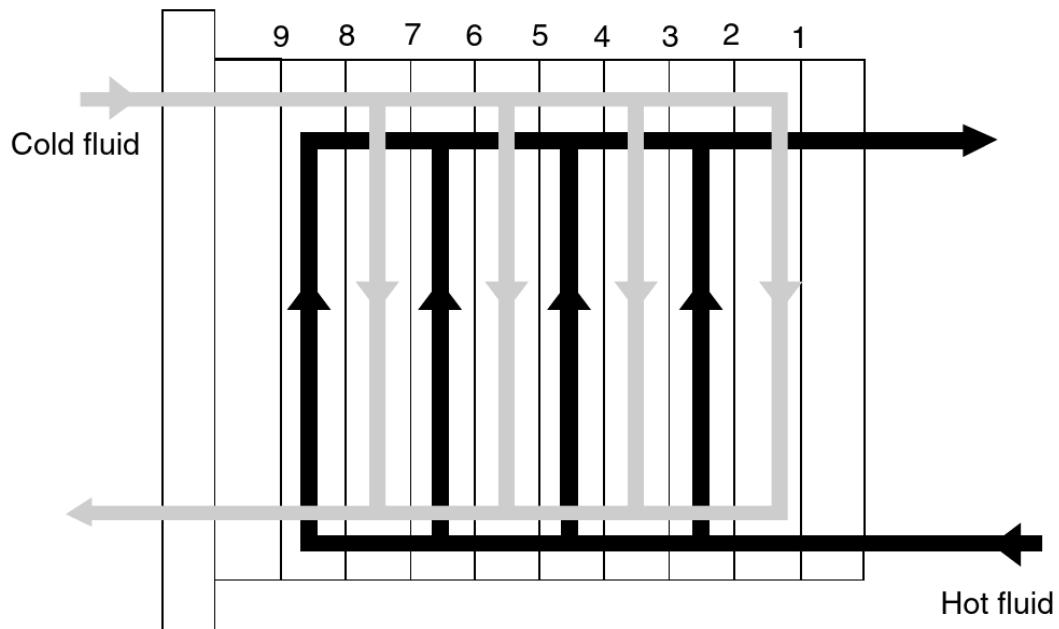


Figure 8.1 – Flow pattern of a typical plate heat exchanger

Basically, a plate heat exchanger consists of a pack of corrugated metal plates with portholes for the passage of the two fluids. Heat transfer takes place through the plates. The plate pack is assembled between a fixed frame plate and a movable pressure plate and is compressed by tightening bolts (Fig. 8.2). The plates are fitted with gaskets that seal the interplate channels. The number of plates is defined by the heat transfer requirements. The plates and pressure plate are suspended by an upper carrying bar and located by a lower guiding bar, both of which are fixed to a support column.

The plates are corrugated. The corrugations provide reinforcement as well as a large number of plate-to-plate contact points. These contact points provide full plate support and make it possible to have high operating pressures with very thin-gauge plates. The usual gap between consecutive plates is about 5 mm. The plates form the channels where the fluids circulate. The fluids enter and exit these channels through

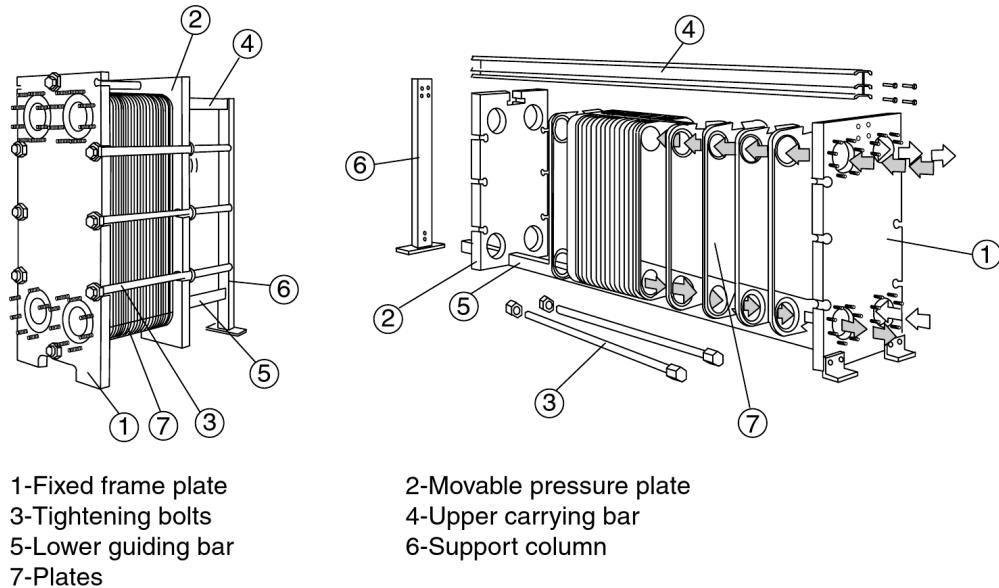


Figure 8.2 – Components of a typical plate heat exchanger

portholes located at the corners of the plates. These portholes may be perforated or not, as required by the desired circulation scheme. The portholes form the distribution headers, which distribute the fluids into the circulation channels.

The fluids enter the unit through inlet nozzles perforated in the frames and pass into the distribution headers. From the headers, the fluids distribute into the circulation channels following a specific flow pattern defined during the design stage. This is achieved by the positioning of elastomeric gaskets to blank off certain flowpaths and by blinding specific portholes of individual heat exchange plates. A suitable design of gaskets and portholes allows different flow configurations.

Shell and Tube heat exchangers are often abbreviated to S&T or even ST heat exchangers.

The flow rates of both hot and cold streams, their terminal temperatures and fluid properties are the primary inputs of thermal design of heat exchangers.

Thermal design considerations

Thermal design of a shell and tube heat exchanger typically includes the determination of heat transfer area, number of tubes, tube length and diameter, tube layout, number of shell and tube passes, type of heat exchanger (fixed tube sheet, removable tube bundle etc), tube pitch, number of baffles, its type and size, shell and tube side pressure drop etc.

Shell

Shell is the container for the shell fluid and the tube bundle is placed inside the shell. Shell diameter should be selected in such a way to give a close fit of the tube bundle. The clearance between the tube bundle and inner shell wall depends on the type of exchanger. Shells are usually fabricated from standard steel pipe with satisfactory corrosion allowance. The shell thickness of 3/8 inch for the shell ID of 12 – 24 inch can be satisfactorily used up to 300 psi of operating pressure.

Tube

Tube OD of $\frac{3}{4}$ and 1 are very common to design a compact heat exchanger. The most efficient condition for heat transfer is to have the maximum number of tubes in the shell to increase turbulence. The tube thickness should be enough to withstand the internal pressure along with the adequate corrosion allowance. The tube thickness is expressed

in terms of BWG (Birmingham Wire Gage) and true outside diameter (OD). The tube length of 6, 8, 12, 16, 20 and 24 ft are preferably used. Longer tube reduces shell diameter at the expense of higher shell pressure drop. Finned tubes are also used when fluid with low heat transfer coefficient flows in the shell side. Stainless steel, admiralty brass, copper, bronze and alloys of copper-nickel are the commonly used tube materials.

Tube pitch, tube-layout and tube-count

Tube pitch is the shortest center to center distance between the adjacent tubes. The tubes are generally placed in square or triangular patterns (pitch) as shown in the.

The number of tubes that can be accommodated in a given shell ID is called tube count. The tube count depends on the factors like shell ID, OD of tube, tube pitch, tube layout, number of tube passes, type of heat exchanger and design pressure.

Tube passes

The number of passes is chosen to get the required tube side fluid velocity to obtain greater heat transfer co-efficient and also to reduce scale formation. The tube passes vary from 1 to 16. The tube passes of 1, 2 and 4 are common in application. The partition built into exchanger head known as partition plate (also called pass partition) is used to direct the tube side flow.

Common tube layouts.

Tube OD, inch.	Pitch Type	Tube Pitch, inch.
$\frac{3}{4}$	Square	1
$\frac{3}{4}$	Triangular	$\frac{15}{16}$

Tube sheet

The tubes are fixed with tube sheet that form the barrier between the tube and shell fluids. The tubes can be fixed with the tube sheet using ferrule and a soft metal packing ring. The tubes are attached to tube sheet with two or more grooves in the tube sheet wall by „tube rolling“. The tube metal is forced to move into the grooves forming an excellent tight seal. This is the most common type of fixing arrangement in large industrial exchangers. The tube sheet thickness should be greater than the tube outside diameter to make a good seal. The recommended standards (IS:4503 or TEMA) should be followed to select the minimum tube sheet thickness.

Baffles

Baffles are used to increase the fluid velocity by diverting the flow across the tube bundle to obtain higher transfer co-efficient. The distance between adjacent baffles is called baffle-spacing. The baffle spacing of 0.2 to 1 times of the inside shell diameter is commonly used. Baffles are held in positioned by means of baffle spacers. Closer baffle spacing gives greater transfer co-efficient by inducing higher turbulence. The pressure drop is more with closer baffle spacing. The various types of baffles are shown in **Figure**. In case of cut-segmental baffle, a segment (called baffle cut) is removed to form the baffle expressed as a percentage of the baffle diameter. Baffle cuts from 15 to 45% are normally used. A baffle cut of 20 to 25% provide a good heat-transfer with the reasonable pressure drop. The % cut for segmental baffle refers to the cut away height from its diameter.

Fouling Considerations

The most of the process fluids in the exchanger foul the heat transfer surface. The material deposited reduces the effective heat transfer rate due to relatively low thermal conductivity. Therefore, net heat transfer with clean surface should be higher to compensate the reduction in

performance during operation. Fouling of exchanger increases the cost of the following:

- (a) construction due to oversizing
- (b) additional energy due to poor exchanger performance
- (c) cleaning to remove deposited materials

A spare exchanger may be considered in design for uninterrupted services to allow cleaning of exchanger.

The effect of fouling is considered in heat exchanger design by including the tube side and shell side fouling resistances. The fouling resistance (fouling factor) for petroleum fractions are available in reference books.

Selection of fluids for tube and the shell side

The routing of the shell side and tube side fluids has considerable effects on the heat exchanger design. Some general guidelines for positioning the fluids are given in Table. It should be understood that these guidelines are not ironclad rules and the optimal fluid placement depends on many factors that are service specific.

Design Theory for Storage Vessels

Storage tanks

These are containers that hold liquids, compressed gases (gas tank; or in U.S.A „pressure vessel“, which is not typically labeled or regulated as a storage tank) or mediums used for the short- or long-term storage of heat or cold. The term can be used for reservoirs (artificial lakes and ponds), and for manufactured containers.

In India, storage tanks operate under no (or very little) pressure, distinguishing them from pressure vessels. Storage tanks are often cylindrical in shape, perpendicular to the ground with flat bottoms, and a fixed frangible or floating roof. There are usually many environmental regulations applied to the design and operation of storage tanks, often depending on the nature of the fluid contained within. Above-ground storage tanks (ASTs) differ from underground storage tanks (USTs) in the kinds of regulations that are applied. Above ground storage tanks can be used to hold materials such as petroleum, waste matter, water, chemicals, and other hazardous materials, all while meeting strict industry standards and regulations.

Reservoirs can be covered, in which case they may be called covered or underground storage tanks or reservoirs. Covered water tanks are common in urban areas.

Storage tanks are available in many shapes: vertical and horizontal cylindrical; open top and closed top; flat bottom, cone bottom, slope bottom and dish bottom. Large tanks tend to be vertical cylindrical, or to have rounded corners transition from vertical side wall to bottom profile, to easier withstand hydraulic hydrostatically induced pressure of contained liquid. Most container tanks for handling liquids during transportation are designed to handle varying degrees of pressure.

Since most liquids can spill, evaporate, or seep through even the smallest opening, special consideration must be made for their safe and

secure handling. This usually involves building a bundling, or containment dike, around the tank, so that any leakage may be safely contained.

Some storage tanks need a floating roof in addition to or in lieu of the fixed roof and structure. This floating roof rises and falls with the liquid level inside the tank, thereby decreasing the vapor space above the liquid level. Floating roofs are considered a safety requirement as well as a pollution prevention measure for many industries including petroleum refining.

Tanks for a particular fluid are chosen according to the flash-point of that substance. Generally in refineries and especially for liquid fuels, there are fixed roof tanks, and floating roof tanks.

1. Fixed roof tanks are meant for liquids with very high flash points, (e.g. fuel oil, water, bitumen etc.) Cone roofs, dome roofs and umbrella roofs are usual. These are insulated to prevent the clogging of certain materials, wherein the heat is provided by steam coils within the tanks. Dome roof tanks are meant for tanks having slightly higher storage pressure than that of atmosphere (e.g. slop oil).
2. Floating roof tanks are broadly divided into external floating roof tanks usually called floating roof tanks: FR Tanks) and internal floating roof types (IFR Tanks).

IFR tanks are used for liquids with low flash-points (e.g., ATF, MS. gasoline, ethanol). These tanks are nothing but cone roof tanks with a floating roof inside which travels up and down along with the liquid level. This floating roof traps the vapor from low flash-point fuels. Floating roofs are supported with legs or cables on which they rest. FR tanks do not have a fixed roof (it is open in the top) and has a floating roof only. Medium flash point liquids such as naphtha, kerosene, diesel, and crude oil are stored in these tanks.

One of the common types found in mining areas is the open roof type tank, usually to store ore slurries. These are the easiest storage tanks to build.

A general storage vessel is shown in the design section (Fig. 10.1).



Figure 10.1 – MDEA storage vessel

Materials

While steel and concrete remain one of the most popular choices for tanks, glass-reinforced plastic, thermoplastic and polyethylene tanks are increasing in popularity. They offer lower build costs and greater chemical resistance, especially for storage of specialty chemicals. There are several relevant standards, such as British Standard 4994 (1989), DVS (German Welding Institute 2205, and ASME (American Society of Mechanical Engineers) RTP-1 which give advice on wall thickness, quality control procedures, testing procedures, accreditation, fabrication and design criteria of final product.

Part III

Material and Energy Balance

Preheater

Material Balance

The data used is given in Table 11.1 and Table 11.2.

Table 11.1 – Sour Gas Data.

Process Variable	Value (SI)	Value (Industry)
Inlet Flow Velocity	615.74m ³ /s	53.2MMSCMD
Outlet Flow Velocity	609.68m ³ /s	52.7MMSCMD
Inlet Pressure	58.83bar	60kgf/cm ²
Outlet Pressure	58.83bar	60kgf/cm ²
Inlet Temperature	300.15K	27°C
Outlet Temperature	305.15K	32°C
Specific Heat Capacity	2100J/kg	2.1kJ/kg
Specific Gravity	0.717	-
Density	717kg/m ³	-
Amount Of Moisture Removed	6000kg/s	6m ³ /s

Table 11.2 – Steam Data.

Process Variable	Value (SI)	Value (Industry)
Inlet Temperature	533.15K	260°C
Outlet Temperature	433.15K	160°C
Specific Heat Capacity	1900J/kg	1.9kJ/kg
Inlet Mass Flow Rate	60000kg/s	-
Density	0.693kg/m ³	-
Moisture Present	-	0.01%

Shell Side

Amount of sour gas entering = Inlet Flow Velocity × Density

$$= 615.74 \times 717 = 441.48 \times 10^3 \text{ kg/s}$$

Amount of sour gas leaving = Amount of sour gas entering – Amount of moisture removed

$$= 615.74 \times 10^3 - 6 = 609.74 \times 10^3 \text{ m}^3/\text{s}$$

Mass Flow rate = Volumetric Flow Rate × Density

$$= 609.74 \times 717 = 437.18 \times 10^3 \text{ kg/s}$$

Tube Side

Amount of steam entering in tube side = Amount of steam leaving tube side = 60000kg/s

Energy Balance

Heat load of steam = $60000 * 1.9 * 130 = 14,820,000 \text{ J}$

Sensible heat of Sour gas = Mass Flowrate × Specific Heat Capacity $\times \Delta T$

$$= 441.48 \times 2.1 \times (32 - 27) = 4635.54 \text{ J}$$

Sensible heat of moisture = Mass Flowrate × Specific Heat Capacity $\times \Delta T$

$$= 6000 \times 4.2 \times (100 - 27) = 1839600 \text{ J}$$

Latent heat of vaporization of moisture = Mass Flowrate × Latent Heat of Water

$$= 6000 \times 2176 = 13056000 \text{ J}$$

Heat load from steam = Sensible heat of sour gas + Sensible heat of moisture + Latent heat of vaporization of moisture + heat losses

Heat losses = 80235J

The data used is given in Table 12.1

Table 12.1 – Absorber Variables

Process Variable	Value
Inlet flow rate of gas	437.18kg/s
Inlet concentration of hydrogen sulphide	100ppm
Outlet concentration of hydrogen sulphide	4ppm
Inlet flow rate of lean amine	0.13kg/s
Specific gravity of lean amine	1.0432
Specific heat of sour gas	2.1kJ/kg
Specific heat of amine	0.6kJ/kg
Inlet temperature of lean MDEA	42°C
Outlet temperature of rich MDEA	40°C
Inlet temperature of sour gas	32°C

Material Balance

$$\text{IN} - \text{OUT} = \text{ACCUMULATION} + \text{DISAPPEARANCE}$$

Since , ACCUMULATION = DISAPPEARANCE = 0 Therefore ,
 $\text{IN} = \text{OUT}$

$$\text{Inlet Flowrate of gas} \times \Delta \text{ Concentration} = 437.18 \times (100 - 4) \times 10^{-6}$$

$$= \text{Inlet Flowrate of Amine} \times \Delta \text{ Concentration}$$

$$= (X - 0)$$

$$\text{Hence , Outlet concentration of rich amine} = X = 0.32$$

Energy Balance

Sensible heat of amine = Mass Flowrate \times Specific Heat Capacity $\times \Delta T$

$$= 0.13 \times 0.6 \times (42 - 40) = 0.156\text{J}$$

Sensible heat of sour gas = Sensible heat of amine = 0.156J

Thus, Outlet temperature of sour gas = $32^\circ C$

The data used is given in Table 13.1

Table 13.1 – Stripper and Reboiler Variables

Process Variables	Values
Inlet flow rate of amine	0.13kg/s
Inlet temperature of rich MDEA	130°C
Outlet temperature of lean MDEA	100°C
Flow rate of steam in the reboiler	60000kg/s
Inlet concentration of hydrogen sulphide in the rich MDEA	0.32ppm
Specific Heat of Amine	0.6kJ/kg

Material balance

$$\text{IN} - \text{OUT} = \text{ACCUMULATION} + \text{DISAPPEARANCE}$$

Since , ACCUMULATION = DISAPPEARANCE = 0 Therefore ,
 $\text{IN} = \text{OUT}$

$$0.13 \times (0.32 - 0) = 60,000 \times (X - 0)$$

$$X = 0.7\text{ppm}$$

Energy Balance

Sensible heat of amine = Mass Flowrate \times Specific Heat Capacity $\times \Delta T$

$$= 0.13 \times 0.6 \times (130 - 100) = 2.34\text{J}$$

Sensible heat of steam = Sensible heat of amine = 2.34J

Thus, Outlet temperature of steam = 260°C

Material Balance

Table 14.1 – Overall Material Balance.

