

Design Guidelines for Ethylene Quench Towers

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Abstract

Ethylene quench oil and quench water towers may have the highest failure rates in fractional distillation due to the rigorous service that these towers perform. Almost all of the challenge areas of distillation are concentrated into one column system; high temperature, solids, fouling potential, oxygenates, polymerization potential, heat removal by pump arounds, and a mixture of Hydrogen, Steam, and C1 to C20s. Given this combination, the guidelines for designing this column have to be flexible yet sturdy to handle each of the challenges.

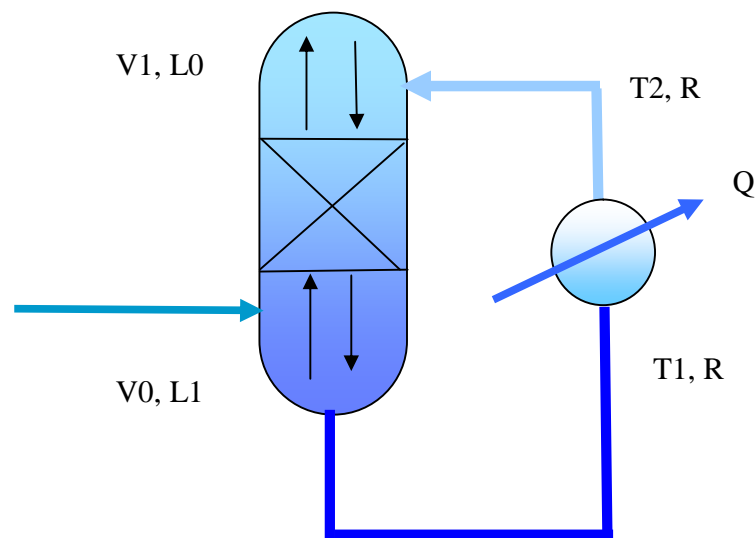
One of the most reliable designs has been the one developed in 1998 (1). These designs have run continually for eight years, with previous designs in these revamped columns having less than one year run length. The authors will examine the history of quench oil and quench water tower's successful and not successful case studies and lessons that can be learned from each of the cases.

Introduction of Quench Towers

There are many process unit quench towers including; Fluidized Catalytic Cracking Units, Vinyl Chloride Monomer Units, Ethylene Oxide, Ethylene Glycol and Ethylene Pyrolysis Cracking Units. The reactor effluent from the process requires cooling for further fractionation and therefore the temperature is reduced or quenched.

Typically quench towers utilize one or more heat transfer sections or pump rounds to remove heat from the column. The use of heat transfer sections or pump rounds results in a better distribution of tower loads than if all of the heat were removed in the tower overhead. Additional benefits include reduced tower diameter at the column upper sections and the recovery of heat at a higher temperature. This higher level of heat can then be utilized in the process for improved energy recovery and higher overall plant efficiency, sometime called specific energy consumption.

Example of a Typical Pump Around Section.



The amount of heat that is removed in the external pump around circuit of a heat transfer section in a column is equal to the exchanger duty Q . The exchanger reduces the pump around liquid temperature from the draw off temperature T_1 to the return temperature T_2 . In a typical design, Q and T_1 are usually set by the heat and material balances and the engineer must select appropriate values of T_2 and the pump around rate R .

These variables are related by the following equation:

$$Q = M C_p \Delta T$$

$$Q = M_{PA} C_{pL} (T_1 - T_2)$$

where:

- Q = Exchanger duty
- R = Pump around liquid rate
- C_{pL} = Liquid specific heat
- T₁ = Pump around liquid draw off temperature
- T₂ = Pump around liquid return temperature

Quench Tower Theory

A reactor produces a hot intense mixture of feed, reactants, inerts and co products. The mixture need to be cooled and separated. In an ethylene plant the mixture is partially cooled by heat exchange and then in a quench tower. The quench tower has heat transfer sections that remove the heat at different levels.

The heat is removed in the heat transfer section by re-circulating liquid against the ascending vapor that enters the tower bottoms. At the bottom of the pump around section the liquid is removed, externally cooled and then returned to the top of the heat transfer section. The challenge of designing the heat transfer sections involves the simultaneous solution of heat and mass transfer equations in which the actual temperature driving force is difficult to evaluate, leading to cases where the number of trays or the height of the packing has been undersized.

The industry method for designing and rating heat transfer section is to utilize a heat transfer unit (HTU). This method is similar to the mass transfer unit (MTU or HEPT) approach to fractionation efficiency. The number of trays or the height of packing for heat transfer can be determined on the basis of heat duty, tower loadings, temperature driving forces and tower area once the mass transfer unit has be established. The mass transfer units are typically established by a process simulator such as PROII or ASPEN.

