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In This Issue

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SPECIAL FEATURES

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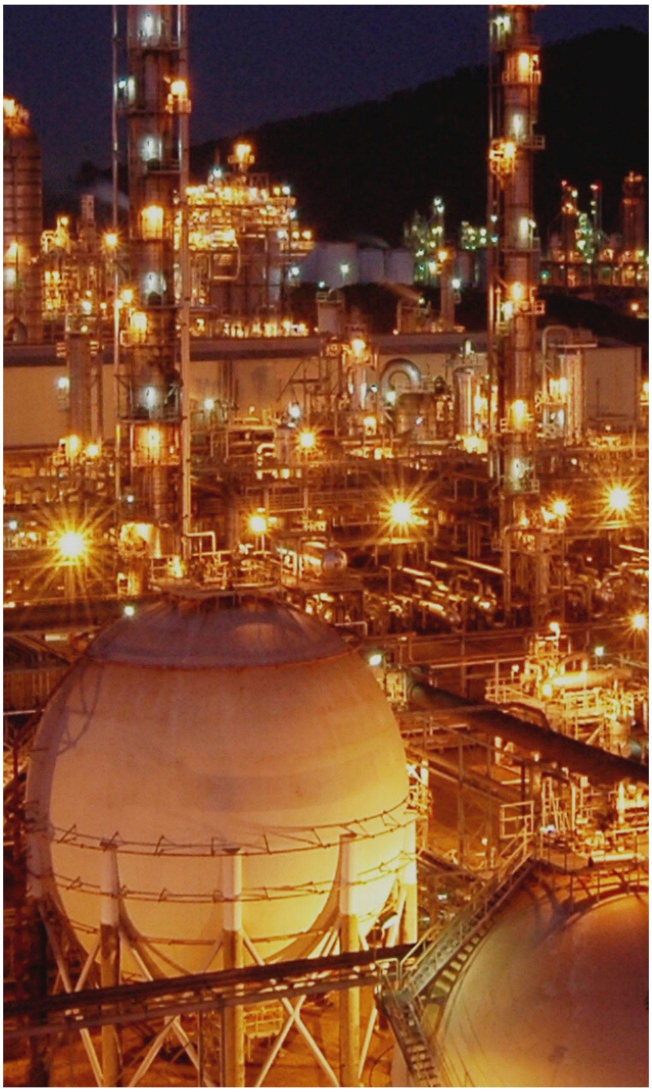

Letter from the Editor

Welcome to March.

It has been particularly cold this year in the USA, so we are looking forward to spring and warmer weather. As we look forward to the future we hope that travel can return and we can again speak at some colleges later this year. We try to inspire students to continue their education by utilizing IACPE Practical Design Modules to learn to utilize the fundamentals they acquired in Universities. The IACPE Title and Certificate is only a symbol of the practical knowledge that you may gain from the modules that will assist you for the rest of your career.

In the Magazine this month we again have great articles from our great contributors. We also have a special Question and Answer Distillation Roundtable.

All the best in Your Career and Life.

All the best,
Karl Kolmetz

2021 EDITORIAL CALENDER

Month	March	May	July	September	November
Industry	Refining	LNG	Midstream & pipelines	Ethylene	Polymers
Unit operation	1. Mass Transfer (Distillation) 2. Catalyst Systems	1 Process control Systems 2. Refrigeration Systems	1. Flaring Systems 2. Corrosion Control	1. Fired heaters 2 Rotating Equipment	1. Reactors Systems 2. Pressure Vessel Design
Safety, Sustainability & Digital	Biorefining	Big data	Process Safety	Water treatment – Boilers and Cooling Towers	Environmentally Friendly Circular plastic

Rock Bottom View

Texas Blizzards & Hydrates….Friend or Foe?

Ronald J. Cormier, *Engineering Practice* Contributing Author

We hope our readers felt at least a bit heartened after scanning The View from Rock Bottom in January's edition of Engineering Practice Magazine. We had entered the dark tunnel that is the COVID-19 pandemic, and were globally ambushed without any existing anecdote for the virus. Hence, 2020/21 registered tremendous societal stresses, strains, and fatalities. Also, a never-before experienced period of huge uncertainly, occurring simultaneously on a global scale. We remain hopeful that as vaccines are now spreading through regional populations, and that some sense of returning normalcy is shining its light at the end of this horrible tunnel. Critical characteristics of petrochemical-based packaging, durable goods, and medical supplies continue to help make this possible and should also bode well for hydrocarbon industry's health.

While we were still pondering the miracle of vaccine science, February's arctic deep freeze rolled into Texas. Ten days of sub-freezing weather, decimated electricity, running water, and home heating. The disaster caused misery and death across the state, and much of the southern US. The expected finger-pointing and blame game ensued including ERCOT, various generators, environmentalists-vs. fossil fuelers, state leaders, and the public.

