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 tower 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. Motivated by Kister and Schwartz's concluding remarks that good engineering judgment is needed in evaluating quench towers (2), the authors will examine the history of quench oil and quench water tower's successful and not successful case studies and lessons than 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 would be the case 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.

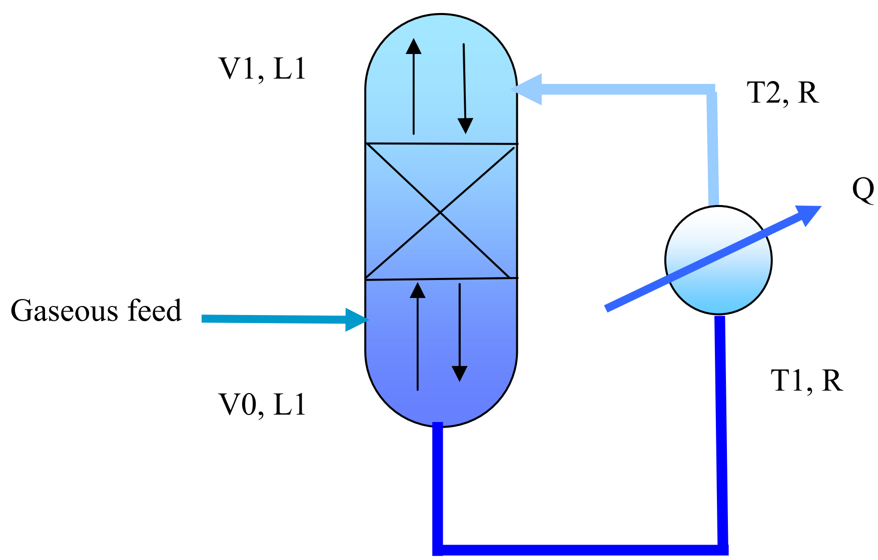

Fig. 1. Example of a Typical Pump Around Section.

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 T1 to the return temperature T2. In a typical design, Q and T1 are usually set by the heat and material balances and the engineer must select appropriate values of T2 and the pump around rate R.

These variables are related by the following equation:

$$Q = M\ C_P\ \Delta T \quad (1)$$

$$Q = M_{PA}\ C_{PL}\ (T1 - T2) \quad (2)$$

where:

Q = Exchanger duty
M_{PA} = Pump around liquid rate
C_{PL} = Specific heat capacity of liquid
$T1$ = 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 exchanger 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 (NTU or HETP) approach to fractionation efficiency, where NTU is number of transfer units and HETP is height equivalent per theoretical plate. 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 (3) 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 (4) 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.

$$NTU_{OG} = \int_{btm}^{top} \frac{dh_G}{h_G - h_I} \tag{3}$$

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

$$NTUM_{OG} = NTU_{OG}/Z \tag{4}$$

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.

$a_{I,eff}$ proportional to $v_L^{0.2}$

where

$a_{I,eff}$ 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 $a_{I,eff}$ of random packing depends more on the liquid load than structured packing. If one utilizes the air/water system, the overall heat transfer coefficient U can be calculated from the $NTUM_{OG}$.

$$U = \frac{NTUM_{OG} \cdot \rho_G \cdot v_G \cdot c_{P,G}}{a_{I,eff}} \tag{5}$$

with

v_G superficial velocity of the gas phase
ρ_G gas density

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

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

$$U = 97.7\ F_V^{0.8} \qquad (6)$$

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_v factor is typical for gas side controlled systems.