In 1985 Kulbe, Hoppe and Keller (2) reviewed load flexibility, heat transfer and condensation in packed beds. They address then need to review the vapor flow in each bed separately and / or the need to possibly section the tower for calculation of heat transfer. They used the term tower load profile. The tower sectioning will depend on the relative amounts of latent and sensible heat transferred in each section, followed by separate heat transfer calculations for each section. Sectioning a tower is to obtain realistic temperature driving forces for heat transfer. In general, different values of temperature driving forces are obtained for each tower section.

In 1995 Spiegel, Bomio and Hunkeler (3) developed and published a method for designing or rating direct contact heat transfer sections with packing. The number of overall transfer units for the gas phase NTU_{OG} was calculated based on the enthalpy difference on the gas side assuming no mass transfer resistance on the liquid side.

$$NTU_{OG} = \int_{\text{btm}}^{\text{top}} \frac{dh_G}{h_G - h_I}$$

with

h_G - enthalpy of the bulk gas phase

h_I - enthalpy of the gas phase at the interface

The number of overall transfer units per meter $NTUM_{OG}$ can be calculated by

$$NTUM_{OG} = NTU_{OG}/Z$$

With Z being the packing height.

The $NTUM_{OG}$ depends on the gas and the liquid loads. For a system that is gas side controlled a dependence on the gas load would be expected. The influence of the liquid load may be attributed to the effective interfacial area which depends on the liquid load being in acceptable ranges. The influence of the liquid load on the effective interfacial area is proportional to the velocity of the liquid phase.

$a_{i,\text{eff}}$ proportional to $v_L^{0.2}$

where

$a_{i,\text{eff}}$ is the effective interfacial area

v_L is the superficial velocity of the liquid phase

This has to be compared to random packing where a much stronger dependence on v_L is found. This may be due to the fact that random packing depends more on the liquid load than structured packing. If one utilizes a water based system the overall heat transfer coefficient U can be calculated from the $NTUM_{OG}$.

$$U = \frac{NTUM_{OG} \cdot \rho_G v_G}{a_{i,\text{eff}} \cdot c_{p,G}}$$

with

v_G superficial velocity of the gas phase

ρ_G gas density

$c_{p,G}$ specific heat capacity of the gas

Combing this data the U correlation for 250 X structured packing can be modeled as

$$U = 97.7 F_V^{0.8}$$

With U in W/m^2K , v_L in m/s and F_V in $Pa^{0.5}$. The exponent 0.8 of the gas load F factor is typical for gas side controlled systems. In 1970 Nemunaitis, Eckert, Foote, and Rollison (5) proposed a heat transfer correlation of

$$U = 11.74 Hi^{0.833} Lo^n Go^m$$

which also has a 0.8 exponent.

The overall heat transfer coefficient is calculated using the theory of heat exchangers.

$$U = \frac{Q}{A \cdot \Delta T_{LM}}$$

with

Q heat flow
 A heat transfer area of the packing
 ΔT_{LM} mean logarithmic temperature difference

The analysis of two sets of laboratory data suggest a relationship of the following dependence of the overall heat transfer coefficient

U proportional to $v_G^{0.8}$

When the overall heat transfer coefficient U is plotted against the gas side Reynolds number

$$Re_G = \frac{\rho_G v_G dh}{\mu_G \cos(\gamma)}$$

with

dh hydraulic diameter of packing
 μ_G dynamic viscosity of gas phase
 γ corrugation angle of packing

and the data is regressed an overall heat transfer correlation is developed.

$$U = 0.0925 Re_G^{0.8}$$

The physical properties of fluids in industrial applications are very different from water. To make a possible comparison of the heat transfer data a dimensionless parameter such as the Nusselt number for forced convection must be utilized. (4)

$$Nu_G / Pr_G^{1/3} \text{ proportional } Re_G^{0.8}$$

with

$$Nu_G \text{ Nusselt number} = 4U/(a_1k_G)$$

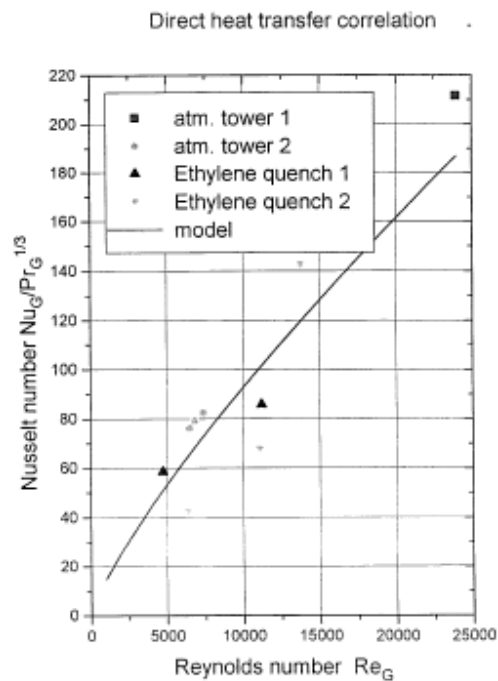
U Overall heat transfer coefficient

k_G thermal conductivity

$$Pr_G \text{ Prandtl number} = \mu_G c_{p,G} / k_G$$

The abscissa in the Nusselt diagram is $Nu_G/Pr_G^{1/3}$, the ordinate is Re_G .