Plenty of unplanned events were blamed for the blackout, but one that continued to confuse me was the significant percentage of generating capacity knocked offline, not due to frozen turbines, generators, transformers, or iced distribution lines, but more upstream in the equation—frozen fuel gas delivery. Whaaaaat???…can the natural gas system freeze?

YES it can. Unrefined natural gas has a bad habit of freezing, forming hydrates in pipes at any temperature below 60 F. Hydrates are a real concern to O&G companies, in and above ground at high pressure natural gas wells. Expanding gas lowers the temperature (the principle behind how your refrigerator and AC functions) and hydrates can form in piping at points where pressure drops exist. Hydrates can form in any weather but cold weather is most common. They can also form in unexpected places in the lines. Hydrates can be prevented but not necessarily removed with the injection of methanol into the piping. Methanol is hygroscopic and can keep hydrates from forming. As the possible culprit, whether or not appropriate application of alcohol was at play in Texas' gas system this time may play a rather insignificant role vs. the phenomena of how hydrate formation might actually HELP us in the future.

The Energy Information Administration estimates that hydrates contain more carbon than each petroleum product accessible on Earth combined. Also, that these ice-like structures could hold somewhere in the range from 10,000 trillion to in excess of 100,000 trillion cubic feet of flammable gas. By comparison, EIA said in 2013 that there are a little more than 7,000 trillion cubic feet of actually recoverable shale gas stores throughout the world.

This potential isn't lost on federal organizations. DOE has been directing an investigation into methane hydrates on and off since the 1980s, and as of late, has re-increased its promise to understanding the substance. Earlier this year the DOE declared that it intends to direct approximately $5 million into subsidizing research ventures investigating the potential fuel source and how it very well may be developed and commercialized.

"We realize that methane hydrates hold immense potential as a future energy asset," said Ray Boswell, the DOE program chief on methane hydrates for the division's National Energy Technology Lab, . "We've moved beyond the subject of 'does this substance truly exist and is it accessible'….now we're truly attempting to sort out its amount…[sic] to where we can reasonably think of it as a portion of our future energy save, and we're running after that."

When inquired as to whether methane hydrates remain a part in the Biden administration's energy thesis, Boswell answered: "Indeed, totally."

Friend or foe?: going forward, keep and eye on hydrates commercialization. Sometimes embracing a problem provides immense opportunity!

Questions & Answers

Distillation Roundtable

Panelist : Andrew W Sloley, Mike Resetarits, Karl Kolmetz, Henry Kister, Christian Geipel

1. WHERE DO YOU FIND MOST OF THE CHALLENGES IN DISTILLATION DESIGN: A) SIMULATION, B) HARDWARE DESIGN, C) INSTALLATION, E) COMMISSIONING?

Andrew: I will define the most challenging area as the one that appears to be most prone to have mistakes. Everyone's opinion will be based on their own work history. I would say I have seen the most problems in hardware design followed closely by simulation. The underlying causes differ.

Most hardware design is done by vendors. They are subject to competitive pressure to deliver the lowest cost equipment. They are also often given insufficient information about the potential range of process conditions.

Simulation work tends to be given to the newest engineers because they are perceived to be most computer savvy. This means a lot of simulations get done by inexperienced people.

Problems do happen during installation and commissioning, but direct comparison of the number of incidents is misleading. A plant may only design a tower (simulation and hardware) once, but that tower may be shut down and started up many times during its life. The number of installation and commission problems may seem high, but that is because there are lots of shutdowns, repairs, and startups.

Mike: All these aspects present incredibly significant challenges. With each aspect, "all" feasible and reasonable options should be considered. Even more importantly, the opinions of experienced engineers and technicians should be sought vigorously.

Karl: As I have been on many tower inspections therefore, I have seen many installation errors which we corrected. Without a senior inspector many of the errors would be undetected. Most of the installation errors will not stop the column from performing, but they will limit the upside potential of capacity and recoveries. As senior distillation inspector that understands hardware and process design can be a multi-million dollar return on investment. Just a small 2% additional capacity gain is a huge return.

Most commissioning are completed without many issues as most plants put together good commissioning teams. Tray and packing damage tend to happen later in unexpected outages and power failures.

Christian: Good simulation requires experience. First, it is important to know the limitations of the selected models and their applicability for the process being simulated. Second, experience is necessary to judge the reliability of the property data being used. We have seen failures of simulations due to both reasons.

To convert simulation results into optimum hardware design, the hardware designer needs to know the background of the simulation cases. For example, "What are startup/ shutdown cases, what are "rating cases" where lower efficiency or higher energy consumption might be acceptable? What are the "real" production cases where you want the optimum performance of the unit?" The same is valid for specified turndown ratios: Does the column really need to perform well at 20% turndown or is it only a startup case?

Based on this, the challenges for the hardware engineer are:

- Minimizing investment: challenge is choice of material, minimizing tower dimensions.
- Maximize flexibility: Find the optimum design to cover the specified turndown range.
- Maximize separation efficiency: challenges are choice of adequate structured or random packing, and / or tray design and tray spacing.