Variables	Values
Inlet capacity of sour gas	$615.74\text{m}^3/\text{s}$ (53.2MMSCMD)
Amount of hydrogen sulphide removed	$0.06\text{m}^3/\text{s}$
Amount of Moisture removed	$6\text{m}^3/\text{s}$
Outlet capacity of sweet gas	$615.74 - 0.06 - 6 = 609.68\text{m}^3/\text{s}$
Outlet capacity of sweet gas	52.67MMSCMD

Energy Balance

This section is equivalent to energy balance done in the preheater section.

Since, the other equipment in use are completely insulated, therefore the overall energy balance suggests energy loss = 80235J

Part IV

Equipment Design

We note again the basic layout of our plant in Fig. 15.1.

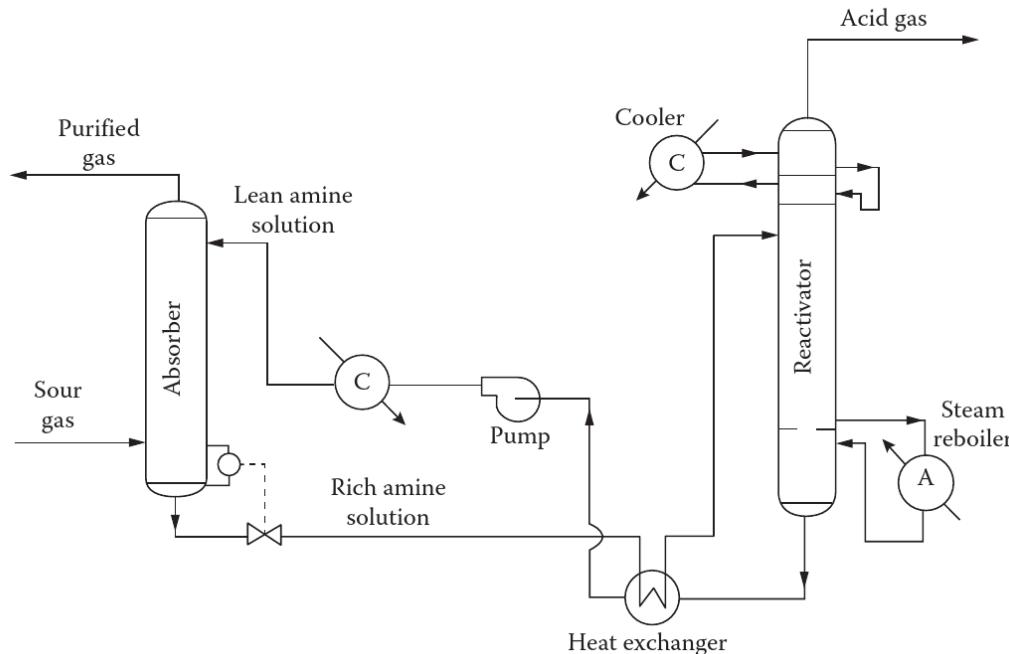


Figure 15.1 – Amine or olamine sweetening from Speight (2014)

More specifically we shall focus on the design parameters of the following:

- Tall Vertical Vessels
 - Absorber
 - Stripper
- Heat Exchangers
 - Shell and Tube Preheaters
 - Plate Heat Exchangers
- Valve Trays
- Centrifugal Pumps

In this section we will determine relevant parameters for the appropriate sizing in mechanical terms of the equipment. We shall not here note

the costs or even the process optimization. The materials determined here are mainly by introspection and for longevity.

For each equipment considered in this section we shall extend the theory of associations and correlations developed before in the literature survey of section II by demonstrating calculations for the same.

Mechanical Design of GSU Absorption Tower

Objective

We will develop here a an estimate of the thickness of the absorption tower used in the Gas Sweetening Unit based on the theory developed in the previous chapter.

Calculations

Table 16.1 – GSU Tower Data.

Variable	Value
Column Diameter	2.5m
Column Height	20m
Valve Trays	14
Glass Wool Insulation thickness	70mm
Operating Pressure	60bar
Joint Efficiency Factor	1
Skirt Support Height	2.5m
Column Material	Stainless Steel
Material Operating Stress	130bar

The data is summarized in Table 16.1 and the material operating stress is the maximum operating stress under the column conditions.

We will follow an iterative process:

1. Design pressure
2. Minimum thickness
3. Preliminary division of trays
4. Weight estimate of trays
5. Stress estimate and analysis (3. if incorrect)

Design Pressure

As we know the design pressure is estimated by Eq. 16.1.

$$P_d = 1.1 \times P_o \quad (16.1)$$

Where:

- P_d is the design pressure
- P_o is the operating pressure

Hence we have:

$$P_d = 1.1 * 60 = 66\text{bar}$$

Minimum Thickness

Our minimum thickness is determined from the column diameter, joint efficiency, and maximum allowable stress as per Eq. 16.2.

$$t_{min} = \frac{J}{2} \frac{D_c}{P_{max}} \quad (16.2)$$

Where:

- t_{min} is the minimum thickness
- J is the joint efficiency factor
- P_{max} is the maximum material operating stress
- D_c is the column diameter

Hence we have:

$$t_{min} = \frac{1}{2} \times \frac{2500}{130} = 9.61\text{mm}$$

Preliminary Tray Division

For the sake of simplicity we shall first try 5 parts for the column as depicted in Table 16.2.

Table 16.2 – GSU Plate Approximation.

Plate Number	Plate Thickness (mm)
1	10
2	12
3	14
4	16
5	18

We note that the factor C_y for plate towers may be taken to be 1.15 and that the 18mm plate is at the bottom of the absorber.

Weight Estimation

$$\text{Weight of plates} = 1.2 \times 4.9 = 5.9\text{kN}$$

Since , there are 14 valve trays ,therefore

$$\text{Total weight of plates in the column} = 82.6\text{kN}$$

Hence,

$$\text{Wall density of glass wool} = 130\text{kg/m}^3$$

$$\text{Volume of insulation} = \pi \times 2 \times 20 \times 0.07 = 8.79\text{m}^2$$

$$\text{Weight of insulation} = 11.21\text{kN}$$

$$\text{Total Weight} \$ = 213 + 82.6 + 11.21 \$$$

$$= 306.81\text{kN}$$

Stress estimates and analysis

Now, calculating the stresses due to wind loading,

Dynamic wind pressure = 1280Pa

Mean Diameter of the tower = $2.5 + (14 + 70) \times 0.002 = 2.668\text{m}$

Wind loading = $1280 \times 2.668 = 3415\text{Pa}$

Bending moment = $\frac{3415}{2} \times 20^2 = 683000\text{Nm}$

Checking our calculations and assumptions using stress analysis

For 18mm plates,

$$F_l = \frac{1 \times 2500}{4 \times 18} = 34.72\text{Pa}$$

$$F_h = \frac{1 \times 2500}{2 \times 18} = 69.44\text{Pa}$$

$$\text{Dead Weight} = \frac{306.81 \times 10^3}{\pi \times 2518 \times 18} = 6.76\text{Pa}$$

Bending length of plate = $2500 + 2 \times 18 = 2536\text{mm}$

$$I_v = \frac{\pi}{64 \times (25364 - 25004)} = 1.128 \times 101\text{mm}^4$$

$$F_b = 683000 \times 10^3 \times \frac{(2500/2+18)}{1.128 \times 10^{11}} = 69.06\text{Pa}$$

Resulting Longitudinal Stress:

Stress based on upward wind flow = $34.72 - 6.76 + 69.06 = 97.02\text{Pa}$

Stress based on downward wind flow = $69.06 + 6.76 - 34.72 = 41.1\text{Pa}$

Difference in principle stress = $97.02 - (-41.1) = 138.73\text{Pa}$

$$\text{Buckling stress} = 2 \times 10^4 \times \frac{18}{2536} = 141.95\text{Pa}$$

Maximum compressive stress without internal pressure = $6.76 + 69.06 = 77.83\text{Pa}$

Mechanical Design of Tray Column

Objective

To find out the height of the column and the thickness of torispherical head as well as the thickness of the trays.

From the previous calculations done with data in Table 17.1 we have summarized the results in Table 17.1.

Table 17.1 – Previously Calculated Data.

Variable	Value
Shell Internal Diameter	2464mm
Valve Trays	14
Tray Spacing	600mm
Hole Diameter	12mm
Plate Thickness	18mm
Weir Height	40mm
Top Disengaging Space	600mm
Bottom Separator Space	1000mm
Allowable Stress	130bar
Column and Tray Material	Stainless Steel (SS340L)

Our process for the design of the column will be:

1. Column height
2. Head calculation

Calculations

Column Height

We obtain the column height by use of Eq. 17.1.

$$h_{column} = t \times (1 + 13T + 14h) + 1000 = 19.2\text{m} \quad (17.1)$$

Where:

- h_{column} is the column height
- t is the thickness

Thus, the height of the column of 20m is justifiable .

Torispherical Head

As per the theoretical considerations discussed previously, our torispherical head will relevant dimensions as given by Eq. 17.2.

$$\text{Thickness} = \frac{pd}{(2 \times f \times J - 0.2 \times p)} = 7.81\text{mm} \quad (17.2)$$

Where:

- t_{head} is the head thickness
- J is the joint efficiency factor

This gives us

$$t_{head} = 7.81\text{mm}$$

Further we note that:

$$\text{Length of the bottom plate} = \frac{20}{5} = 4\text{mm}$$

$$\text{Bottom separator space} = 1\text{mm}$$

$$\text{Thickness of the bottom tray} = 18\text{mm}$$

Result

The results are summarized in Table 17.2

Table 17.2 – GSU Tower Data.

Parameter	Value
Thickness of ellipsoidal head	8mm
Thickness of the bottom tray	18mm
Height of the column	20m

Mechanical Design of Reboiler

18

Objective

To calculate the number of tubes in the reboiler.

Data

Variables	Values
Velocity of the MDEA in the tube side (Data from Flow Measuring Device place after the Flow control Valve)	1m/s
Density of MDEA	1.024g/cc
Flow rate of MDEA	0.2kg/s

Calculations

Applying mass balance to calculate the number of tubes:

$$nPAv = 2$$

$$\frac{n \times 1.024 \times 103 \times \pi \times (0.002)^2 \times 1}{4} = 0.2$$

$$n = 64$$

Result

Number of tubes in the kettle reboiler = 64

Material of construction = SS340L

Thickness of shell (from previous calculations) = 18mm # Mechanical Hand Analysis of Plate Heat Exchanger ## Objective

To find the number of plates in one stack of a PHE

Data:

The data used is given in Table 18.1

Table 18.1 – Plate Heat Exchanger Data.

Variables	Values
Specific gravity of MDEA	1.024
Gap between the plates	1.5mm
Thickness of the plate	2mm
Dimensions of the plate	15mm × 30mm
Viscosity of MDEA	38cP
Thermal conductivity of carbon steel plate	45W/K
Inlet temperature of rich MDEA	42°C
Outlet temperature of rich MDEA	130°C
Inlet temperature of lean MDEA	100°C
Outlet temperature of rich MDEA	40°C
Specific Heat Capacity	600J/kg

Calculations:

Assume:

- $U = 100\text{W/m}^2\text{K}$
- $\text{LMTD} = 42^\circ\text{C}$

$$\text{Heat load} = \text{Flowrate} \times \text{Specific Heat Capacity} \times \Delta T$$

$$Q = 0.13 \times 600 \times 88 = 6864\text{W}$$

$$A = \frac{Q}{U} \times LMTD = 0.194\text{m}^2$$

$$\text{Number of plates in a stack} = \frac{A}{LB} = 21$$

Applying mass balance:

$$p \times A \times v = \frac{m}{n}$$

$$v = 0.027 \text{ m/s}$$

Since laminar flow is maintained to prevent leakage of toxic MDEA

$$Re = \frac{pvL}{u} = 106.5$$

$$h = 1.86 \times \frac{k}{l} = 558 \text{ W/m}^2\text{K}$$

Assuming 0.3mm of fouling resistance on one plate

Total Fouling resistance = $0.3 \times 21 = 6.3 \text{ mm}$

Using: $U_{o,cal} = \frac{1}{\frac{1}{h_0} + R_{do} + \frac{A_o}{A_i}(\frac{d_o - d_i}{2k_w}) + \frac{A_o}{A_i}(\frac{1}{h_i}) + \frac{A_o}{A_i}(R_{di})}$ we calculate the new value of $U_{o,cal}$.

$$U_{o,cal} = 101.66 \text{ W/m}^2\text{K}$$

Error = $1.66\% < 30\%$

This implies our assumption was correct

Result

Number of plates in one stack = 21

Number of stacks = 15 (given)

Number of plates = 315

Comment

This is in close approximation to total 350 plates used in PHE. # Mechanical Design of Centrifugal Pump ## Objective

To calculate the minimum power of the pump to be installed for pumping MDEA to absorber from MDEA storage tank.

Data

The data used is given in Table 18.2

Table 18.2 – Centrifugal Pump Data.

Variable	Values
Height of the absorber	20m
Height of the skirt support	2.5m
Density of MDEA	\$ 1.024 g/cc\$

Calculations

Total head required for pumping = $20 + 2.5 = 22.5\text{m}$

Loss coefficient , based on the surface roughness of the carbon steel
= 0.7

Neglecting the effects of bends we have:

$$\text{Total head loss} = \frac{kv^2}{2g}$$

Calculating velocity from the Bernoulli's equation we have:

$$V = (2gh)^{0.5} = 21\text{m/s}$$

$$\text{Total head loss} = 15.75\text{m}$$

$$\text{Total head required in the pump} = 22.5 + 15.75 = 38.25\text{m}$$

$$\text{Power of the pump} = pQHg = 150\text{kW}$$

Result

$$\text{Power of the pump} = 150\text{kW}$$

Comments

Result obtained is in close accordance with the 200hP = 201.6kW pump used in the plant. # Mechanical Design of Storage Vessel ## Objective To calculate the thickness of the storage vessel ## Data The data used is given in Table 18.3

Table 18.3 – Storage Vessel Data.

Variables	Values
Internal pressure	\$ 1 atm\$
Diameter of the vessel	6m
Maximum permissible stress of the stainless steel vessel	130bar
Joint efficiency factor	1
Height of the storage vessel	5m

Calculations

Applying the formula for thin pressure vessel under internal pressure ,

$$\text{Thickness}(t) = \frac{\text{Pressure}(P) \times \text{Diameter}(D)}{2 \text{jointefficiencyfactor}(f_j)} = 24\text{mm}$$

Result

The thickness of the bottom plate = 24mm

Comments

There is no roof provide for the storage vessel to cut the cost. Blanketing is done by flue gases.

The thickness of the plate remains constant throughout at ONGC, Surat.

Part V

Hazards and Safety

Material Safety and Toxicology

Before considering the safety of a plant we would do well to discuss briefly the nature of toxicity in general. A colloquial definition, which captures much of the nuances of the difficulties associated with attempts is succinctly paraphrased as:

Every substance is deadly, it's only the amount required which varies.

Naturally there are more quantitative definitions, especially for the chemical process industries. Commonly cited ones are (Crowl and Louvar 2002):

Toxicology The toxicity of a chemical or physical agent is a property of the agent describing its effect on biological organisms. This is a substance property.

Toxic Hazard This is the likelihood of damage to biological organisms based on exposure resulting from transport and other physical factors of usage. This may be reduced by usage of industrial hygiene practices.