With these correlations and the lab and field data a graph can be constructed of the model and actual field data. (3)



Utilizing this model which matches the field data one can calculate the heat transfer limits of pump around sections. Utilizing a simulator such as PRO II one can then calculated the equilibrium limits and confirm the design of heat transfer pump around sections.

Quench Tower Fractionation Device Selection

In non fouling services most fractionation devices can be utilized for heat transfer sections. Typically trays cost less than other fractionation devices and would be the first choice. In a revamp where higher capacity is required structured packing can be utilized. Packing is best when low pressure drop is desired, while still providing good heat transfer and efficiency. Compared to grid, beds heights can be lower with packing to achieve the same separation.

Fouling services are where the fluids contain solids such as coke, catalyst or scale, and other components that might lead to solid, crystallization or polymer formation. (6) In fouling service the order of preference would be grids, trays, structured packing, and last random packing. The disadvantage of random packing in fouling service is that occasionally one of the random packing will be vertical and the liquid on the horizontal section will have a high residence time leading to fouling. Once the fouling starts it will grow and eventually block the vapor and liquid flows.

Grids are preferred over trays when low pressure drop is desired, entrainment needs to be reduced, and when coking or fouling potential is high due to their low liquid hold up and resident times. Grids have seen excellent service in many quench towers.

Several fouling phenomena can be experienced in quench towers and quench systems; solid fouling, polymer fouling, and coke fouling. Typically quench oil fouling is lowered by the to the following process conditions.

1. There is 30% steam in the quench oil column therefore fouling will be reduced by this inert.
2. There is a high percentage of hydrogen which will reduce the fouling potential, and
3. Many of the fouling components are in the vapor phase.

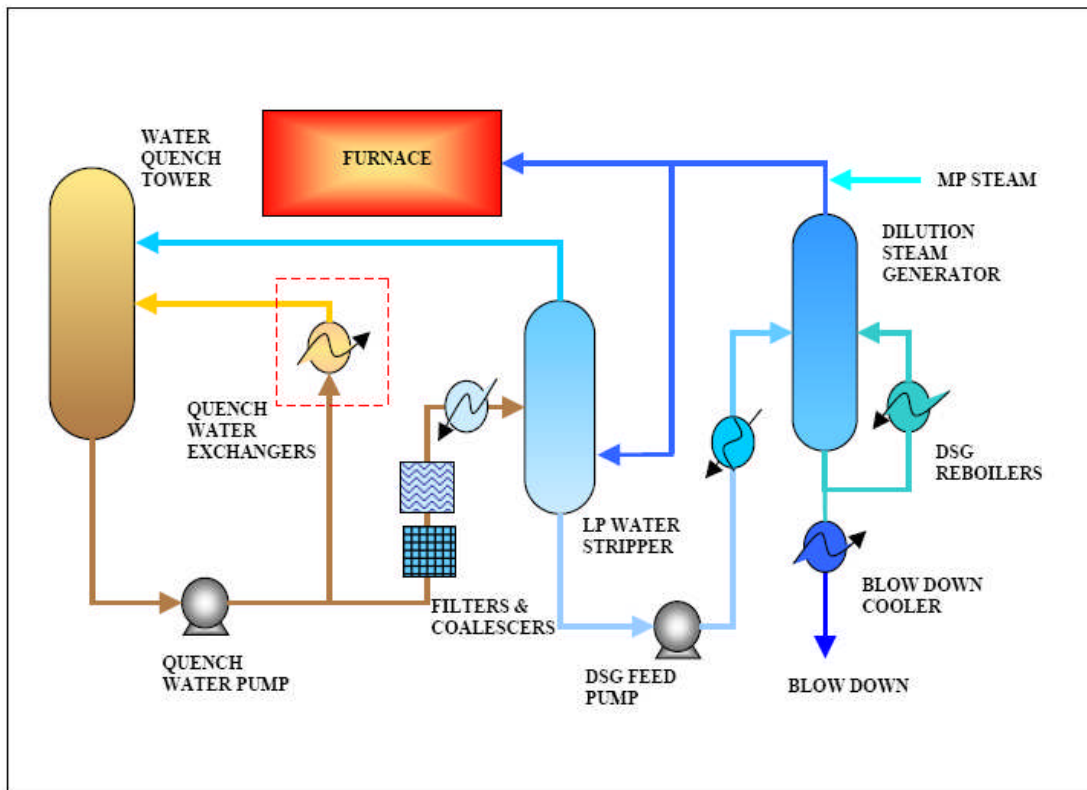
Fouling still does occur in quench oil columns due to the rigorous conditions of the feed stream and designs that increase resident time in the columns. Poor column design is the main reason for quench oil tower fouling. (10)