In 1970, before structured packing entered the industrial market, Nemunaitis, Eckert, Foote, and Rollison (5) used pilot scale data to correlate the following equation for heat transfer coefficient using random packing (2-inch polypropylene rings):

$$U = C'T_p\ Lo^n\ Go^m \qquad (7)$$

where

C'Tp	system constants
Lo	liquid rate
Go	Gas rate
n	= 0.68 (for dehumidification) or 0.82 (humidification)
m	= 1.0 (for dehumidification) or 0.5 (humidification)

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

$$U = \frac{Q}{A \cdot \Delta T_{LM}} \qquad (8)$$

with

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

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

$$\mathrm{Re}_G = \frac{\rho_G v_G d_h}{\mu_G \cos(\gamma)} \tag{9}$$

with

d_h 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.

$$U = 0.0925\mathrm{Re}_G^{\ 0.8} \tag{10}$$

The physical properties of fluids in industrial applications are very different from air/water system. To make a possible comparison of the heat transfer data a dimensionless parameter such as the Nusselt number for forced convection must be utilized (6).

$$\frac{\mathrm{Nu}_G}{\mathrm{Pr}_G^{\ 1/3}} \text{ proportional } \mathrm{Re}_G^{\ 0.8} \tag{11}$$

with

Nu_G Nusselt number $= 4U/(a_I k_G)$
U Overall heat transfer coefficient
k_G thermal conductivity
Pr_G Prandtl number $(= \mu_G\ c_{p,G}/k_G)$

The abscissa in the Nusselt diagram is $\mathrm{Nu}_G/\mathrm{Pr}_G^{1/3}$, the ordinate is Re_G.

With these correlations with lab and field data a graph can be constructed of the model and actual field data in Fig. 2 (4).

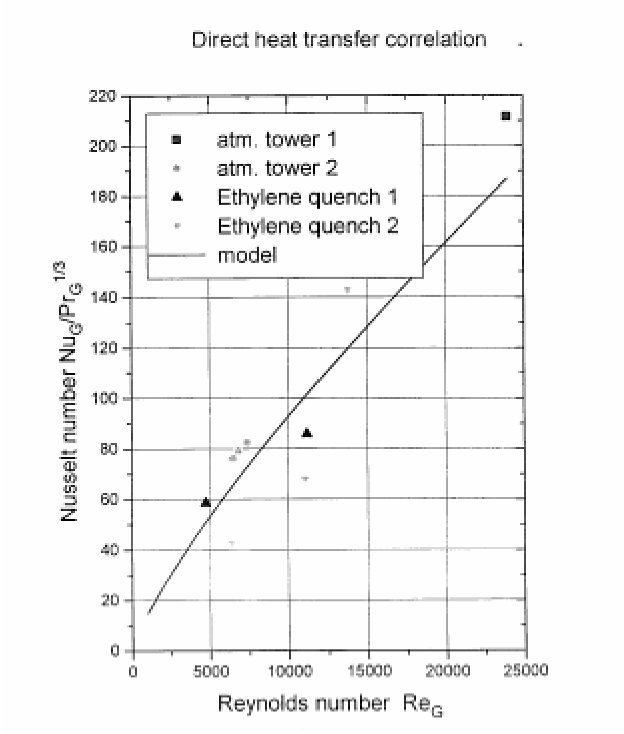

Fig. 2. Test of the heat transfer model by operating data of industrial applications

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. (7) 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. In 2002 Gondolfe and Mueller noted that poor column design is the main reason for quench oil tower fouling. (8)

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 218°C 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.