Global Standards

The gold standard of chemicals and affiliated safety issues is undoubtedly the phenomenal guidance document (Nationen 2017) published by the United Nations and updated every two years. This document, popularly known as the UN GHS Purple Book is the epitome of global expertise and research and is currently in its seventh iteration.

Overview

Very broadly, has the following salient features:

- Defines physical, health and environmental hazards of chemicals and harmonizes classification criteria;

- Standardizes the content and format of chemical labels and Safety Data Sheets.

The entire GHS range of hazards may be gleaned by inspection of Fig. 19.1.

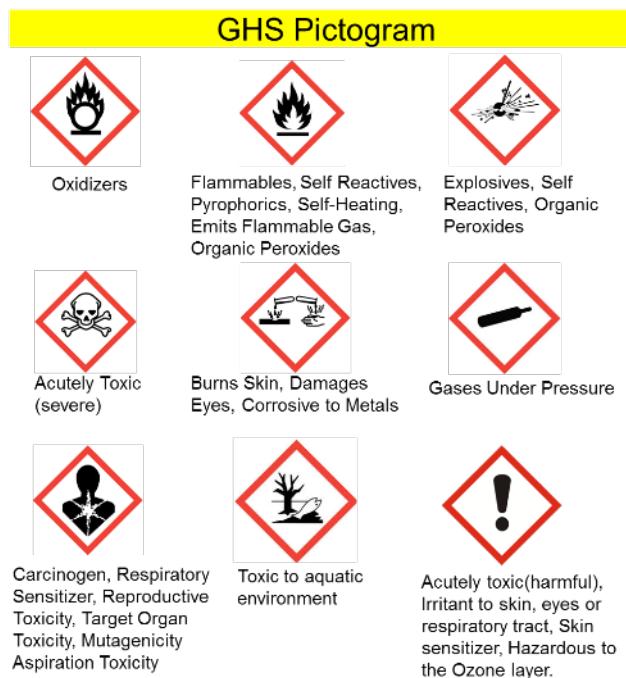


Figure 19.1 – GHS Pictogram

Safety Data Sheets

A Safety Data Sheet (SDS), also called a Material Safety Data Sheet (MSDS) in the pre-GHS era, is a very important document which is meant to list major hazards of an industrial substance or mixture and to provide advice on safety precautions.

An SDS should be prepared and provided for a substance or mixture meeting classification criteria for a hazard class or for a mixture containing an ingredient meeting criteria for carcinogenic, toxic to reproduction or specific target organ toxicity exceeding certain defined cut-off value limits.

The latest harmonized scheme for the chemical properties under consideration are divided into 16 sections:

- Section 1 Identification of the substance or mixture and of the supplier;
- Section 2 Hazard identification;
- Section 3 Composition/information on ingredients;
- Section 4 First-aid measures;
- Section 5 Fire-fighting measures;
- Section 6 Accidental release measures;
- Section 7 Handling and storage;
- Section 8 Exposure controls/personal protection;
- Section 9 Physical and chemical properties;
- Section 10 Stability and reactivity;
- Section 11 Toxicological information;
- Section 12 Ecological information;
- Section 13 Disposal consideration;
- Section 14 Transport information;
- Section 15 Regulatory information;
- Section 16 Other information.

Caveats and legality

Like all UN guidelines and documents the ones listed in the Purple book are also not legally binding in any sense of the term, except in terms of a compliance in moral terms of a government towards its citizens.

As such there are many countries which adopt the guidelines and format specifications as stipulated by the Globally Harmonized System of Classification and Labelling of Chemicals but there are as many which adopt only a subset of a superset of the same as per the needs of the country in question. This modular adoption policy is often called the GHS Building Blocks approach.

Indian Perspective

In India, the government has in the past been very proactive in terms of both industrial growth and also the safety and conservation measures, having established by Act of Parliament, both the Bureau of Energy Efficiency and adopting a form of the UN specifications that same year. As the last amendment to the Manufacture, Storage and Import of Hazardous Chemicals Rule, 1989 was in 2000, India therefore lists 9 mandatory sections for the MSDS as described in the Schedule to the aforementioned act on hazardous chemicals.

Material Safety Data Sheets

- Material Safety Data Sheets (MSDSs) are documents prepared by the manufacturers/ suppliers of the chemicals and contain information on physical and chemical properties of the material, potential hazards of the material and how to work safely with these materials. They also contain information on usages, storage, handling and emergency procedures related to the hazards of the materials. In fact, they provide a single reference for all information about hazardous substances.
- MSDSs provide information on chemicals, chemical compounds and chemical mixtures and are primarily intended for use by the personnel working with hazardous chemical substances.
- MSDSs need to be updated every three years, but if any significant new information becomes available about the material, then it should be incorporated into MSDS even before lapse of three years.

Indian MSDS Sections

As mandated by the Manufacture, Storage and Import of Hazardous Chemicals Rule, 1989 (amended in 2000) the following sections must be present for a document to be considered to be a material safety data sheet.

- Section 1: Chemical Identity

- Section 2: Physical and Chemical Data
- Section 3: Fire and Explosion Hazard Data
- Section 4: Reactivity Data
- Section 5: Health Hazard Data
- Section 6: Preventive Measures
- Section 7: Emergency and First Aid Measure
- Section 8: Additional Information/ References
- Section 9: Manufacturer/ Supplier Data

Based on a typical MSDS and data made available via (“Material Safety Data Sheets and Their Relevance to Customs Work,” n.d.) we infer the following about these sections.

Section I: Chemical Product and Company Identification This section contains product identification, name and address of the supplier of the MSDS and emergency telephone numbers.

Section II: Composition and Information on Ingredients This section contains percentage composition by weight of the ingredients and toxicological data on the ingredients of the given chemical.

Section III: Hazard Identification This section contains the information on toxicology of the chemical and its acute and chronic health effects.

Section IV: First Aid Measures This section provides for exposure routes, signs and symptoms of exposure, and emergency and first aid procedures.

Section V: Fire and Explosion Data This section contains information about the flammability, stability and the reactivity of the substance.

Section VI: Accidental Release Measures This section provides for measures to be adopted during accidental release of the chemical due to spills, leaks, etc.

Section VII: Handling and Storage This section contains precautions for safe handling, storage conditions and reactivity with other materials.

Section VIII: Exposure Controls or Personal Protection This section provides details about the types of Personal Protective Equipment (PPE) required by the person working with these

chemicals, hygiene practices and ventilation requirements to reduce potential for exposure.

Section IX: Physical and Chemical Properties This section contains the physical and chemical properties of the chemical.

MDEA Toxicity

Much of the physical and chemical properties have been elaborated in the solvent selection of the previous sections however we will here note the unique nature of MDEA in the Indian legal and industrial system.

- MDEA is not among the 800 hazardous chemicals listed in the Schedule to the Manufacture, Storage and Import of Hazardous Chemicals Rule of 1989 (amended in 2000).
- Inspite of this MDEA may not be sold or manufactured without the dissemination of a material safety data sheet.

This apparent paradox is resolved by a careful examination of the Schedules to the relevant Act of Parliament, where it is mentioned in Schedule - IV that the substances involved in the desulphurization, manufacture and transformation of sulphur containing compounds as well as compounds part of installations for distillation, refining or other processing of petroleum or petroleum products are to be regulated by the rules mentioned in the aforementioned Act.

We therefore note that in the quantities used by the industry, MDEA is indeed a chemical with hazardous potential and thus its MSDS has been listed in the Appendix. We also note the hazardous quantities of the sulphur in Fig. 19.2 which is removed by the action of gas sweetening, which warrants such special treatment of the mild MDEA.

CONCENTRATIONS IN AIR				
Percent by Volume	Parts per Million by Volume	Grains Per 100 Standard Cubic Feet ^a	Milligrams Per Cubic Meter ^a	Physiological Effects
0.00013	0.13	0.008	0.18	Obvious and unpleasant odor generally perceptible at 0.13 ppm and quite noticeable at 4.6 ppm. As the concentration increases, the sense of smell fatigues and the gas can no longer be detected by odor.
0.002	10	1.26	28.83	Acceptable ceiling concentration permitted by federal OSHA standards.
0.005	50	3.15	72.07	Acceptable maximum peak above the OSHA acceptable ceiling concentrations permitted once for 10 min per 8-h shift, if no other measurable exposure occurs.
0.01	100	6.30	144.14	Coughing, eye irritation, loss of sense of smell after 3–15 min. Altered respiration, pain in eyes, and drowsiness after 15–30 min, followed by throat irritation after 1 h. Prolonged exposure results in a gradual increase in the severity of these symptoms.
0.02	200	12.59	288.06	Kills sense of smell rapidly, burns eyes and throat.
0.05	500	31.49	720.49	Dizziness, loss of sense of reasoning and balance. Breathing problems in a few minutes. Victims need prompt artificial resuscitation.
0.07	700	44.08	1008.55	Unconscious quickly. Breathing will stop and deaths will result if not rescued promptly. Artificial resuscitation is needed.
0.10+	1000+	62.98	1440.98+	Unconsciousness at once. Permanent brain damage or death may result unless rescued promptly and given artificial resuscitation.

^aBased on 1% hydrogen sulfide = 629.77 gr/100 SCF at 14.696 psia and 59 °F, or 101.325 kPa and 15 °C.

Figure 19.2 – Toxicity data

Introduction

Hazard and operability studies (HAZOP) is defined by Perry, Green, and Maloney (1997) to be:

HAZOP A set of formal hazard identification and elimination procedures designed to identify hazards to people, process plants, and the environment.

Along with failure mode and effect analysis (FMEA), and other plant safety studies showcase the combined efforts of many industry and academic collaborations. The process hazards analysis recommends appropriate measures to reduce the risk, including (but not limited to) the safety interlocks to be implemented in the safety interlock system.

We formalize here the definition and therefore the relevant differences between a **hazard** and a **risk** as sometimes in common parlance they are used interchangeably.

Hazards The inherent potential of a material or activity to harm people, property, or the environment. Hazards do not have a probability component.

Risks These are a measure of economic loss or injury in terms of both the incident likelihood and magnitude of loss of injury. Risks imply the probability of an incident occurring. As shown in Fig. 20.1

Fault Trees

Due to the complexities in the form of many variables in the analysis of a typical plant or even a sub portion of the plant, there exist computational fault tree generators which are used to somewhat automate the HAZOP studies.

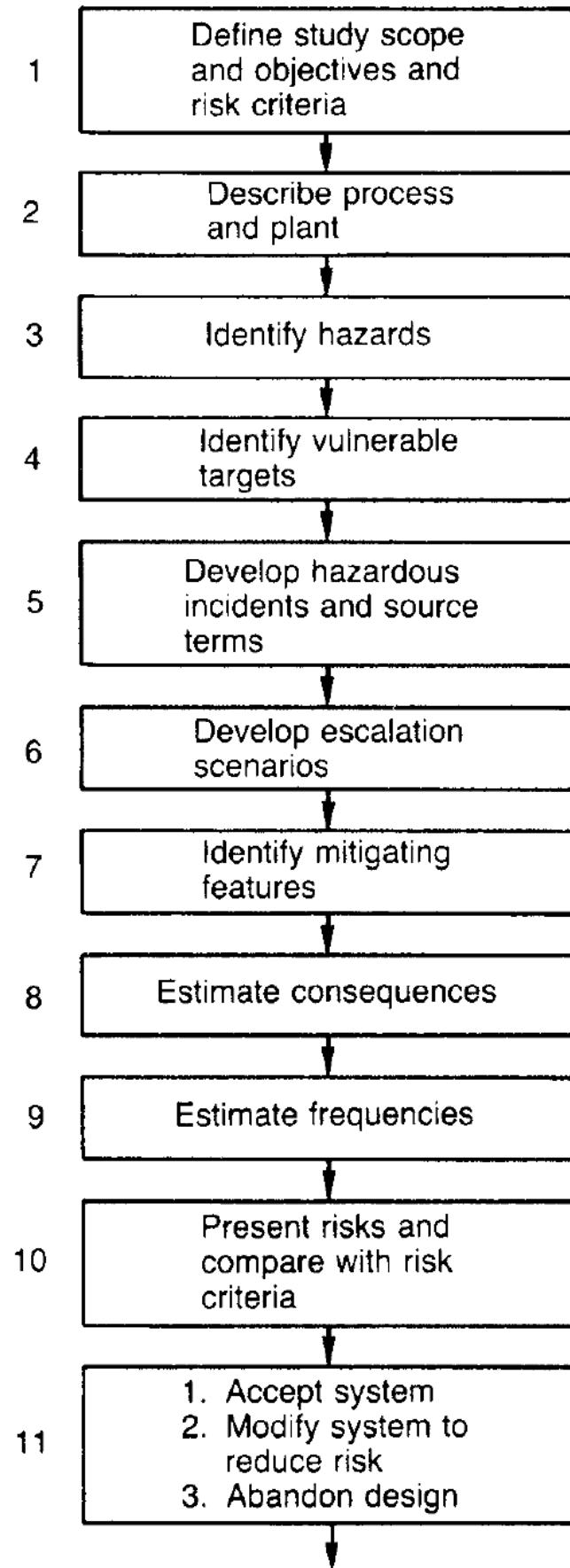


Figure 20.1 – Risk Management

These computer generated studies should only be used as a guideline as they are based on simple boolean logic and probability which is no substitute for an experienced process engineer and HAZOP analyst.

Boolean Operator Symbols

The logical symbols are enlisted in Fig. 20.2, Fig. 20.3, Fig. 20.4 as mentioned in Lees (2012).

A Events	Symbol
	Primary, or base, event – basic fault event requiring no further development
	Undeveloped, or diamond, event – fault event which has not been further developed
	Intermediate event – fault event which occurs due to antecedent causes acting through a logic gate
	Conditioning event – specific condition which applies to a logic gate (used mainly with PRIORITY, AND and INHIBIT gates)
	External, or house, event – event which is normally expected to occur ^a

Figure 20.2 – Boolean Event Symbols

Plant HAZOP Preliminaries

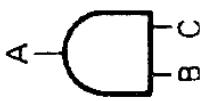
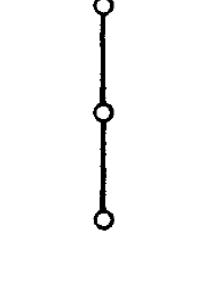
Hazard warning flows for major system master tree and subsidiary trees are shown in Fig. 20.5 and Fig. 20.6.

B Logic gates, etc.

Symbol	Alternative symbol
	AND gate – output exists only if all inputs exist
	OR gate – output exists if one or more inputs exists
	INHIBIT gate – output exists if input occurs in presence of the specific enabling condition (specified by conditioning event to right of gate)
	PRIORITY AND gate – output exists if all inputs occur in a specific sequence (specified by conditioning event to right of gate)
	EXCLUSIVE OR gate – output exists if one, and only one, input exists
	VOTING gate – output exists if there exist r-out-of-n inputs ^b
	TRANSFER IN – symbol indicating that the tree is developed further at the corresponding TRANSFER OUT symbol
	TRANSFER OUT – symbol indicating that the portion of the tree below the symbol is to be attached to the main tree at the corresponding TRANSFER IN symbol

^a This is the definition given by Vesely *et al.* (1981). Other authors such as Henley and Kumamoto (1981) use this symbol for an event which is expected to occur or not to occur.

Figure 20.3 – Boolean Logic Gates

A Basic probability relations ^a			
Logic symbol	Reliability graph	Boolean algebra relation	Probability relations
		$A = BC$	$P(A) = P(B)P(C)$
		$A = B + C$	$P(A) = P(B) + P(C) - P(B)P(C)$
B Relations involving frequencies and/or probabilities ^a			
Gate	Inputs	Outputs	
OR	$P_B \text{ OR } P_C$ $F_B \text{ OR } F_C$ $F_B \text{ OR } P_C$	$P_A = P_B + P_C - P_B P_C \approx P_B + P_C$ $F_A = F_B + F_C$ Not permitted	
AND	$P_B \text{ AND } P_C$ $F_B \text{ AND } F_C$ $F_B \text{ AND } P_C$	$P_A = p_B p_C$ Not permitted; reformulate $F_A = F_B P_C$	

^a F, frequency; P, probability.

