CHAPTER IV

PRELIMINARY EQUIPMENT DESIGN

A preliminary design is required to use as a basis for determining the cost of the plant. At this stage, the cost estimated may vary from the practical one by 25%. The real cost of the equipment applied to the real plant is to be estimated from the detailed design and the price from suppliers of equipments. This step is to be taken after the preliminary study has been made.

The preliminary design is based on approximate design methods and provides only a rough cost estimate. If this study shows that further work is justified, a detailed estimate design may then be developed. The preliminary design work on a process which is already in existence can be done by using the available data on the existing process for further approximation; on the other hand, in a new process more details as to equipment size must be obtained in order to make a more reliable estimate of cost.

Equipment design for this preliminary process evaluation involves determining the size of equipment in terms of parameters as volume, flow per unit time, or surface area, etc. The major equipment for the process are listed in Table IV-1. The size determination of the equipment is done by the following methods.

Table IV-1

Major Equipment for the Ammonia Plant. Utilising Natural Gas as Raw Material.

CP-1	Compressor for compressing feed natural gas
DS-1	Desulfurisation tank
HE-1	Heat exchanger for heating natural gas before the primary reformer
HE-2	Heat exchanger for heating air before the secondary reformer
CP-2	Compressor for compressing feed air
HE-8	Heating exchanger used as waste heat boiler
PR	Primary reformer
SR	Secondary reformer
HE-3	Heat exchanger for cooling the synthesis gas before high temperature shift converter
HTS	High temperature shift converter
HE-4	Heat exchanger for cooling the synthesis gas before the low-temperature shift converter
HE5	Heat exchanger before the flash drum -1
FD-1	Flash drum for separation of water condensate before the carbon dioxide remover
CR	Carbon dioxide remover
RG	Regeneration unit for CO_2 - removing solution
HE6	Heat exchanger for heating synthesis gas before methanation
ME	Methanation unit
FD-2	Flash Drum before the synthesis loop
CP-3	Compressor for compressing synthesis gas
HE7	Heat exchanger for cooling synthesis gas
RF	Refrigeration unit for cooling of synthesis gas
FD-3	Flash drum for separation of ammonia
HE9	Heat exchanger for heating synthesis gas before the synthesis
SC	Synthesis converter
CP-4	Compression of regrigerant for refrigeration
DS-2	Desulfurisation column before the low temperature shift converter
LTS	Low temperature shift converter
HE10	Heat exchanger for the CO_2 refrigerant

4.1 Heat Exchanger Design

Two types of heat exchanger are used in this process. One is the ordinary shell and tube type and the other is the finned-tube type for gas to gas heat exchanger.

The preliminary design is calculated from

$$U_c A \Delta T_{lm} = \text{Heat transferred}$$

where ΔT_{lm} is calculated from the conditioned of the inlet and the outlet, heat load is computed, Uc is the approximate overall coefficient given by

$$Uc = 0.7\ h_i h_o/(h_i + h_o)$$

and the values of the h_i and h_o is approximated from Backhurst and Harker (1973).

4.2 Flash Drum Design

The design criteria for the flash drum is that the flash drum must be able to hold the liquid quantity equal to the volume of liquid flow in 20 minutes. The volume of the liquid holding part of the drum is computed from the volumetric flow rate of the liquid stream leaving the drum; the total height of the drum will be four to five feet higher than liquid holding part height due to the need for space for vapor-liquid separation.

The drum dimension is considered from the most economic figure with the length to diameter ratio in the range of 2.5-5 (Backhurst & Harker, 1973). In this design, the length to diameter ratio is chosen to be 3-5. However for higher operating pressures, FD-3, the length to diameter ratio is increased to 10 to withstand higher stress.

4.3 Pressure Vessels Design

Pressure vessels are used in the desulfurisation units, catalytic reactors, flash drums. In fact, the whole process is under pressurised conditions. Designed pressure is approximately 10% or 10-25 psi, whichever is higher, above the maximum operating pressure. The design temperature is taken to be 50^{O}F above the operating condition. In case of temperatures higher than 650^{O}F, an approximate of 10% over operating temperature is provided for the design temperature (Backhurst & Harker, 1973).