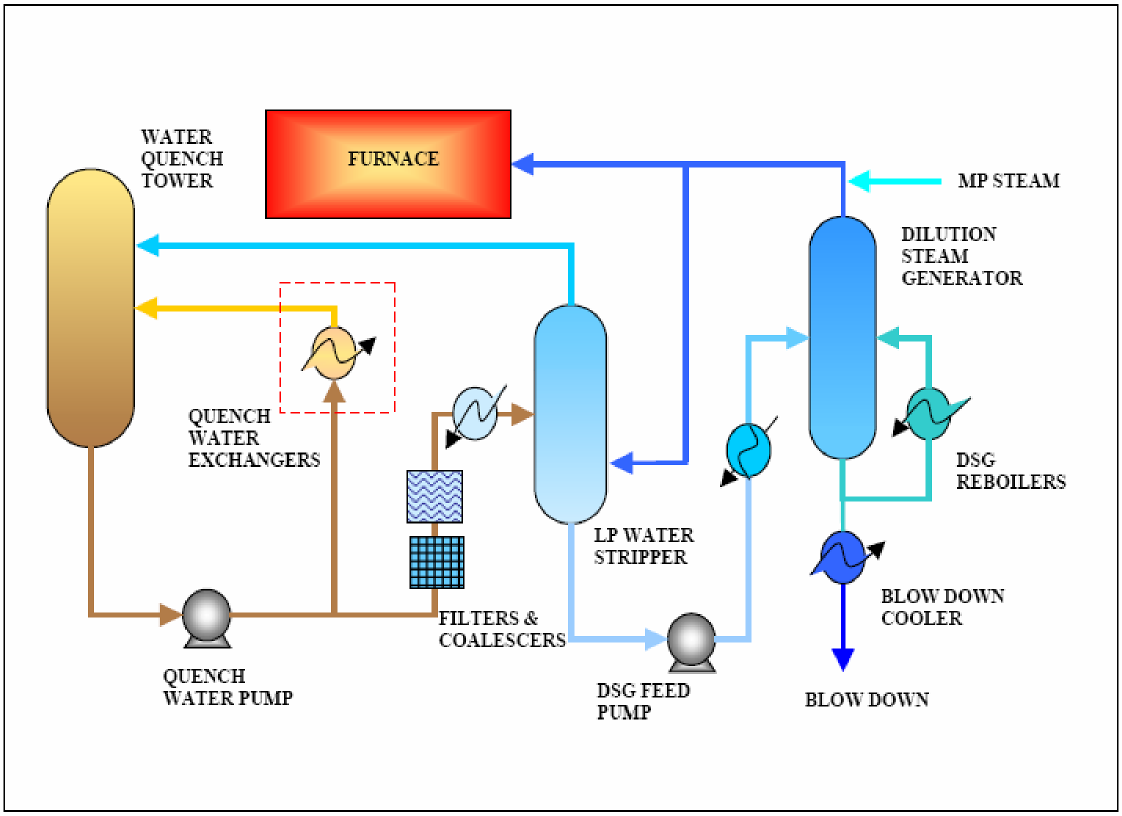

Fig. 3. Process flow diagram of water quench tower and dilution steam generator.

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 caused by 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.

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.

Case One Example - US Gulf Coast Quench Water Tower

An ethylene water quench tower was commissioned in 1992 at a US Gulf Coast ethane gas cracker. The original design of the tower had random packing in the top bed and structure packing in the bottom bed. The tower was upset during start-up and both beds collapsed.

The tower was then hot tapped and spray nozzles were inserted through ring of hot taps around the top of the tower. With the ring of hot taps and spray nozzles, the unit was able to run at full rates, however, the tower's top temperature was higher than design, about 115 to 120 °F compared to a design of 105°F. This led to increase energy consumption in the compressor. The temperature in the bottom of the column was 140 to 160 °F, instead of the design of 180°F. The led to a loss of heat transfer causing increased energy usage per ton of product.

The commissioning of quench water towers can be challenging. The tower is designed to have condensables (steam) and non condensables (cracked gas). During the unit and furnace commissioning there are times when few non condensables are in the furnace effluent, this is called decoking or hot steam stand by. Steam is fed to the furnace to remove the heat in the coils while the furnace is being brought to operating temperatures.

This condensable steam can cause a vacuum in the quench water tower if no non condensables are added. The tower bottoms is kept at 180°F, if the pressure is allowed to drop into a vacuum, the water in the tower bottoms can flash, leading to damage to the structured or random packed bed.

The quench water tower is equipped with a vacuum breaker. It is normally methane or nitrogen that is added by a regulator. If the regulator is inadvertently left blocked in during a start up, and a vacuum is created inside the tower, damage to the tower internals can occur.

A second challenge of vacuum breakers is the location. If the vacuum breaker is placed on the tower outlet, the outlet which is the compressor suction, can still be a positive pressure, while the tower bottoms can be flashing. The vacuum breaker of a quench tower should be placed on the tower inlet to always maintain positive pressure at the tower bottoms.