Figure 20.4 – Probability Relations

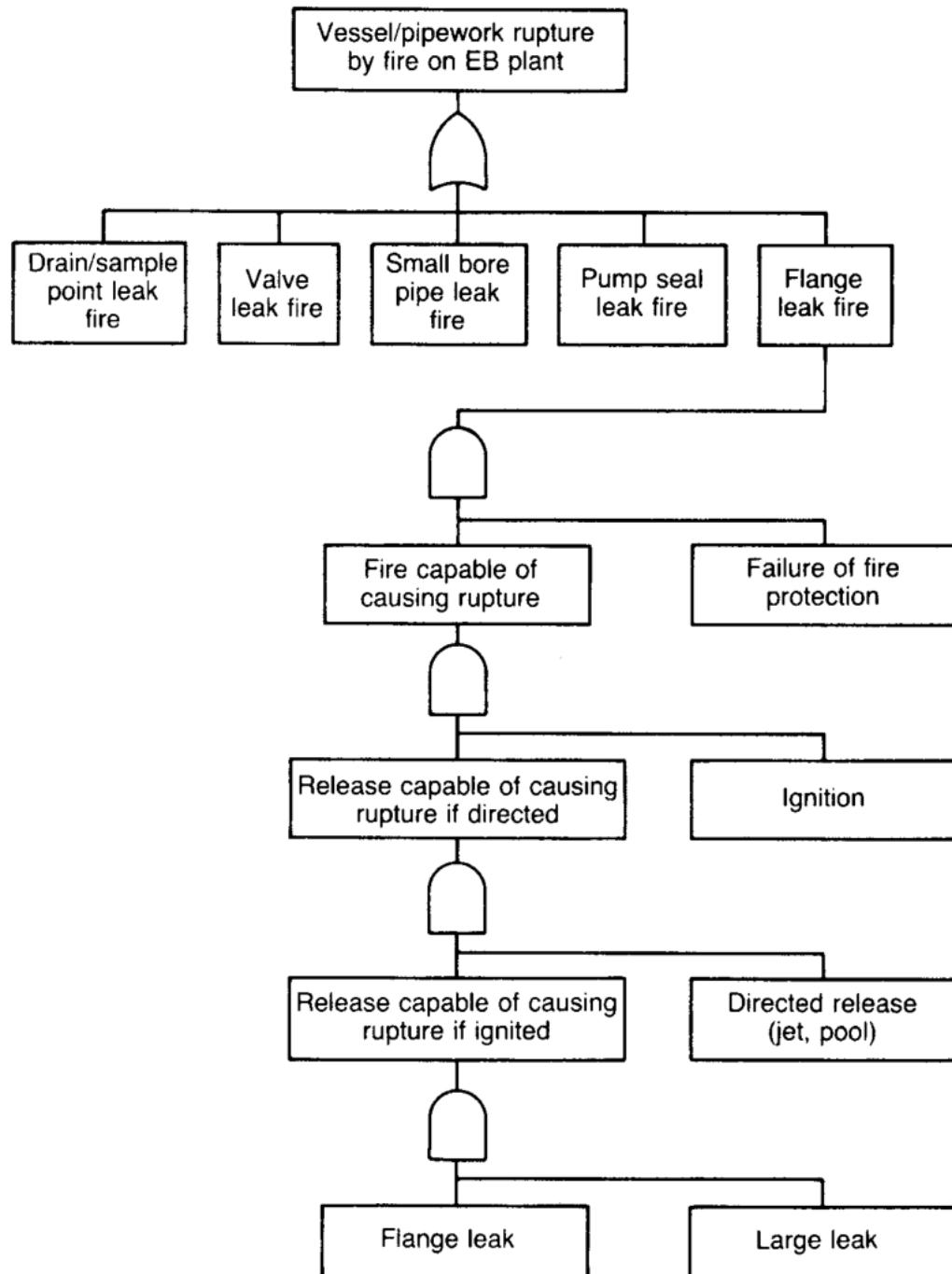


Figure 20.5 – Failure via mechanical rupture

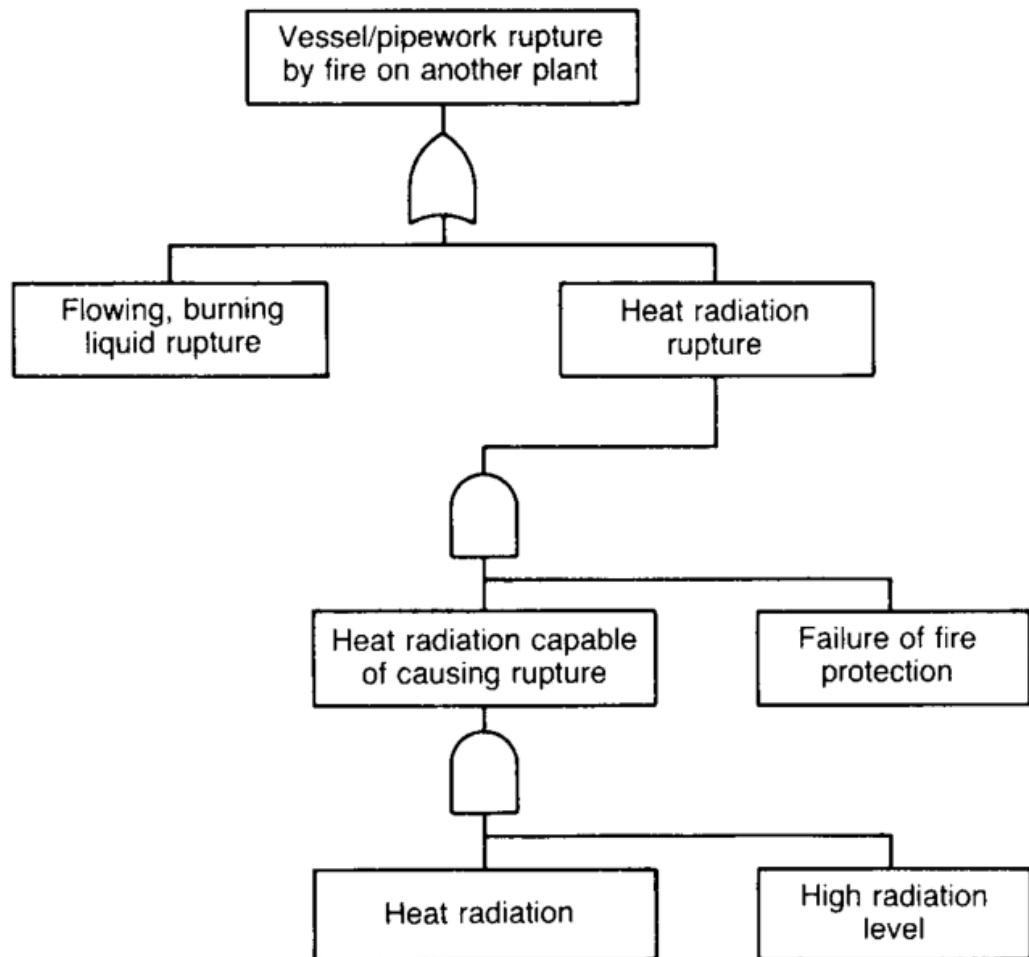


Figure 20.6 – Failure via temperature rupture

We shall discuss the control of the process shown in Fig. 15.1 and Fig. 21.1.

Standard Equipment

Each amine gas sweetening plant has at the very least, the following:

- amine contactor tower
- reboiler or surge tank
- reflux condenser or solution cooler in a common aerial cooler package
- reflux accumulator
- reflux pump or pumps
- main solvent pump or pumps
- solvent booster pump or pumps
- full-flow filter for rich solvent
- charcoal filter
- control panel or motor starters

The solvents which may be handled by the generalized flowsheets and equipment noted previously are:

- Monoethanolamine (MEA)
- Diethanolamine (DEA)
- Methyl-diethanolamine (MDEA)
- Diglycolamine (DGA)
- Sulfinol solvent

Standard Process

The steps followed in ammonia sweetening systems across the world have been largely standardized as a testament to the maturity and utility of this process.

These standard process steps are:

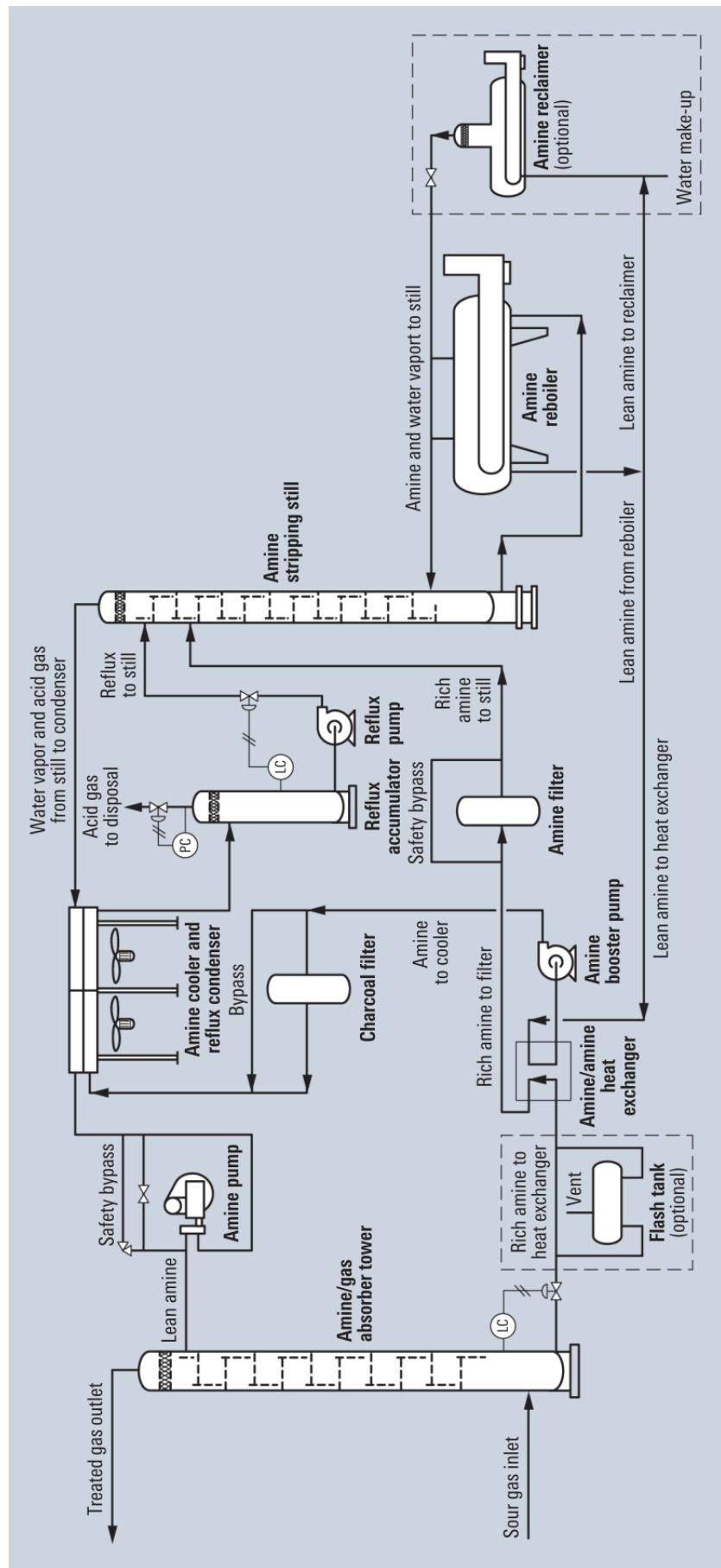


Figure 21.1 – Sweetening gas industrial layout

1. Sour gas enters the contactor tower and rises through the descending amine solution.
2. Purified gas flows from the top of the tower.
3. The amine solution, carrying absorbed acid gases, leaves the tower for the heat exchanger or optional flash tank.
4. Rich amine is heated by hot regenerated lean amine in the heat exchanger.
5. Rich amine is further heated in the regeneration still column, by heat supplied from the reboiler. The steam rising through the still liberates water and CO₂, regenerating the amine.
6. Steam and acid gases separated from the rich amine are condensed and cooled, respectively, in the reflux condenser.
7. Condensed steam is separated in the reflux accumulator and returned to the still. Acid gases may be vented, incinerated, or directed to a sulfur recovery system.
8. Hot regenerated lean amine is cooled in a solvent aerial cooler and circulated to the contactor tower, completing the cycle.
9. A variety of heat sources can be used for the still reboiler—direct fired, waste heat, hot oil, and

Instrumentation and Process Control

For the purpose of our discussion on the control of this fascinating process we shall refer to the process flow sheet given in Fig. 22.1.

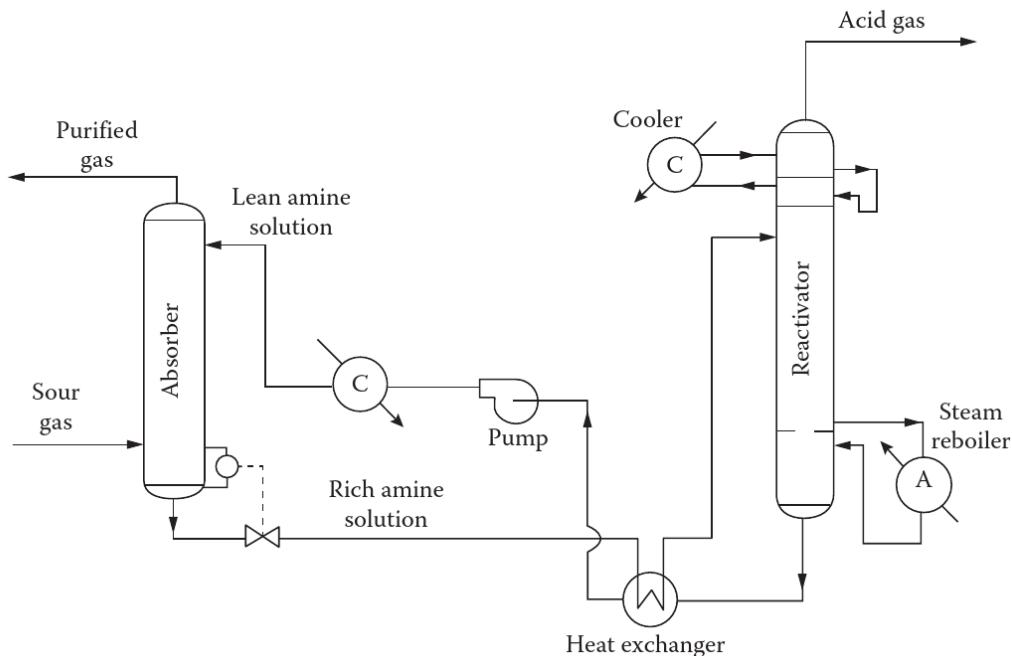


Figure 22.1 – Amine or olamine sweetening from Speight (2014)

Absorber

Differential Pressure

- Normal Range: 0.1 - 0.2 psi/tray
- Instrument: Continuous recorder w/alarm. *Note:* By locating the lower d/p cell tap at roughly the same elevation as the top sight glass connection, it doubles as an independent high-level alarm.
- Status check: At alarm or sign of upset
- Deviation: Consistently low
 - Cause: Possible tray damage
 - Consequence: Poor efficiency, off-spec treated gas
 - Action: Mechanical repair
- Deviation: Gradual increase
 - Cause: Possible fouling
 - Consequence: Flooding, poor efficiency, off-spec treated gas, capacity limit
 - Action: Cleanout, identify root cause (e.g. corrosion, contamination)
- Deviation: Sudden increase, erratic action
 - Cause: Foaming, flooding
 - Consequence: Poor efficiency, liquid carryover, general plant upset
 - Action: Add antifoam (discriminately), reduce gas and/or liquid rates, check relative gas/amine temperatures to determine likelihood of HC condensation, check feed gas for entrained HC's

Feed Gas Flow Rate

- Normal range: Site specific
- Instrument: Continuous recorder
- Status check: At sign of upset or change in process conditions
- Deviation: Low
 - Cause: Upstream process change
 - Consequence: Reduced amine demand, potential reduction in mass transfer due to weeping
 - Action: Reduce amine flow, supplement feed gas with recycle or clean gas if warranted
- Deviation: High
 - Cause: Upstream process change
 - Consequence: Increased amine demand, possible jet flooding
 - Action: Increase amine flow

Feed Gas Temperature

- Normal range: 80-120°F
- Instrument: Indicator
- Status check: Once/round or at sign of upset
- Deviation:
 - Low
 - Cause: Change in upstream process and/or ambient conditions
 - Consequence: Reduced acid gas recovery in extreme cases
 - Action: Increase temperature of feed gas and/or amine
 - High
 - Cause: Change in upstream process and/or ambient conditions
 - Consequence: Potentially reduced acid gas recovery
 - Action: Decrease feed gas temperature, or increase amine flow rate to improve heat balance

Lean Amine Temperature

- Normal range: 90-130°F
- Instrument: Indicator
- Status check: Once/round or at sign of upset
- Deviation:
 - Low
 - Cause: Change in upstream process and/or ambient condition
 - Consequence: Potentially reduced acid gas removal from high viscosity or low rate of reaction
 - Action: Reduce lean amine cooling or supply heat
 - High
 - Cause: Change in upstream process and/or ambient conditions
 - Consequence: Potentially reduced acid gas removal due to poor equilibrium at high absorber temperatures
Excessive moisture in treated gas, with potential downstream condensation and resultant corrosion/fouling
 - Action: Increase lean amine cooling

Lean Amine / Feed Gas Temperature Differential

- Normal range: Lean amine at least 10°F hotter than feed gas, where gas is HC-saturated
- Instrument: Indicators on individual streams
- Status check: Once/round or at sign of upset
- Deviation: Low
 - Cause: Change in upstream process and/or ambient conditions
 - Consequence: Condensation of HCs, potentially resulting in foaming and/or emulsification
 - Action: Increase lean amine temperature

Rich Amine Loading

- Normal range: Maximum within target (see Appendix for typical rich amine loadings), contingent on satisfactory gas cleanup
- Analysis: Spot lab sample
 - Caution: Samples exposed to air or allowed to weather may not yield accurate results.
Relative H₂S loss is significantly increased by high contactor pressures, LPG entrainment and excessive (> 0.5) mole loadings.
- Hazard: Potential H₂S release
- Status Check: Site-specific, ranging from once/shift to weekly
- Deviation: Low
 - Cause: Overcirculation
 - Consequence: Excessive utility consumption
 - Action: Reduce amine circulation rate
- Deviation: High
 - Cause: Undercirculation
 - Consequence: Reduced acid gas removal, corrosion
 - Action: Increase amine circulation rate

H₂S in Treated Gas

- Normal range: 10-50 ppm
- Instrument: Optional online analyzer
- Analysis: Length-of-stain tube
- Status check: Once/round or at sign of upset
- Deviation: High
 - Cause: One or more of following: flooding, foaming, inadequate stripping, amine rate, temperature or quality
 - Consequence: Contractual/environmental non-compliance
 - Action: Adjust process conditions as warranted

Reboiler Control

Reboiler Temperatures (In/Out)

- Normal Range: 230-260° F (dependent upon tower pressure, heat stable salt content, tray loadings, amine type and concentration)
- Instrument: Field, temperature gages; board, thermocouple/transmitter/recorder/alarm
- Status Check: Field, once/(round), board, at alarm or upset
- Deviation:
 - Low
 - Cause: Loss of reboiler heating media, fouling of reboiler or lean/rich exchangers due to excessive corrosion rates, failure of reflux/pumparound control leading to overcooling, loss of containment
 - Consequence: High lean loadings leading to off-spec product, emission of toxic gases to atmosphere
 - Action: Confirm initial and final state and flows of heating media and calculate duty to exchanger.
Compute tower heat balance.
Compute required heat requirement and compare to available heat.
Compute reboiler and lean/rich exchanger heat transfer coefficients to determine fouling. Clean exchangers if fouled.
Isolate leak.
Remove heat stable salts and anions.
Remove amine degradation products (reclaim)
 - High
 - Cause: Tower overhead or acid gas product line plugged or pressure control problem, too hot heating media, too high heat input, hydrocarbon incursion causing higher pressures, high heat stable salt anion and sodium content, improper circulation of amine through reboiler, improper design of reboiler, high amine degradation product levels
 - Consequence: Increased corrosion rates leading to filter plugging and equipment fouling and plugging, corrosion damage leading to reboiler failure, decreased throughput, acceleration of amine degradation
 - Action: Clear overhead line pressure restrictions
Confirm temperature and pressure of heating media, adjust as needed
Skim rich flash drum/reflux drum for hydrocarbons
Decrease reboiler heat input
Remove heat stable salt anions and sodium
Remove amine degradation products (reclaim)
Evaluate heat transfer equipment for coefficient and hydraulics.
Clean reboiler and/or lean/rich exchangers

Part VI

Cost Analysis

Capital Cost and Economic Analysis

Economic Overview

The economics of an industry is a most complicated topic, more suited to the analysis by macroeconomics and labor economics.