Wall thickness is determined from the relation

$$t = PD/(SE - 0.6P) + C$$

where
- t = Wall thickness, in.
- P = Designed pressure, psi.
- D = Internal Diameter, in.
- S = Allowable working stress, psi.
- E = Welding joing efficiency, usually 0.8-0.95
- C = Corrosion allowance, in. (0.25 in.)

4.4 Column Design

Two dimensions are needed in sizing the tray columns. One is the column diameter and the other is the column height. Column diameter is estimated from flooding considerations by using the Fair's correlation represented in Figure 6-3 of Backhurst & Harker (1973). The chart is a plot of a flow parameter, $(l/V)\ (e_v/e_l)^{0.5}$, versus a capacity parameter,

$U_n \cdot (e_v/e_l - e_v)^{0.5}$, with tray spacing as an additional parameter. In order to stay on the safe side, the maximum value of vapor density, vapor flow rate, liquid density and liquid flow rate are used to determine U_n, vapor velocity for 100 per cent of flooding, 0.8 U_n, is used to calculate the active area of the column by dividing the total volumetric vapor flow rate by this 80 per cent flooding velocity. The gross area of column is equal to 125 per cent of the active area. Then, the column design diameter is selected to the nearest 0.5 ft of the diameter calculated from the gross area. The height of the columns is found by using the tray spacing from which U_n is calculated times the number of actual trays required and plus the height of the top and bottom section. The number of actual trays used is computed from the number of theoretical trays, divided by the overall tray efficiency.

4.5 Compressor Design

The centrifugal compressor is used in this process because of simplicity, maintenance free design and ability to handle long periods of continuous operation. Also the economy of using centrifugal compressor over reciprocating is also another factor. Centrifugal compressors are more economical than reciprocating compressors at the capacity of 600 tons of ammonia/day and above, the details of which is shown by Noyes (1967) and Baasel (1976).

The design procedure for centrifugal compressors is given by Neerken (1975) for the preliminary selection of compressors by using overall polytropic head. From the suction and discharge conditions required and the average properties of the process stream, the following are calculated:

1. The specific heat ratio, $k = c_p/c_v$
2. The average compressibility
3. The acoustic velocity at the inlet from

   $U_a = \sqrt{k\ g\ R\ T_s\ Z_s}$
4. Head coefficient is found from Table IV of Neerken (1976)
5. Polytropic efficiency, n_{poly} from Figure IV of neerken (1976)
6. Polytropic head is computed from

   $$H_{poly} = (\bar{Z}) \frac{(1,545)}{M.W.} (T_s) \frac{(r_c^{(n-1)/n} - 1)}{(n-1)/n}$$
7. $$\frac{(n-1)}{n} = \frac{(k-1)}{k\ n_{poly}}$$
8. The gas horsepower from

   $HP_g(poly) = W\ H_{poly}/33{,}000\ n_{poly}$
9. Total brake horse power is found by adding the friction horse power of the bearing and seals, obtained from Figure VI of Neerken (1976) to the gas horse power.
10. The actual impeller tip speed is computed from

    $$U = \frac{h_{poly} g}{N_{st}\ u}$$
11. The number of stages, N_{st}, is assumed in calculating the U.
12. The resultant speed is cOmpared to the maximum impeller tip speed, which is equal to 0.9 to 1.0 U_a. If the computed value is greater than the maximum allowable value, then a

new value of the number of stages must be assumed.

13. Rotating speed is computed from

$$N = 299 \frac{U}{D} \text{ rpm}$$

In this design, the rpm of every compressor is fixed in order to use the single shaft for all compressors. So the impeller tip speed is computed first from 12, then from 10, the number of stages is determined.

The results of calculation and specifications of the equipment are shown in Appendix B.