Due to this flashing phenomenon the tower should be designed with additional mechanical strength. The tray and internals uplift rating should be increased. Normal tray design up lift is 0.1 lbs per square foot. Heavy duty design is 1.0 lbs per square foot. The designs that were installed by the authors had 2.0 lbs per square foot rating. This also tends to point toward utilizing baffle trays and grids in the tower bottoms, because they can withstand the flashing phenomenon better than packing or trays.

In 1995 the tower was revised with random packing in the top and a draw pan for spray circulation in the tower bottom. The draw pan was upset during start up. Later inspection found hold-down clamps were not properly installed and bolts were missing. The ring of spray nozzles was reinstalled.

In 2000 the circulation rate was suddenly reduced, as if a line plugged overnight. The tower was shut down for inspection and a loose vapor riser hat from the draw pan was found in circulation draw. The vapor riser hats were welded in place improving the previous bolting design.

In September 2003 the tower was converted to v-notch liquid distributors and grid packing. The tower is still performing with lower pressure drop and improved heat transfer capacity than the original design.

Case Two Example - US Gulf Coast Quench Water Tower

An ethylene water quench tower was commissioned in 1997 at a US Gulf Coast ethane/propane gas cracker. (11) Because the unit was designed to crack E/P only, an oil quench column was not installed. The water quench tower had a combination of spray nozzles and random packing.

The column was configured with two beds of packing. The top bed of packing used nominal 2" random packing with a spray nozzle distributor for liquid irrigation of the primary quench water feed. A deck type re-distributor was used to collect liquid from the upper bed and redistribute it to the lower bed. A secondary water quench was introduced over the lower packing bed via a spray nozzle distributor, which was located below the deck re-distributor.

The bulk of the bottom bed was packed with a nominal 3" random packing plus a small section of grid style packing for higher fouling resistance.

This column had a history of pressure drop problems. The pressure drop in this column typically ranged from 1 psi to 1.5 psi while the calculated pressure drop at design conditions was approximately 0.3 psi. The hydraulic capacity ratings of the packing were less than 70% at the design conditions. The column was kept in service despite continued pressure drop problems and attempts to mitigate the fouling by modifying the unit operations and maintenance practices.

In 2000, the quench tower was starting to show signs of reduced heat transfer capacity. Additionally, the three pound pressure drop increase is a very large operating cost due to the fact that the pyrolysis furnace reactor yields are much higher with lower pressure.

The high quench tower pressure drop increases the reactor pressure and greatly reduces the net yield of the plant.

Table 1. Comparison of operating data before and after revamp.

	2000 Operating Data Prior to Revamp	**2005 Operating Data After Revamp**
Cracked Gas Flow, Lb/Hr	**100% of original design**	**121% of original design rates**
Quench Water Bed 1, GPM	**2197**	**2294**
Quench Water Bed 2, GPM	**1850**	**1668**
Total Quench Water, GPM	**4047**	**3962**
Top Temperature Approach, °F	**2.3**	**0.4**
Overhead Vapor Temperature, °F	**101**	**95.3**
Bottom Water Temperature, °F	**186**	**179.4**
Column Temp Delta, °F	**85**	**84**
Overhead Pressure, psig	**13.2**	**16.9**
Column Pressure Drop, psi	**3.0**	**0.23**
Number of Heat Transfer Units developed per Bed	**-**	**6.7 top / 3.3 bottom**
Total Number of Stages	**-**	**4 top / 2-3 bottom**

Case Three - Light Naphtha Crackers

Two identical light naphtha crackers were commissioned in South East Asia. The first unit was commissioned in 1995 and the second in 1999. These crackers were designed for a feed stock of heart-cut naphtha with about 80 to 85% paraffins. A second option for designing naphtha crackers may be to what is called an “open spec” naphtha which has higher end points and 60% paraffins. A third option is to design for gas oil cracking which has much higher feed stock end point and much lower paraffins.