That the overall profitability is determined by the various laws of scale as determined by the microeconomic theories coming to fruition in our century from the initial work of Marshall (2013) and Evensky (2015).

Capital Cost

Let us once again consider the plant layout given in Fig. 23.1.

As noted in Peters, Timmerhaus, and West (2003) and Couper (2003), the major contributions to the capital cost are the process industry equipment average and the all industry equipment average which forms the basis of the Marshall and Swift Cost Index, originally the Marshall and Stevens Index, established in the year 1926.

Another more pragmatic, industry agnostic cost index is that of the Chemical Engineering Index, established in the 1960's with the major components being listed in Table 23.1.

Table 23.1 – Major components of CE.

Component	Weight Factor (%)
Equipment, machinery, and supports	61
Construction labor	22
Building materials and labor	7
Engineering and supervision	10
Total	100

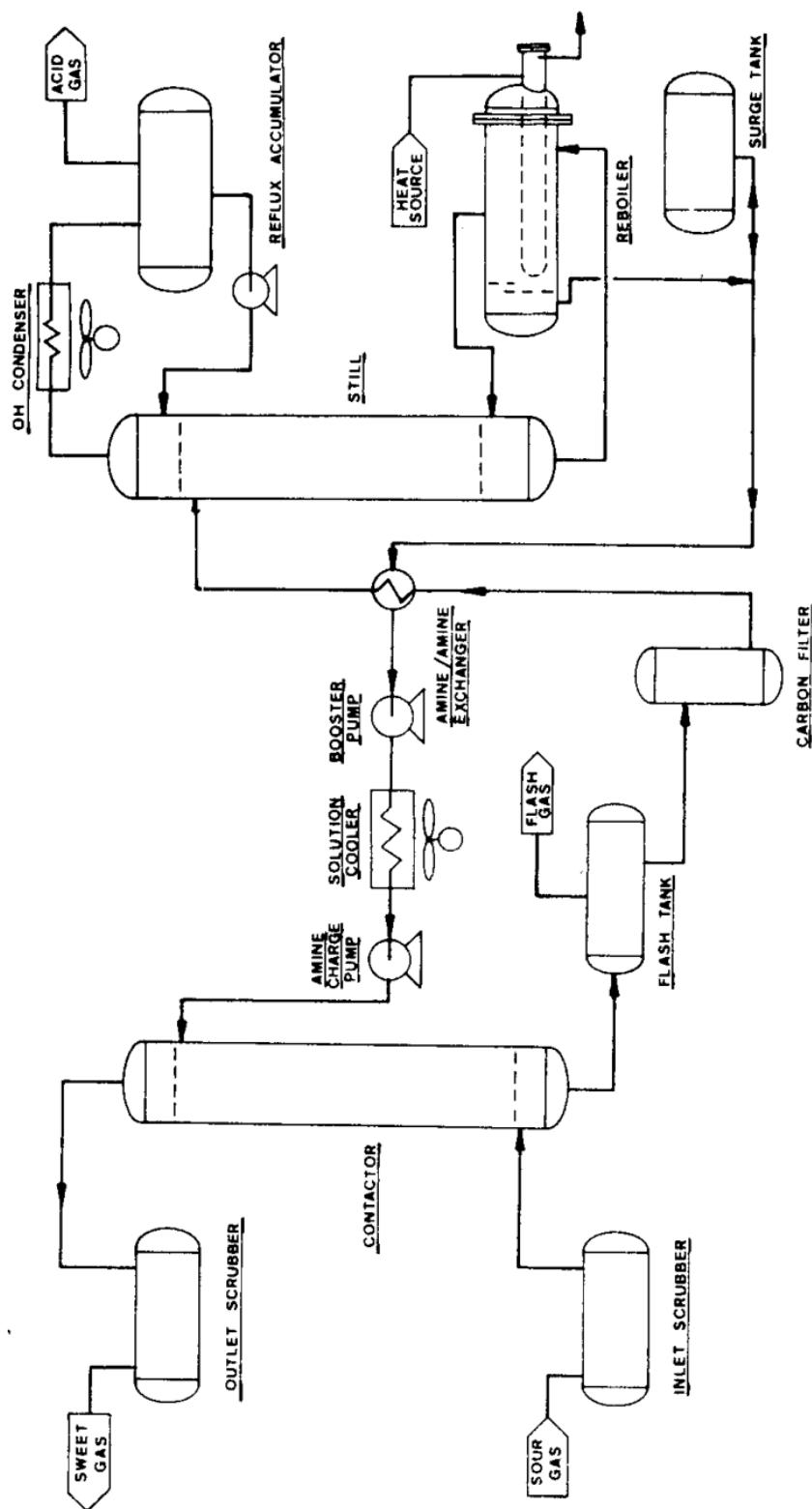


Figure 23.1 – Plant layout from Maddox (1982)

Having procured the relevant costs from the prevalent sources, we can then use a simple cost correlation or a more complicated in house correlation which involves local governments and other factors.

Plant Costs

From our deep analysis of the process and it's material requirements we have prepared the data listed in table below.

Cost Data

Component	Source	Cost
MDEA	Sigma Aldrich (Merck)	INR 36,084.47 per 18 L
SS340 Steel	MEPS International Ltd.	INR 34,000

The steel data is for Hot-Rolled Plates of 15 – 40mm thickness and width over 2.0m.

CEPCI Analysis

The preliminary value for the March CE Plant Cost Index(CEPCI; the most recent value available) rose compared to the previous month's value, continuing a string of six straight months of increasing values. All four of the major subindices - Equipment, Construction Labor, Buildings and Engineering & Supervision - increased in March compared to their February values. The preliminary overall monthly CEPCI value for March 2017 stands at 5.0% higher than the corresponding value from March 2016. Meanwhile the latest Current Business Indicators (CBI) saw a small rise in the CPI Output Index for April, while the March value for the CPI Value of Output declined slightly. The April Productivity Index decreased by a small margin.

The Index is shown in Fig. 23.2.

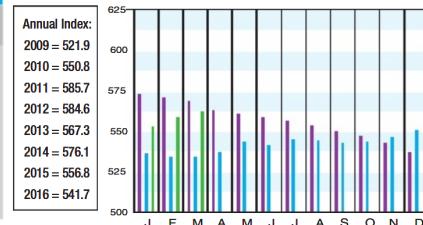
Economic Indicators

2015 2016 2017

Download the CEPCI two weeks sooner at www.chemengonline.com/pci

CHEMICAL ENGINEERING PLANT COST INDEX (CEPCI)

(1957-59 = 100)	Mar.'17 Prelim.	Feb.'17 Final	Mar.'16 Final
CE Index	562.1	558.3	535.3
Equipment	676.6	672.0	638.0
Heat exchangers & tanks	591.0	587.3	545.2
Process machinery	672.1	671.1	644.8
Pipe, valves & fittings	863.7	852.0	800.3
Process instruments	403.4	403.0	383.0
Pumps & compressors	982.3	973.1	969.7
Electrical equipment	514.3	512.1	508.3
Structural supports & misc	733.3	729.7	697.4
Construction labor	326.4	323.1	323.3
Buildings	555.4	552.2	538.4
Engineering & supervision	315.7	315.0	315.7

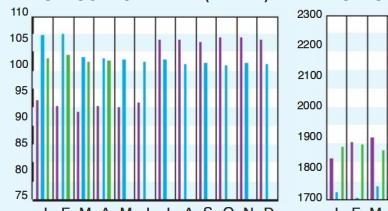


Starting with the April 2007 Final numbers, several of the data series for labor and compressors have been converted to accommodate series IDs that were discontinued by the U.S. Bureau of Labor Statistics

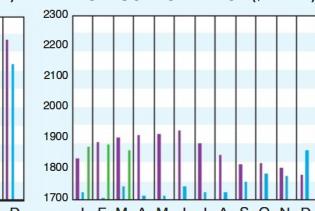
CURRENT BUSINESS INDICATORS

	LATEST	PREVIOUS	YEAR AGO
CPI output index (2012 = 100)			
CPI value of outp ut, \$ billions			
CPI operating rate, %			
Producer prices, industrial chemicals (1982=100)			
Industrial Production in Manufacturing (2012=100)*			
Hourly earnings index, chemical & allied products (1992 = 100)			
Productivity index, chemicals & allied products (1992 = 100)			

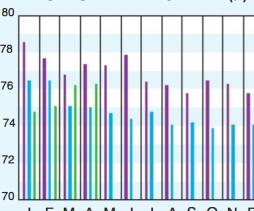
CPI OUTPUT INDEX (2000 = 100)†



CPI OUTPUT VALUE (\$BILLIONS)



CPI OPERATING RATE (%)



*Due to discontinuance, the Index of Industrial Activity has been replaced by the Industrial Production in Manufacturing index from the U.S. Federal Reserve Board.

†For the current month's CPI output index values, the base year was changed from 2000 to 2012.

Current business indicators provided by Global Insight, Inc., Lexington, Mass.

Figure 23.2 – March 2017 Data

Investment decisions are often based upon several criteria, such as annual return on investment (ROI), payback period (PBP), net present value (NPV), the average rate of return (ARR), present value ratio (PVR), or the internal rate of return (IRR). Discounted cash flow rate on return (DCFRR) is another popular means of evaluating the economic viability of a project.

To explain the computational aspect we shall describe various cash flow measures here.

Return on Investment

In engineering economic evaluation, rate of return on investment is the percentage ratio of average yearly profit (net cash flow) over the productive life of the project, divided by the total initial investment. This is calculated after income taxes have been deducted from the gross or pre-tax income. The remainder or net income may be used either for paying dividends, reinvestment, or can be spent for other means. ROI is defined by Eq. 24.1.

$$ROI = \frac{\text{Annual return}}{\text{Investment}} \times 100 \quad (24.1)$$

The annual return may be the gross income, net pre-tax income, net after-tax income, cash flow, or profit. These may be calculated for one particular year or as an average over the project life. Investment may be the original total investment, depreciated book-value investment, lifetime average investment, fixed capital investment, or equity investment. The investment includes working capital and sometimes capitalized expenses such as interest on capital during construction.

Payback Period

Payback period is widely used when long-term cash flows are difficult to forecast, because no information is required beyond the break-even point. It may be used for preliminary evaluation or as a project screening device for high risk projects in times of uncertainty. Payback period is usually measured as the time from the start of production to recovery of the capital investment. The payback period is the time taken for the cumulative net cash flow from start-up of the plant to equal the depreciable fixed capital investment ($C_{FC} - S$). It is the value of t that satisfies Eq. 24.2.

$$\sum_{t=0}^{t=\text{PBP}} C_{CF} = (C_{FC} - S) \quad (24.2)$$

Where:

- C_{CF} is the net annual cash flow
- C_{FC} is the fixed capital cost
- S is the salvage value

Breakeven Point

The payback period is amenable to being calculated by means of a digital computer and a sample labeled figure for understanding graphs generated in this manner is given by Fig. 24.1.

Fig. 24.1 shows the cumulative cash flow diagram for a project. The PBP is the time that elapses from the start of the project, A, to the break-even point, E, where the rising part of the curve passes the zero cash position line. The PBP thus measures the time required for the cumulative project investment and other expenditure to be balanced by the cumulative income.

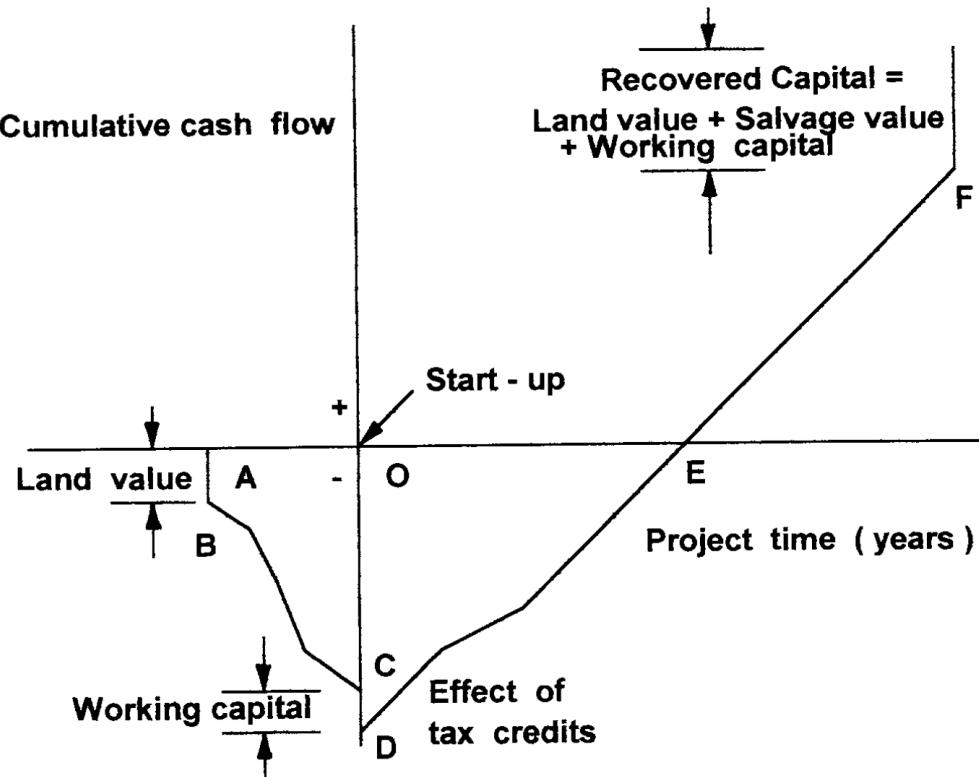


Figure 24.1 – Cumulative cash flow diagram

Present Worth

In an economic evaluation of a project, it is often necessary to evaluate the present value of funds that will be received at some definite time in the future. The present value (PV) of a future amount can be considered as the present principal at a given rate and compounded to give the actual amount received at a future date. The relationship between the indicated future amount and the present value is determined by a discount factor. Discounting evaluates each year's flow on an equal basis. It does this by means of the discount, or present value factor, and the reciprocal of the compound interest factor $(1 + i)^n$ with

i is the interest rate n is the year in which the interest is compounded

The discount factor (D_F) is given by Eq. 24.3

$$D_F = \frac{1}{1+i}^n \quad (24.3)$$

If C_n represents the amount available after n interest periods, P is the initial principal and the discrete compound interest rate is i . The present value, PV , can be expressed as shown in Eq. 24.4.

$$PV = P = \frac{C_n}{(1 + i)^n} \quad (24.4)$$

Part VII

Computational Analysis

For our given system and the well commented code in the Appendix, we used the data as given in Fig. 25.1.

```
1 MDEA
2 64.0 0.787 0.984 1.0
3 S
4 16.0 6.0 0.003 107.6
5 266.0 0.143 38 63.9
6 311.985 1587.33
7 STEAM
8 3.0 255.906 3.0 59.05
9 0.001 212.0 104.0 0.484
10 0.012 0.43 0.17055 47000000
```

Figure 25.1 – Data File Input

- We note that as expected the Reynolds number is laminar, the heat transfer coefficient is obviously a much cleaner and closer approximation to the true value of the system.