Solid Fouling

1. Solid Fouling can be seen in quench water towers by naphthalene. Naphthalene is an aromatic compound with a chemical formula of $C_{10}H_8$ and a molecular weight of 128.2. Physical properties include a normal boiling point of 218C and a specific gravity of 1.02. Naphthalene can form a white crystal solid at temperature below its boiling point and may cause fouling in the Quench Water System. Mostly is found in the Quench Water Loop because the Specific Gravity of Naphthalene is close to water, so it preferentially goes with the water phase. Sometimes white solid naphthalene crystals can be found in the suction of quench water pump. This can be remedied by a small stream of pyrolysis gasoline to the system.

Polymer Fouling

1. Polymer fouling can be seen in quench system by polystyrene, many times not in the quench towers, but the adjacent equipment such as the Dilution Steam Generator.



A dilution steam generator was revamped from pan distributors and random packing in 1999 (1). Previously the system was cleaned yearly and after the revamp to low resident time fouling resistant notched distributor and grids, the system is still in service today – 8 eight years later.

Styrene fouling is typically brown to black in color and very hard formations. Polystyrene usually occurs at high temperature with Ferric Oxide as Catalyst. It can be minimize by avoiding hot vapor contact directly to packing with no liquid reflux.

Styrene should not be found in a DSG. If the quench water system pH is not controlled, an emulsion will form and carry the styrene and other hydrocarbon with the water to the DSG, where the polymerization will occur.

2. Polymer fouling can be seen in quench oil towers by polyindene if the residence time is high. It is a yellowish color powder and can accumulate in trays and packing. It can be

minimized by maintain the gasoline reflux and prevent oxygen and oxygenates from entering into system.

Coke Fouling

1. Coke is a co product of olefin production. It is a catalytic reaction with the Fe in the furnace tubes. The coke will partially condense on the furnace tube and partially be swept with the feed and steam mixture to the quench system. The coke will then collect in the any collection system such as pan distributors and random packing. It will also settle into low resident time areas such at the tower bottoms. Some of the coke is removed in the quench oil circulation pump filters.

Olefin Plant Quench Water Systems

In olefins plants, the potential for significant fouling exists in the quench columns that are used to cool the hot process gas from the pyrolysis cracking furnaces. The pyrolysis furnace effluent is a full range mixture of hydrogen, hydrocarbons and steam. Coke fines from the cracking furnaces are entrained with the gas to the first column in the quench unit. This first column will be an oil quench primary fractionator in a liquid cracker or a water quench column in a gas cracker. (9)

In many ethylene units because of the coke fines, the column section above the cracked gas inlet will often utilize open-type baffle trays such as angle trays, disk and donut trays or splash decks depending on the licensor. In light naphtha units dual flow ripple trays have been successful, but in heavy naphtha units there have been some issues with dual flow trays. Some units will use a grid style packing in this section or a combination bed of grid packing with structured packing or trays. Some units still utilize random packing and pan distributors even though they have been shown to be problematic.

As the vapor cools and the worst coke fouling is eliminated, the packing type can be changed to a higher efficiency style. As a result of the additional packing efficiency, the liquid outlet temperature from the column can be increased, resulting in greater heat recovery from the ethylene quench unit. In the upper section of these columns, where fouling is less of a concern, high performance structured packing or trays can be utilized to provide greater efficiency for increased cooling of the process gas.

With the coke fines being removed from the process vapor by the pump around liquid, the liquid at the bottom of the quench column is usually heavy in solids. The pump around liquid is re-circulated to the tower after filtering and heat removal. The filtered re-circulating pump around liquid still contains some fouling material and requires a fouling resistant liquid distributor design such as a larger size spray nozzle distributor or a v-notched weir trough distributor.

The quench water decanter settler can have emulsification problems when the pH of the water is not neutral. This results in the circulation of hydrocarbons back to the quench

water tower with what should be water circulation and to the DSG System. This can be the primary source of fouling.

The water pump around circuits are sometimes integrated with a dilution steam generator (DSG). The DSG performance can impact the quench column operation and styrene fouling can be developed. Some olefins plant has removed the tower internals of the DSG System, but environmentally this is not a good option as one of the DSG functions is to remove phenol from the excess DSG water that is sent to waste water treatment for final disposal.