These light naphtha crackers had quench oil and quench water columns with dual flow ripple trays. There was no feed inlet device, baffle trays or middle draw on the quench oil tower. These columns have performed very well on light naphtha service but would be challenged on open spec or gas oil feed stocks.

An open spec Naphtha Cracker in South East Asia has been performing well with shed decks in the tower bottoms, followed by spray nozzles, random packing and pan distributors. Several points of operation are beneficial to South East Asia tower operations and they may not be representative of other regions.

1) There was a government mandated shutdown every 24 - 36 months for the first few years of operations. A shutdown and system cleaning every three years can eliminate

many of the potential fouling precursors. The challenge is the 4 and 5 years of tower operation.

2) There is a shortage of propylene in the region and the units have low cracking severity to produce more propylene. Lower cracking severity produces less tar, coke and other co products.

3) Due to overcapacity in the region the units have ran at about design rates during 2000 to 2003 time period. Higher rates challenge the towers more than design rates.

Guidelines for Increased run length guidelines include;

1. Restrict the use of pan distributors and random packing in fouling services. (7) Quench Oil and Water towers can experience coke fine, particles and flashing. Pan distributors and random packing serve as excellent collectors for the coke and other solids. These designs have a few successes and many not successful operations. In 2002 Kister and Schwartz presented their findings on differences between shed decks and packings in quench towers. (2)

2. Restrict the use of movable valve trays in fouling services. Dual flow trays or fixed valves trays that direct the flow across the tray are a better choice. Consider utilizing devices like push valves that eliminate the high residence time areas of cross flow trays. Stepped outlet weirs can be utilized, but are not required in the upper section of the columns.

3. Be cautious of component trapping in quench oil towers. Some species of hydrocarbon are too light to exit out the bottoms and too heavy to be removed in the overhead, and can even be concentrated if the quench oil tower has a middle draw pan.

When component trapping occurs a very high residence time can be encountered, partially in conventional or dual flow trays, leading to fouling on the tray. Pictures of up to 18 inches (450 mm) of fouled material on a tray were presented in a conference in 2001. (9) Restrict the use of pan distributors, conventional trays or dual flow trays in high residence and fouling areas of the tower.

4. Utilize low residence time devices in high fouling areas. Residence time and efficiency are inversely proportional. The best devices for fouling are baffle trays and grids. They have lower efficiency but 5+ years run length versus 1 year run length or in one case 45 days.

5. The most reliable designs have baffle trays, followed by grids, then other devices can be utilized above the grids such as dual flow ripple trays or conventional cross flow trays. This configuration has less efficiency than other configurations, but will have a very good run length. In the upper section the choice between ripple trays and cross flow trays can be the vendor’s choice.

6. In quench water towers structured packing has been used in the top sections successfully, but there is mixed results for structured packing in quench oil towers and very poor results for most random packing in each service. But some quench oil towers have good results even with random packing, because on these towers the fouling potential is low. If the unit fouling potential is low, structured packing can be utilized on the top of the quench oil tower, but trays are a safer choice.

The fouling potential is a function of feed stocks, cracking severity and design capacity. If the unit processes light naphtha feed stocks, maintains cracking severity low to produce propylene and keeps the capacity about the rated design, then fouling potential is reduced.

7. Consider heavy duty designs in quench towers. The potential exist for flashing to occur if the quench water tower is allowed to condense into a vacuum. The authors have recommended and utilized a 2 lb per square foot uplift design.

Conclusions

The optimum number of heat transfer trays or packed height is based on an economic study involving tower height and diameter, external heat exchanger size, pump and power costs and tower reliability. By far the largest economic driver is reliability. Many quench towers run lengths are measured in days due to lack of understanding of the basic principles

One definition of insanity is to repeat the same experiment and expect different results. If a particular design has had multiple failures, one should consider modifying the design or expect continued failures. Practical engineering is an ongoing learning experience from past successes and failures. If you have had no failures, you have not been working hard enough, but it is very important to be learning from your mistakes and the mistakes of others.

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