The code is adapted from Coker (1995) with modifications in the form of the correlation for long tubes.

The formatted output produced by the program is given in Fig. 25.2

```

1          *SHELL AND TUBE HEAT EXCHANGER RATING*
2 ****
3          TUBE SIDE      SHELL SIDE
4 -----
5  FLUID TYPE:           MDEA        STEAM
6  FLUID FLOW RATE , lb/hr.:    1587.     47000000.
7  SPECIFIC HEAT CAPACITY, Btu/lb.oF:   0.143      0.484
8  FLUID DENSITY, lb/ft^3:       63.900     0.430
9  FLUID VISCOSITY, cP:        38.000      0.012
10 FLUID THERMAL COND. Btu/ft.h.oF:  311.985     0.171
11 FLUID INLET TEMPERATURE, oF:     108.       212.
12 FLUID OUTLET TEMPERATURE, oF:    266.       104.
13 INSIDE DIAMETER, inch:        0.787      255.906
14 OUTSIDE DIAMETER, inch:       0.984
15 TUBE PITCH, inch:            1.000      SQUARE PITCH
16 TUBE LENGTH, ft.:             16.000
17 NUMBER OF TUBES:              64.
18 NUMBER OF PASSES:             6.          3.
19 NUMBER OF BAFFLES:            3.
20 BAFFLE SPACING, inch:         59.0
21 FLUID VELOCITY, ft/sec.:      0.191      ****
22 REYNOLDS NUMBER:              31.        49793652.
23 PRANDTL NUMBER:               0.0422     0.0824
24 FRICTION FACTOR (ft^2/in^2):  0.00000    0.00043
25 HEAT COEFF., Btu/hr.ft^2.oF:  1161.8     17722.5
26 PRESSURE DROP, psi:           0.0038      ****
27 FOULING FACTOR:                0.0030     0.0010
28 LOG MEAN TEMP. DIFF. oF:     18.61
29 OVERALL HEAT COEFF., Btu/hr.ft^2.oF: 203.4
30 OUTSIDE AREA OF UNIT, ft^2:    251.9
31 HEAT LOAD ON UNIT, Btu/hr:    35955.
32 , -----

```

Figure 25.2 – Program Output

We have here presented the project profitability of suitably scaled financial data with judicial factors and assumptions to showcase the generalized technique of our program.

Here we compare the profitability of the sweetening plant in terms of its cash flow based on the natural gas produced and sold, along with an alternate hypothetical situation of the same amount of money being given over to the production and refining of a close competitor, LPG.

The data is taken from ONGC's annual report (2016 – 2017) and covers scaled economic data for 6 years.

- We have compared natural gas and LPG, with LPG data being multiplied by 6.9 to get initial values
- The payback period is within the year for both of these as they are of plants in operation

The LPG input file is given in Fig. 26.1 and the input file for natural gas is Fig. 26.2.

The output files are shown in Fig. 26.4 and Fig. 26.3.

From the data above we assert that given our assumptions and conditions, that LPG is more profitable than natural gas.

- However this does not take into account the diseconomies of scale faced by natural gas production as well as the more stringent environmental regulations on it.
- Additionally it may be physically unfeasible to produce the same amount of cash equivalence in real life due to scarcity of resources.

1	6
2	10
3	126746.1
4	163605.9
5	217239.6
6	208000.5
7	237222
8	241161.9
9	257004.4

Figure 26.1 – Input for LPG

1	6
2	10
3	127544
4	141397
5	165400
6	183291
7	187381
8	182239
9	139398

Figure 26.2 – Input for natural gas

```

1          NET PRESENT VALUE CALCULATION
2 , ****
3
4          7    YEARLY CASH FLOWS INCLUDING YEAR 0
5          FOR LPG
6          10.00  PERCENTAGE ANNUAL DISCOUNT RATE
7 ,
8
9
10         YEAR      CASH FLOWS (INR)      CUMULATIVE
11                           CASH FLOWS (INR)      DISCOUNT
12                           FACTOR      PRESENT
13                           VALUE (INR)
14 ,
15         0        126746.10        126746.10      1.0000      126746.10
16         1        163605.90        290352.00      0.9091      148732.63
17         2        217239.60        507591.60      0.8264      179536.85
18         3        208000.50        715592.10      0.7513      156273.85
19         4        237222.00        952814.10      0.6830      162025.80
20         5        241161.90        1193976.00     0.6209      149742.55
21         6        257004.40        1450980.40     0.5645      145072.27
22
23 0      THE NET PRESENT VALUE (INR):      1068130.04
24 0      PRESENT VALUE RATIO:            7.427
25 0      THE NET RETURN RATE:           140.46 %  

26 0      THE AVERAGE RATE OF RETURN:   174.13 %
27 0      THE PAYBACK PERIOD IS BETWEEN: 0 AND 1 YEARS

```

Figure 26.3 – Output for LPG

```

1          NET PRESENT VALUE CALCULATION
2 , ****
3
4          7    YEARLY CASH FLOWS INCLUDING YEAR 0
5          FOR NATURAL GAS
6          10.00  PERCENTAGE ANNUAL DISCOUNT RATE
7 ,
8
9
10         YEAR      CASH FLOWS (INR)      CUMULATIVE
11                           CASH FLOWS (INR)      DISCOUNT
12                           FACTOR      PRESENT
13                           VALUE (INR)
14 ,
15         0        127544.00        127544.00      1.0000      127544.00
16         1        141397.00        268941.00      0.9091      128542.72
17         2        165400.00        434341.00      0.8264      136694.21
18         3        183291.00        617632.00      0.7513      137709.23
19         4        187381.00        805013.00      0.6830      127983.73
20         5        182239.00        987252.00      0.6209      113156.07
21         6        139398.00        1126650.00     0.5645      78686.53
22
23 0      THE NET PRESENT VALUE (INR):      850316.49
24 0      PRESENT VALUE RATIO:            5.667
25 0      THE NET RETURN RATE:           111.11 %
26 0      THE AVERAGE RATE OF RETURN:   130.56 %
27 0      THE PAYBACK PERIOD IS BETWEEN: 0 AND 1 YEARS

```

Figure 26.4 – Output for natural gas

Part VIII

Conclusions

Summary and Conclusions

To conclude we would to reassert the goals and their appropriate

- We have analyzed with a fully modern and modular approach an important chemical industrial process.
- We acquainted ourselves with the methods and situations faced in the industry in terms of analysis, HAZOP, profit etc.
- We have emphasized the importance of both reference books and industrial indicators in our work.

Apart from the above, we also gained a better understanding of this most fascinating process in terms of its computational and theoretical nuances and we have chosen to end our report on the data in Fig. 27.1, Fig. 27.2 and Fig. 27.3.

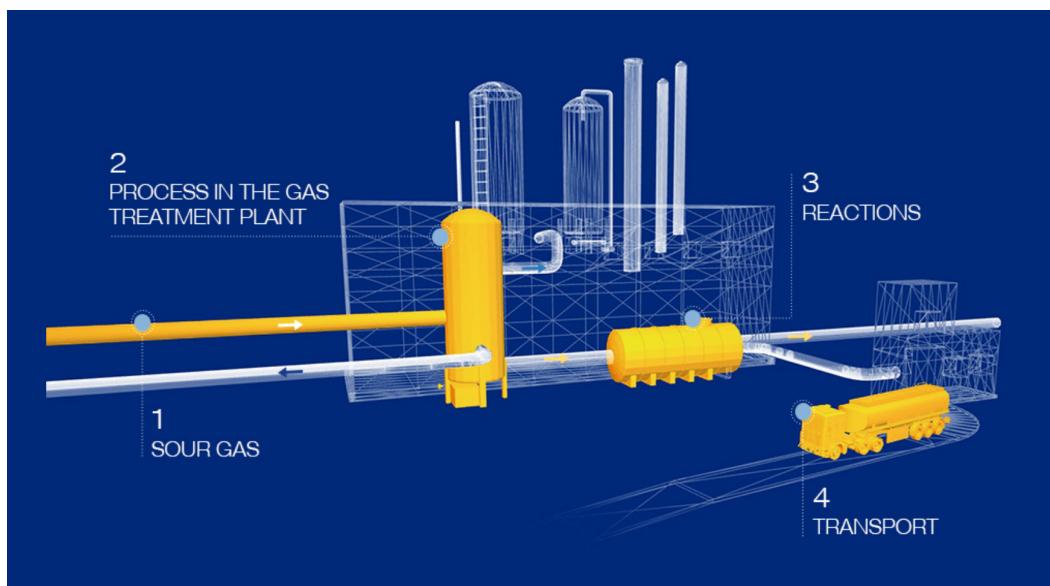
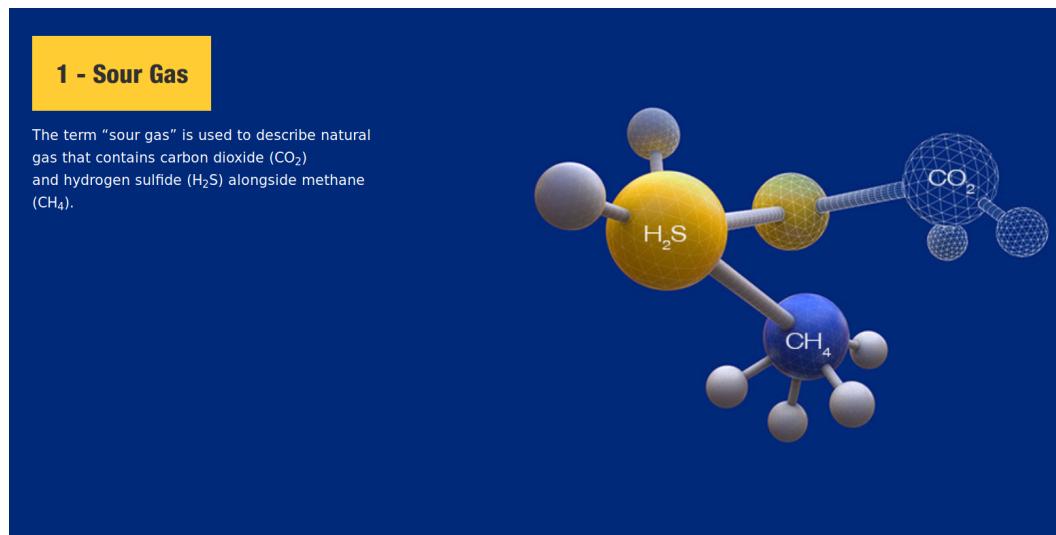
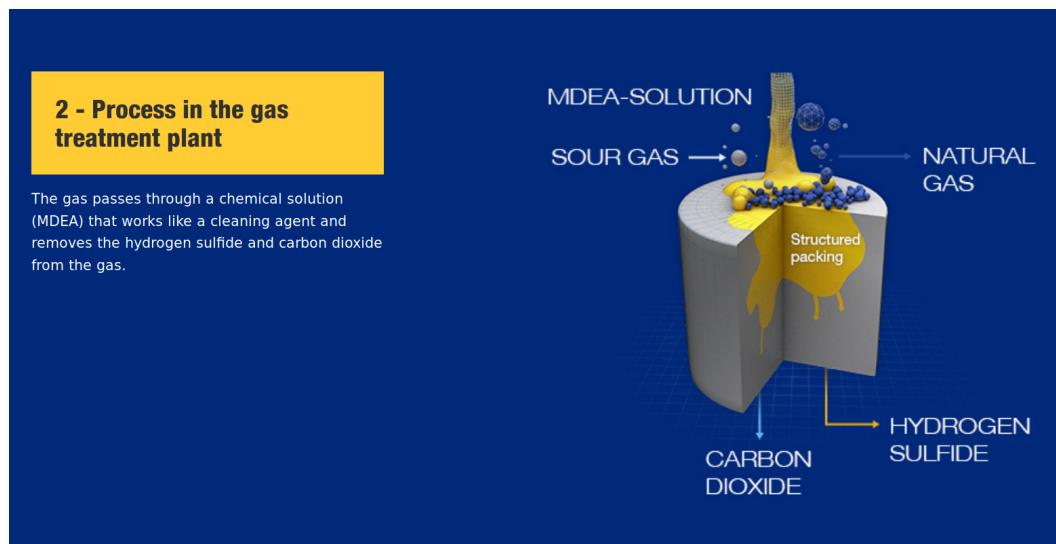


Figure 27.1 – Plant overview

**Figure 27.2 – Sour gas****Figure 27.3 – Process in absorption tower**

Part IX

Appendices

Appendix A: Material Safety Data Sheets



MATERIAL SAFETY DATA SHEET

N-Methyldiethanolamine

Section 1 - Chemical Product and Company Identification

MSDS Name:	N-Methyldiethanolamine, 99+%		
Synonyms:	MDEA; 2,2'-(Methylimino)diethanol		
Company Identification: (INDIA)	Veritas House, 70 Mint Road, Fort, Mumbai - 400 001. INDIA		
For information in the INDIA, call:	Tel: +91 - 22 - 2275 5555 / 6184 0000, Fax: +91 - 22 - 2275 5556 / 6184 0001		

Section 2 - Composition, Information on Ingredients

CAS#	Chemical Name:	%	EINECS#
105-59-9	N-Methyldiethanolamine	99+%	203-312-7

Hazard Symbols:	XI
Risk Phrases:	36

Section 3 - Hazards Identification

EMERGENCY OVERVIEW	
<i>Irritating to eyes.</i>	
Potential Health Effects	
Eye:	Causes eye irritation.
Skin:	May cause skin irritation. May be harmful if absorbed through the skin.
Ingestion:	May cause irritation of the digestive tract.
Inhalation:	May cause respiratory tract irritation. May be harmful if inhaled.
Chronic:	

Section 4 - First Aid Measures

Eyes:	Flush eyes with plenty of water for at least 15 minutes, occasionally lifting the upper and lower eyelids. Get medical aid.
Skin:	Get medical aid. Flush skin with plenty of water for at least 15 minutes while removing contaminated clothing and shoes.
Ingestion:	Get medical aid. Wash mouth out with water.
Inhalation:	Remove from exposure and move to fresh air immediately.
Notes to Physician:	



Section 5 - Fire Fighting Measures

General Information:	As in any fire, wear a self-contained breathing apparatus in pressure-demand, MSHA/NIOSH (approved or equivalent), and full protective gear.
Extinguishing Media:	In case of fire, use water, dry chemical, chemical foam, or alcohol-resistant foam.

Section 6 - Accidental Release Measures

General Information:	Use proper personal protective equipment as indicated in Section 8.
Spills/Leaks:	Absorb spill with inert material (e.g. vermiculite, sand or earth), then place in suitable container.

Section 7 - Handling and Storage

Handling:	Avoid breathing dust, vapor, mist, or gas. Avoid contact with eyes, skin, and clothing.
Storage:	Store in a cool, dry place. Store in a tightly closed container.

Section 8 - Exposure Controls, Personal Protection

Engineering Controls:	Facilities storing or utilizing this material should be equipped with an eyewash facility and a safety shower. Use adequate ventilation to keep airborne concentrations low.
Exposure Limits	CAS# 105-59-9:

Personal Protective Equipment

Eyes:	Wear appropriate protective eyeglasses or chemical safety goggles as described by OSHA's eye and face protection regulations in 29 CFR 1910.133 or European Standard EN166.
Skin:	Wear appropriate protective gloves to prevent skin exposure.
Clothing:	Wear appropriate protective clothing to prevent skin exposure.
Respirators:	Follow the OSHA respirator regulations found in 29 CFR 1910.134 or European Standard EN 149. Use a NIOSH/MSHA or European Standard EN 149 approved respirator if exposure limits are exceeded or if irritation or other symptoms are experienced.

Section 9 - Physical and Chemical Properties

Physical State:	Clear liquid
Color:	colorless - light yellow
Odor:	amine-like
pH:	11.5(100g/l aq.sol.)
Vapor Pressure:	0.026 mbar @ 40 deg C
Viscosity:	101mPa.s @ 20 deg C
Boiling Point:	243 deg C @ 760mmHg (469.40°F)
Freezing/Melting Point:	-21 deg C (-5.80°F)
Autoignition Temperature:	265 deg C (509.00 deg F)
Flash Point:	137 deg C (278.60 deg F)
Explosion Limits: Lower:	0.9 vol %
Explosion Limits: Upper:	8.4 vol %
Decomposition Temperature:	Not available
Solubility in water:	Miscible
Specific Gravity/Density:	1.0380 g/cc
Molecular Formula:	C5H13NO2
Molecular Weight:	119.16



Section 10 - Stability and Reactivity

Chemical Stability:	Not available
Conditions to Avoid:	Incompatible materials.
Incompatibilities with Other Materials	Incompatible materials, acids.
Hazardous Decomposition Products	Nitrogen oxides, carbon monoxide, carbon dioxide, nitrogen.
Hazardous Polymerization	Has not been reported.

Section 11 - Toxicological Information

RTECS#:	CAS# 105-59-9: KL7525000
LD50/LC50:	RTECS: CAS# 105-59-9: Draize test, rabbit, eye: 5 uL Mild; Draize test, rabbit, skin: 500 uL Mild; Oral, rat: LD50 = 1945 mg/kg; Skin, rabbit: LD50 = 5990 uL/kg; Other:
Carcinogenicity:	N-Methyldiethanolamine - Not listed as a carcinogen by ACGIH, IARC, NTP, or CA Prop 65.
Other:	See actual entry in RTECS for complete information. Mutagenicity: Ames-test: negative.

Section 12 - Ecological Information

Ecotoxicity:	Daphnia: EC50: 233 mg/l; 48H; acute Algae: EC50: 37 mg/l; 72H; . Bacteria: EC50: 410 mg/l; 17H; . Fish: Leuciscus idus: LC50: >1000 <2200 mg/l; 96H;
Other:	Biodegradable. log Pow: -1.08

Section 13 - Disposal Considerations

Dispose of in a manner consistent with federal, state, and local regulations.

Section 14 - Transport Information

	IATA	IMO	RID/ADR
Shipping Name:	Not available	Not available	Not available
Hazard Class:			
UN Number:			
Packing Group:			

Section 15 - Regulatory Information

European/International Regulations

European Labeling in Accordance with EC Directives

Hazard Symbols: Xi

Risk Phrases:

- R 36 Irritating to eyes.

Safety Phrases:

- S 24 Avoid contact with skin.

WGK (Water Danger/Protection)

- CAS# 105-59-9: 1

Canada

- CAS# 105-59-9 is listed on Canada's DSL List

US Federal



-
- | |
|--|
| ➤ TSCA |
| ➤ CAS# 105-59-9 is listed on the TSCA Inventory. |

Section 16 - Other Information

MSDS Creation Date:	July 22, 2015
Revision #0 Date	

The Information above is believed to be accurate and represents the best information currently available to us. However, we make no warranty of merchantability or any other warranty, express or implied, with respect to such information, and we assume no liability resulting from its use. Users should make their own investigations to determine the suitability of the information for their particular purposes. In no event shall the company be liable for any claims, losses, or damages of any third party or for lost profits or any special, indirect, incidental, consequential, or exemplary damages howsoever arising, even if the company has been advised of the possibility of such damages.

Appendix B:

Computational Programs

HER

The fortran 90 code for heat exchanger rating is shown below. (abbreviated to HER)

PROGRAM HeatRater

```

! *****
! Adapted and updated to f90 by Rohit Goswami
! and Shaivya Anand
! from Kayode's Fortran programs for chemical
! process design analysis and simulation
! (c) 2018 Rohit Goswami [HaoZeke] <haozeke@yandex.com>
! License: MIT

! THIS PROGRAM RATES A SHELL AND TUBE HEAT EXCHANGER WITH
! EITHER A SINGLE SHELL PASS AND ANY MULTIPLE OF TWO TUBE
! PASSES OR TWO OR MORE SHELL PASSES.
! THE PROGRAM ASSUMES THAT THERE IS NO CHANGE OF PHASE.

!
! THE PROGRAM USES DEVELOPED CORRELATIONS FOR FRICTION FACTORS
! IN BOTH THE SHELL AND TUBE SIDES. HEDRICK'S CORRELATION
! (Chem. Eng. June, 1990) FOR ESTIMATING THE HEAT TRANSFER
! COEFFICIENT FOR TUBES IN THE TRANSITION REGION IS EMPLOYED
! IN THE PROGRAM
! *****

!
! NOMENCLATURE:                                     TUBE SIDE
!
! 1.   FLUID TYPE                               =      FTS
! 2.   NUMBER OF TUBES                          =      NT
! 3.   TUBE INSIDE DIA., inch                  =      TID

```

```

!    4.    TUBE OUTSIDE DIA., inch          =      TOD
!    5.    TUBE PITCH, inch                =      TP
!    6.    PITCH TYPE (TRIANGLE/SQUARE)   =      T/S
!    7.    TUBE LENGTH, ft.               =      TL
!    8.    NUMBER OF PASSES              =      TNP
!    9.    FOULING FACTOR              =      TFF
!   10.    INLET TEMPERATURE, oF        =      TT1
!   11.    OUTLET TEMPERATURE, oF       =      TT2
!   12.    SPECIFIC HEAT CAPACITY (Btu/lboF) =      TSHC
!   13.    VISCOSITY, cP                 =      TVIS
!   14.    DENSITY, lb/ft^3             =      TDEN
!   15.    FLUID THERMAL CONDUCTIVITY (Btu/ft.h.oF) =      TFK
!   16.    MASS FLOWRATE, lb/hr         =      TW
!
!    NOMENCLATURE:                      SHELL SIDE
!
!    1.    FLUID TYPE                  =      FSS
!    2.    NUMBER OF SHELL PASSES     =      NS
!    3.    SHELL INSIDE DIA., inch    =      SID
!    4.    NUMBER OF BAFFLES         =      NB
!    5.    BAFFLE SPACING, inch       =      BSP
!    6.    FOULING FACTOR           =      SFF
!    7.    INLET TEMPERATURE, oF      =      TS1
!    8.    OUTLET TEMPERATURE, oF     =      TS2
!    9.    SPECIFIC HEAT CAPACITY (Btu/lboF) =      SSHC
!   10.    VISCOSITY, cP             =      SVIS
!   11.    DENSITY, lb/ft^3          =      SDEN
!   12.    FLUID THERMAL CONDUCTIVITY (Btu/ft.h.oF) =      SFK
!   13.    MASS FLOWRATE, lb/hr       =      SW
!
! ****

```

PARAMETER (PI = 3.1415927, G=32.174)

CHARACTER*30 FTS, FSS

CHARACTER TYPE

!

```

COMMON/TUBE1/ FTS, TYPE
COMMON/TUBE2/ NT, TID, TOD, TP, TL, TNP
COMMON/TUBE3/ TFF, TT1, TT2, TSHC, TVIS, TDEN
COMMON/TUBE4/ TFK, TW
COMMON/TUBE5/ RET, PRT, TFRF, HT, TV
COMMON/TUBE6/ TDELP, TRDELP, TOTDP
COMMON/SHELL1/ FSS
COMMON/SHELL2/ NS, SID, NB, BSP, SFF
COMMON/SHELL3/ TS1, TS2, SSHC, SVIS, SDEN
COMMON/SHELL4/ SFK, SW
COMMON/SHELL5/ RES, PRS, SFRF, HS, SV
COMMON/SHELL6/ SDELP

COMMON/OUT/ Q, AREA, F, LMTD, U, Q1, CMTD

REAL NT, NS, NB, LMTD

OPEN (UNIT = 3, FILE = 'DATA82.DAT', STATUS ='OLD', ERR = 18)
OPEN (UNIT = 1, FILE = 'PRN')

!     READ THE TUBE SIDE HEAT EXCHANGER PARAMETERS

!     READ THE FLUID TYPE: FTS

READ (3, 11, ERR = 19) FTS
11 FORMAT (A)

!     READ THE NUMBER OF TUBES: NT
!     READ THE TUBE INSIDE DIAMETER, inch : TID
!     READ THE TUBE OUTSIDE DIAMETER, inch : TOD
!     READ THE TUBE PITCH, inch : TP
!     READ THE TYPE OF PITCH (TRIANGULAR: T) or (SQUARE: S): TYPE

READ (3, *, ERR= 19) NT, TID, TOD, TP
READ (3, 12, ERR= 19) TYPE
12 FORMAT    (A)

```

```

!    READ THE TUBE LENGTH, ft. : TL
!    READ THE NUMBER OF PASSES : TNP
!    READ THE TUBE SIDE FOULING FACTOR : TFF
!    READ THE INLET TEMPERATURE, oF : TT1
!    READ THE OUTLET TEMPERATURE, oF : TT2
!    READ THE SPECIFIC HEAT CAPACITY OF FLUID, (Btu/hr.ft.oF) : TSHC
!    READ THE FLUID THERMAL CONDUCTIVITY (Btu/ft.h.oF) : TFK
!    READ THE FLUID VISCOSITY, cP : TVIS
!    READ THE FLUID DENSITY, lb/ft^3 : TDEN
!    READ THE FLUID FLOWRATE, lb/hr.: TW

READ (3, *, ERR=19) TL, TNP, TFF, TT1
READ (3, *, ERR=19) TT2, TSHC, TVIS, TDEN
READ (3, *, ERR=19) TFK, TW

!    READ THE SHELL SIDE HEAT EXCHANGER PARAMETERS

!    READ THE FLUID TYPE: FSS

READ (3, 13, ERR=19) FSS

```

13 FORMAT (A)

```

!    READ THE NUMBER OF SHELL: NS
!    READ THE SHELL SIDE INSIDE DIAMETER, inch: SID
!    READ THE NUMBER OF BAFFLES: NB
!    READ THE BAFFLE SPACING, inch: BSP
!    READ THE SHELL SIDE FOULING FACTOR, SFF
!    READ THE INLET TEMPERATURE, oF: TSI
!    READ THE OULET TEMPERATURE, oF: TS2
!    SHELL SIDE SPECIFIC HEAT CAPACITY, Btu/ib.oF.: SSHC
!    READ THE FLUID VISCOSITY, cP: SVIS
!    READ THE FLUID DENSITY, ib/ft^3.: SDEN
!    READ THE FLUID THERMAL CONDUCTIVITY, Btu/hr.ft.oF.: SFK
!    READ THE FLUID FLOWRATE, ib/hr.: SW.

```

```

READ (3, *, ERR=19) NS, SID, NB, BSP
READ (3, *, ERR=19) SFF, TS1, TS2, SSHC
READ (3, *, ERR=19) SVIS, SDEN, SFK, SW

GO TO 15

18 WRITE (*, 20)
20 FORMAT (3X, 'ERROR MESSAGE IN THE DATA VALUE')
GO TO 999
19 WRITE (*, 21)
21 FORMAT (3X, 'DATA VALUE DOES NOT EXIST')
GO TO 999

! TUBE SIDE CALCULATION

15 CALL TUBE

! SHELL SIDE CALCULATION
CALL SHELL
! CALCULATE THE HEAT LOAD
Q = TW*TSHC*ABS(TT2-TT1)
! CALCULATE THE LOG MEAN TEMPERATURE DIFFERENCE (LMTD)

VAL1 = (TS1-TT2)-(TS2-TT1)
VAL2 = ALOG((TS1-TT2)/(TS2-TT1))
LMTD = ABS(VAL1/VAL2)
! CALCULATE THE APPROACH FACTOR
R = (TS1-TS2)/(TT2-TT1)
P = (TT2-TT1) / (TS1-TT1)

IF (NS .EQ. 1.0 .AND. R .EQ. 1.0) THEN
PX = P/(NS-NS*P+P)
SUM1 = (PX*(R*R+1.0)**0.5)/(1.0-PX)
SUM2 = 2.0/PX-1.0-R+(R*R+1.0)**0.5
SUM3 = 2.0/PX-1.0-R-(R*R+1.0)**0.5

```

```

SUM4 = ALOG(SUM2/SUM3)
F = SUM1/SUM4
GO TO 100
ELSEIF (NS .EQ. 1.0 .AND. R .GT. 1.0) THEN
55   SUM5 = ((R*P-1.0)/(P-1.0))**1.0/NS
PX = (1.0-SUM5)/(R-SUM5)
SUM6 = (R*R+1.0)**0.5/(R-1.0)
SUM7 = ALOG((1.0-PX)/(1.0-R*PX))
SUM41 = ((2.0/PX)-1.0-R+(R*R+1.0)**0.5)
SUM42 = ((2.0/PX)-1.0-R-(R*R+1.0)**0.5)
!   Broken IF maybe. (dangling)

IF (SUM41 .LT. 0.0 .OR. SUM42 .LT. 0.0) THEN
NS = NS+1.
TNP = TNP+2.
GO TO 55
ELSE
SUM43 = ALOG(SUM41/SUM42)
ENDIF

F = (SUM6*SUM7)/SUM43

IF (F .LT. 0.8) THEN
NS = NS+1.0
TNP = TNP+2.
GO TO 55
ELSE
END IF

GO TO 100

ELSEIF (NS .GE. 2.0 .AND. NT .GE. 4.0) THEN
SUM8 = (R*R+1.0)**0.5/(2.0*(R-1.0))
SUM9 = ALOG((1.0-P)/(1.0-P*R))
SUM10 = 2/P-1.0-R+((2/P)*((1.0-P)*(1-P*R))**0.5)+(R*R+1.0)**0.5
SUM11 = 2/P-1.0-R+((2/P)*((1.0-P)*(1.-P*R))**0.5)-(R*R+1.0)**0.5

```

```
F = (SUM8*SUM9)/ALOG(SUM10/SUM11)
ENDIF

! CORRECTED MEAN TEPERATURE

100 CMTD=F*LMTD

! CALCULATE THE OVERALL HEAT TRANSFER COEFFICIENT
! Btu/hr.ft^2.oF

U = 1.0/(1./HT+1./HS+TFF+SFF)

! CALCULATE THE OUTSIDE AREA OF THE HEAT EXCHANGER UNIT ft^2
AREA=(TOD/12.0*TP*NT*NS*TL)

Q1 = AREA*U*CMTD
CALL RESULT

! FORM FEED THE PRINTING PAPER TO TOP OF THE NEXT PAGE.

WRITE (1, *) CHAR(12)

CLOSE (UNIT=3, STATUS='KEEP')
CLOSE (UNIT=1)
999 STOP
END

! *THIS PROGRAM CALCULATES TUBE SIDE HEAT TRANSFER COEFFICIENT,
! * REYNOLDS NUMBER, PRANDTL NUMBER, FRICTION FACTOR AND PRESSURE DR
```

SUBROUTINE TUBE

```
PARAMETER (PI = 3.1415927, G = 32.174)
CHARACTER*30 FTS, FSS
CHARACTER TYPE

COMMON/TUBE1/ FTS, TYPE
COMMON/TUBE2/ NT, TID, TOD, TP, TL, TNP
COMMON/TUBE3/ TFF, TT1, TT2, TSHC, TVIS, TDEN
COMMON/TUBE4/ TFK, TW
COMMON/TUBE5/ RET, PRT, TFRF, HT, TV
COMMON/TUBE6/ TDELP, TRDELP, TOTDP
COMMON/SHELL1/ FSS
COMMON/SHELL2/ NS, SID, NB, BSP, SFF
COMMON/SHELL3/ TS1, TS2, SSHC, SVIS, SDEN
COMMON/SHELL4/ SFK, SW
COMMON/SHELL5/ RES, PRS, SFRF, HS, SV
COMMON/SHELL6/ SDELP

COMMON/OUT/ Q, AREA, F, LMTD, U, Q1, CMTD

REAL NT, NS, NB, LMTD

!      CALCULATE THE FLOW AREA: TFA, ft^2
TFA = (PI*TID**2*NT)/(4.0*144.0*TNP)

!      CALCULATE THE MASS VELOCITY, lb/hr.ft^2
GT = TW/TFA

!      CALCULATE THE REYNOLDS NUMBER: RET
RET = (GT*TID)/(12.0*2.42*TVIS)

!      CALCULATE THE PRANDTL NUT[BER: PRT
PRT = (TSHC*TVIS*2.42)/TFK
```

```

!      CHECK THAT THE REYNOLDS NUMBER IS IN THE TRANSITION REGION
!      (i.e. 2000<RET<10,000)

IF (RET .GT. 2000.0 .AND. RET .LT. 10000.0) THEN
X1 = RET/1000.0
X2 = 10.0*TID/TL
X3 = (1.0-(X1/10.0)**0.256)
X4 = X2**X3
B1 = (-3.08+3.075*X1+(0.32567*X1**2)-(0.02185*X1**3))*X4
HT = (16.1/TOD)*(B1*TFK*PRT**0.33)
GO TO 110

ELSE IF (RET .GT. 10000.0) THEN
HT = 0.027*TFK*(12.0/TID)*(RET**0.8)*(PRT**0.33)
GO TO 110
ELSE
!      Using the Mills Correlation for longer tubes
HT = (3.66+((0.065*RET*PRT*(TID/TL))/ &
& (1+0.04*(RET*PRT*(TID/TL)**0.67))))*TFK/TID
ENDIF

110 HT      = HT*TID/TOD

!      CALCULATE THE TUBE FRICTION FACTOR

IF (RET .GT. 1000.0) THEN
TFRF = 0.0029/(RET**0.25603)
ELSE
ENDIF

!      CALCULATE THE FLUID VELOCITY, ft./s.

TV = GT/(3600.0*TDEN)

```

! CALCULATE THE TUBE SIDE PRESSURE DROP, psi.

TDELP = (TFRF*(GT/3600.0)**2*TL*TNP)/(2.0*G*TDEN*TID/12.0)

! CALCULATE THE PRESSURE DROP OF RETURN LOSSES, psi.

! USING FRANK'S CORRELATION.

TRDELP = (2.5*TNP*TDEN*TV**2)/(2*G*144.0)

*! CALCULATE THE TOTAL PRESSURE DROP EXCLUDING THE NOZZLE IN THE
! TUBE SIDE, psi.*

TOTDP = TDELP+TRDELP

RETURN

END

*!
! *****
! *THIS PROGRAM CALCULATES THE SHELL SIDE HEAT TRANSFER
! *
! *COEFFICIENT,
! REYNOLDS NUMBER, PRANDTL NUMBER, FRICTION FACTOR *
! *AND PRESSURE DROP.
! *
! ******

SUBROUTINE SHELL

PARAMETER (PI = 3.1415927, G=32.174)

CHARACTER*30 FTS, FSS

CHARACTER TYPE

COMMON/TUBE1/ FTS, TYPE

```

COMMON/TUBE2/ NT, TID, TOD, TP, TL, TNP
COMMON/TUBE3/ TFF, TT1, TT2, TSHC, TVIS, TDEN
COMMON/TUBE4/ TFK, TW
COMMON/TUBE5/ RET, PRT, TFRF, HT, TV
COMMON/TUBE6/ TDELP, TRDELP, TOTDP
COMMON/SHELL1/ FSS
COMMON/SHELL2/ NS, SID, NB, BSP, SFF
COMMON/SHELL3/ TS1, TS2, SSHC, SVIS, SDEN
COMMON/SHELL4/ SFK, SW
COMMON/SHELL5/ RES, PRS, SFRF, HS, SV
COMMON/SHELL6/ SDELP

COMMON/OUT/ Q, AREA, F, LMTD, U, Q1, CMTD

REAL NT, NS, NB, LMTD

!      CHECK WHETHER PITCH TYPE IS TRIANGULAR OR SQUARE

IF (TYPE .EQ. 'T' .OR. TYPE .EQ. 't') THEN
DE = ((1.72*TP**2)-(0.5*PI*TOD**2))/(0.5*PI*12.0*TOD)
GO TO 120
ELSE IF (TYPE .EQ. 'S' .OR. TYPE .EQ. 's') THEN
DE = 4.*((TP**2)-(PI*TOD**2/4.0))/(PI*12.0*TOD)
ENDIF

!      CALCULATE THE TUBE CLEARANCE, inch
120      TC = TP-TOD

!      CALCULATE THE SHELL FLOW AREA, ft^2.

IF (NS .GE. 2.0 .AND. NT .GE. 4.0) THEN
SFA = (0.5*(SID*TC*BSP))/(144.0*TP)
GO TO 130
ELSE
SFA = (SID*TC*BSP)/(144.0*TP)
ENDIF

```

```
!      CALCULATE THE SHELL SUPERFICIAL FLOW RATE, Ib/ft^2.hr.  
130 GS = SW/SFA  
  
!      CALCULATE THE SHELL VELOCITY, ft./sec.  
SV = SW/(3600.0*SDEN*SFA)  
  
!      CALCULATE THE REYNOLDS NUMBER  
RES = (DE*GS)/(2.42*SVIS)  
  
!      CALCULATE THE PRANDTL NUMBER  
PRS = (2.42*SSHC*SVIS)/SFK  
  
!      CALCULATE THE SHELL SIDE HEAT TRANSFER COEFFICIENT  
HS = 0.36/DE*SFK*(RES**0.55)*(PRS**0.333)  
  
!      CHECK WHETHER THE SHELL'S REYNOLDS NUMBER IS LESS OR GREATER  
!      THAN 500  
IF (RES .LT. 500.0) THEN  
SFRF = 0.11183/(RES**0.59246)  
GO TO 140  
ELSEIF (RES .GT. 500.0) THEN  
SFRF = 0.01159/(RES**0.18597)  
ENDIF  
  
!      CALCULATE THE SHELL SIDE PRESSURE DROP, psi.  
140 VAL1 = SFRF*(GS/3600.0)**2*SID/12.0*(NB+1.)  
VAL2 = 64.4*SDEN*DE  
SDELP = (VAL1/VAL2)*NS  
  
!  
  
RETURN  
END
```

```
! ****
! THIS PROGRAM PRINTS THE RESULTS OF THE SHELL AND TUBE HEAT
! EXCHANGER ONTO A PRINTER
! ****

SUBROUTINE RESULT

CHARACTER*30 FTS,FSS
CHARACTER TYPE

COMMON/TUBE1/ FTS, TYPE
COMMON/TUBE2/ NT, TID, TOD, TP, TL, TNP
COMMON/TUBE3/ TFF, TT1, TT2, TSHC, TVIS, TDEN
COMMON/TUBE4/ TFK, TW
COMMON/TUBE5/ RET, PRT, TFRF, HT, TV
COMMON/TUBE6/ TDELP, TRDELP, TOTDP
COMMON/SHELL1/ FSS
COMMON/SHELL2/ NS, SID, NB, BSP, SFF
COMMON/SHELL3/ TS1, TS2, SSHC, SVIS, SDEN
COMMON/SHELL4/ SFK, SW
COMMON/SHELL5/ RES, PRS, SFRF, HS, SV
COMMON/SHELL6/ SDELP

COMMON/OUT/ Q, AREA, F, LMTD, U, Q1, CMTD

REAL NT, NS, NB, LMTD

WRITE (1, 150)
150 FORMAT (///,18X,'*SHELL AND TUBE HEAT EXCHANGER RATING*')

WRITE (1, 160)
160 FORMAT (78(1H*))

WRITE (1, 170)
170 FORMAT (40X,'TUBE SIDE',10X,'SHELL SIDE')

WRITE (1, 180)
```

180 FORMAT (78(1H-))

WRITE (1, 190) FTS, FSS

190 FORMAT (2X, 'FLUID TYPE:', T40, A15, T60, A15)

WRITE (1, 200) TW, SW

200 FORMAT (2X, 'FLUID FLOW RATE , lb/hr.:', T40, F10.0, T60, F10.0)

WRITE (1, 210) TSHC, SSHC

210 FORMAT (2X, 'SPECIFIC HEAT CAPACITY, Btu/ib.oF:', T40, F8.3, T60,
& F8.3)

WRITE (1, 220) TDEN, SDEN

220 FORMAT (2X, 'FLUID DENSITY, lb/ft^3:', T40, F8.3, T60, F8.3)

WRITE (1, 230) TVIS, SVIS

230 FORMAT (2X, 'FLUID VISCOSITY, cP:', T40, F8.3, T60, F8.3)

WRITE (1, 240) TFK, SFK

240 FORMAT (2X, 'FLUID THERMAL COND. Btu/ft.h.oF:', T40, F8.3, T60, F8.3)

WRITE (1, 250) TT1, TS1

250 FORMAT (2X, 'FLUID INLET TEMPERATURE, oF:', T40, F8.0, T60, F8.0)

WRITE (1, 260) TT2, TS2

260 FORMAT (2X, 'FLUID OUTLET TEMPERATURE, oF:', T40, F8.0, T60, F8.0)

WRITE (1, 270) TID, SID

270 FORMAT (2X, 'INSIDE DIAMETER, inch:', T40, F8.3, T60, F8.3)

WRITE (1, 280) TOD

280 FORMAT (2X, 'OUTSIDE DIAMETER, inch:', T40, F8.3)

IF (TYPE .EQ. 'T' .OR. TYPE .EQ. 't') THEN

WRITE (1, 290) TP

290 FORMAT (2X, 'TUBE PITCH, inch:', T40, F8.3, T60, 'TRIANGULAR PITCH')

```
ELSEIF (TYPE .EQ. 'S' .OR. TYPE .EQ. 's') THEN
    WRITE (1, 300) TP
    300 FORMAT (2X,'TUBE PITCH, inch:',T40,F8.3,T60,'SQUARE PITCH')
ENDIF

WRITE (1, 310) TL
310 FORMAT (2X,'TUBE LENGTH, ft.:',T40,F8.3)

WRITE (1, 315) NT
315 FORMAT (2X,'NUMBER OF TUBES:', T40, F6.0)

WRITE (1, 320) TNP, NS
320 FORMAT (2X,'NUMBER OF PASSES:',T40,F4.0,T60,F4.0)

WRITE (1, 340) NB
340 FORMAT (2X,'NUMBER OF BAFFLES:',T60,F4.0)

WRITE (1, 350) BSP
350 FORMAT (2X,'BAFFLE SPACING, inch:',T60,F4.1)

WRITE (1, 360) TV, SV
360 FORMAT (2X,'FLUID VELOCITY, ft/sec.:',T40,F8.3,T60,F8.3)

WRITE (1, 370) RET, RES
370 FORMAT (2X,'REYNOLDS NUMBER:',T40,F9.0,T60,F9.0)

WRITE (1, 380) PRT, PRS
380 FORMAT (2X,'PRANDTL NUMBER:',T40,F8.4,T60,F8.4)

WRITE (1, 390) TFRF, SFRF
390 FORMAT (2X,'FRICTION FACTOR (ft^2/in^2):',T40,F8.5,T60,F8.5)

WRITE (1, 400) HT, HS
400 FORMAT (2X,'HEAT COEFF., Btu/hr.ft^2.oF:',T40,F9.1,T60,F9.1)

WRITE (1, 410) TOTDP, SDELP
```

```
410 FORMAT (2X,'PRESSURE DROP, psi:',T40,F8.4,T60,F8.4)

      WRITE (1, 420) TFF, SFF
420 FORMAT (2X,'FOULING FACTOR:',T40,F8.4,T60,F8.4)

      WRITE (1, 430) F
430 FORMAT (2X,'APPROACH FACTOR, F:', T40, F8.4)

      WRITE (1, 440) LMTD
440 FORMAT (2X,'LOG MEAN TEMP. DIFF. °F:',T40,F8.2)

      WRITE (1, 450) CMTD
450 FORMAT (2X,'CORRECTED MEAN TEMP. DIFF. °F:',T40,F8.2)

      WRITE (1, 460) U
460 FORMAT (2X,'OVERALL HEAT COEFF., Btu/hr.ft^2.°F:',T40,F9.1)

      WRITE (1, 470) AREA
470 FORMAT (2X,'OUTSIDE AREA OF UNIT, ft^2:',T40,F9.1)

      WRITE (1, 480) Q
480 FORMAT (2X,'HEAT LOAD ON UNIT, Btu/hr:',T40,F12.0)

      WRITE (1, 490) Q1
490 FORMAT (2X,'HEAT TRANSFERRED, Btu/hr:',T40,F12.0)

      WRITE (1, 500)
500 FORMAT (1H,78(1H-))

!     FORM FEED THE PRINTING PAPER TO THE TOP OF THE NEXT PAGE.

      WRITE (1, *) CHAR(12)

      RETURN
      END
```

PPA

The fortran 90 code for project profitability analysis is shown below. (abbreviated to PPA)

PROGRAM ecoCheck

```

!      Adapted and updated to f90 by Rohit Goswami
!      and Shaivya Anand
!      from Kayode's Fortran programs for chemical
!      process design analysis and simulation
!      (c) 2018 Rohit Goswami [HaoZeke] <haozeke@yandex.com>
!      License: MIT
!
!*****THIS PROGRAM CALCULATES THE YEARLY CASH FLOWS, THE CUMULATIVE
!      * CASH FLOWS, DISCOUNT FACTORS AT A GIVEN RATE AND PRESENT
!      * VALUES. THE PROGRAM FURTHER CALCULATES THE NET PRESENT VALUE,
!      * PRESENT VALUE RATIO, THE AVERAGE RATE OF RETURN, THE NET
!      * RETURN RATE, AND THE PAYBACK PERIOD.
!
!*****ARR      =      AVERAGE RATE OF RETURN
!      YCF       =      YEARLY CASH FLOW
!      DF        =      DISCOUNT FACTOR
!      CUCF     =      CUMULATIVE CASH FLOW
!      NPV       =      NET PRESENT VALUE
!      NOCF     =      NUMBER OF PROJECT LIFE EXCLUDING YEAR 0
!      NRR        =      NET RETURN RATE
!      PV         =      PRESENT VALUE
!      PVR       =      PRESENT VALUE RATIO
!      TIME      =      PAYBACK PERIOD
!
!*****

```

REAL*8 YCF, DF, PV, NPV, CUCF

EXTERNAL XNPV

REAL NRR

INTEGER TIME, TIME1

```

DIMENSION YCF(0:100), DF(0:100), PV(0:100), CUCF(0:100)
!
OPEN (UNIT = 3, FILE = 'DATA91_B.DAT', STATUS ='OLD', ERR=18)
OPEN (UNIT = 1, FILE = 'PRN_ECO_B')
!      READ THE NUMBER OF YEARLY CASH FLOWS EXCLUDING YEAR 0: NOCF
READ (3, *, ERR=19) NOCF
!      READ THE ANNUAL DISCOUNT RATE IN PERCENTAGE: DISC
!      READ THE YEARLY CASH FLOW: YCF
READ (3, *, ERR=19) DISC
READ (3, *, ERR=19) (YCF(K), K = 0, NOCF)

GO TO 10

18 WRITE (*, 21)
21 FORMAT (6X,'DATA FILE DOES NOT EXIST')

GO TO 999

19 WRITE (*, 23)
23 FORMAT(6X,'ERROR MESSAGE IN THE DATA VALUE')

GO TO 999

10 WRITE (1, 100)
100 FORMAT (//,25X,'NET PRESENT VALUE CALCULATION',/1H, 78(1H*))

NOCF1 = NOCF + 1

WRITE (1, 110) NOCF1
110 FORMAT(/,20X,I4,3X,'YEARLY CASH FLOWS INCLUDING YEAR 0')

WRITE (1, 120) DISC
120 FORMAT (/,20X,F8.2,3X,'PERCENTAGE ANNUAL DISCOUNT RATE ',/,
& 1H, 78(1H-))

```

```

WRITE (1, 130)
130 FORMAT (//, 7X, 'YEAR', 5X, 'CASH FLOWS ($)', 4X, 'CUMULATIVE',
& 10X, 'DISCOUNT', 8X, 'PRESENT')

WRITE (1, 140)
140 FORMAT (34X, 'CASH FLOWS ($)', 6X, 'FACTOR', 10X, 'VALUE ($)', 
& /, 1H, 78(1H-))

!      CALCULATE THE CUMULATIVE CASH FLOW

CUCF(0) = YCF(0)
DO I = 1, NOCF
  CUCF(I) = CUCF(I-1) + YCF(I)
END DO

R1 = DISC/100.0
PV(0) = YCF(0)*1.0

DO 15 I1 = 0, 0
  IF (I1 .EQ. 0) THEN
    R1 = 0.0
    DF(I1) = 1.0/((1.0+R1)**I1)
    PV(I1) = YCF(I1)*DF(I1)
  ELSE
    ENDIF
  15 END DO

R1 = DISC/100.0

DO 20 I1 = 1, NOCF
  DF(I1) = 1.0/((1.0+R1)**I1)
  PV(I1) = YCF(I1)*DF(I1)
  20 END DO

```

```

DO K = 0, NOCF
WRITE (1, 150) K, YCF(K), CUCF(K), DF(K), PV(K)
150 FORMAT (5X,I3,5X,F14.2,5X,F14.2,5X,F9.4,5X,F14.2)
END DO

WRITE (1, 160)
160 FORMAT (78(1H-))

!      CALCULATE THE NET PRESENT VALUE FROM THE PRESENT VALUE

NPV = 0.0
DO J = 0, NOCF
NPV = NPV + PV(J)
END DO

WRITE (1, 170) NPV
170 FORMAT (1HO, 6X,'THE NET PRESENT VALUE ($):', 3X, T40, F14.2)

!      CALCULATE THE PRESENT VALUE RATIO, PVR
!      PVR = PRESENT VALUE OF ALL POSITIVE CASH FLOWS/
!      PRESENT VALUE OF ALL NEGATIVE CASH FLOWS
SUM1 = 0.0

DO K = 1, NOCF
SUM1 = SUM1 + PV(K)
END DO

PVR = SUM1 /ABS(YCF(0))

WRITE (1, 180) PVR
180 FORMAT (1HO, 6X, 'PRESENT VALUE RATIO:', 3X, T40, F8.3)

!      CALCULATE THE NET RETURN RATE (NRR) i.e. THE NET PRESENT VALUE
!      DIVIDED BY THE PRODUCT OF INITIAL INVESTMENT AND THE PROJECT LIFE
!      (EXCLUDING YEAR 0).

```

```

NRR = ABS(NPV/(ABS(PV(0))*NOCF))*100

WRITE(1, 190) NRR
190 FORMAT (1HO, 6X, 'THE NET RETURN RATE:', 3X, T40, F8.2, 3X, '%')

!      CALCULATE THE AVERAGE RATE OF RETURN (ARR) i.e THE AVERAGE
!      CASH FLOW DURING THE LIFE OF THE PROJECT (EXCLUDING YEAR 0).

SUM = 0.0
DO I = 1, NOCF
SUM = SUM + YCF(I)
END DO

AVSUM = SUM/NOCF
ARR = (AVSUM/ABS(YCF(0)))*100.0

WRITE (1, 200) ARR
200 FORMAT(1HO, 6X, 'THE AVERAGE RATE OF RETURN:', 3X, T40, F8.2,
& 3X, '%')

!      CALCULATE THE PAYBACK TIME FROM THE CUMULATIVE CASH FLOW, TIME:

DO I=0, NOCF
IF ( CUCF(I+1) .GT. CUCF(I) .AND. CUCF(I+1) .GE. 0.0) THEN
TEMP = CUCF(I+1)
CUCF(I+1) = CUCF(I)
CUCF(I) = TEMP
TIME = I
TIME1 = TIME + 1
GO TO 40
ELSE
ENDIF
END DO

40 WRITE (1, 210) TIME, TIME1
210 FORMAT (1HO,6X,'THE PAYBACK PERIOD IS BETWEEN:', ,

```

```
& 3X,T45,I2,2X,'AND',2X,I2,3X,'YEARS')
```

```
!      FORM FEED THE PRINTING PAPER TO TOP OF THE NEXT PAGE.
```

```
WRITE(1, *) CHAR(12)
```

```
CLOSE (3, STATUS = 'KEEP')
```

```
CLOSE (1)
```

```
999 STOP
```

```
END
```

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