I would say that simulation and hardware design are the two main challenges in distillation design. Installation and commissioning also have their challenges and failures can lead to tremendous costs, but I would not say that these problems are typically "design related".

2. DESCRIBE THE CHALLENGES IN SIMULATION AND CHOOSING THE CORRECT EFFICIENCY.

Andrew: I will stick to tray efficiency, though many of the same general observations apply to packing. Efficiency estimates have different approaches when evaluating an existing tower versus designing a new tower.

Nearly every time one evaluates an existing tower, the best approach is to adjust the number of theoretical stages in the model until the results match the unit performance in the field. The theoretical stages divided by actual trays gives a tray efficiency that is then applicable to an entire section of the tower. Only very rarely does using partial tray efficiencies in simulation software give useful results. While they allow for different efficiencies for multiple components, they tend not to reflect what engineers mean by efficiency. Additionally, using the partial tray efficiencies option often leads to solution instability and high execution times, even with modern software and computers.

When designing a new tower, the best starting point is to find demonstrated section efficiency in similar equipment being used in a similar or identical service. If such information is not available, the most useful efficiency correlations are the O'Connell correlation,

$$\eta_{section} = 0.503(\mu\alpha)^{-0.226}$$

or the more recent Taylor-Duss variation,

$$\eta_{section} = 0.503\mu^{-0.226}\alpha^{-0.08}$$

.

Note that these equations only look at two factors: viscosity and relative volatility. No tray design parameters are included.

Hundreds of papers have been written and many researchers have worked on developing fundamental relationships for equipment efficiency. The reason so much work has been done is that most correlations do not produce useful results—that's why people keep trying. There are certainly general trends in how tray design and system properties affect efficiency. But the specifics of individual situations are especially important. If you can find published data on the system you are looking at, you are in luck. Otherwise, stick to the simpler correlations. They are just as accurate as the more

Mike: If an existing column has been simulated with a wrong efficiency, a redesign can still work out okay as long as the same, wrong efficiency is employed again. A better path, of course, is to correctly determine the true efficiency of the existing trays/packings especially via the use of the correct thermo (VLE) model and the use of good operating data, wherein the masses balance.

Karl: The challenge of simulation is to have your simulation model what happens in the field. Taking a simulation to the field and choosing the correct tray efficiency takes multiple years of experience. There are complex models for Refinery Vacuum Towers and other systems that have unusual behavior in the field. Ensure that your simulation models actual field behavior.

There is a large data base in the industry of estimated tray efficiencies. Many are based on SRK VLE data. Be conservative when choosing tray efficiency even when using SRK.

3. DESCRIBE THE CHALLENGES IN HARDWARE DESIGN.

Andrew: The two most challenging aspects of hardware design are (1) fluid flow under low pressure drop conditions and (2) vapor-liquid mixing and separation on trays.

Little pressure drop is available inside towers to drive fluid flow. Outside of external feed pumping to the tower, only gravity head is available for liquid flow. Also, the pressure drop inside the tower created by the vapor flow increases liquid hold-up. These issues apply to both tray towers and packed towers but has different implications for the different tower types. Understanding how and where pressure drop is created and where static head accumulates is critical to understanding equipment hydraulics.

Vapor-liquid mixing issues apply much more to trays than packing. Trays mix liquid and vapor and then separate them. Phase mechanics can range from froth to unclearly defined intermediates to spray, depending upon flow rates and physical properties. The hydraulics of a given system can cause entrainment (jet flood) and downcomer filling. Knowledge and understanding of these factors are critical to successful tray design.

Mike: Hardware design methodologies are by hardware companies. As a result, their designs often cannot be checked by the purchasers or selectors of the hardware. Purchasers of such hardware should insist that the purveyors prove that similar designs in similar applications have performed well subsequent to start-ups.

Karl: The hardware designer assumes that you have put the design safety margin in the simulation, so they rarely will add any additional design margin. I was consulting on a refinery crude tower revamp at the startup. I reviewed the hydraulic design – and the jet flood was designed at 30%. I would normally like the design to be above 50%. This low jet flood leads to low tray pressure drop, and lower tray efficiency.

Christian: The challenge in hardware design is to combine/optimize mass transfer equipment (packing/tray) with corresponding tower internals (distributors, collecting trays, and droplet separators) to ensure optimum performance of the whole system.

The challenges are in detail:

- Maximum capacity, especially for Revamps
- turn down ratio/flexibility,
- consideration of fouling, foaming, etc. where applicable
- Minimization of droplet entrainment at high capacity

A topic that is often underestimated is the individual mechanical design of the hardware, especially in revamp cases. A knowledgeable mechanical design engineer can minimize installation time with tailor-made, innovative mechanical designs. The use of existing weld-in parts and support rings and adapting the hardware design to the individual scenario can avoid or at least reduce cutting and welding in the column.

One trend we currently see in hardware design is that higher investment is accepted if the operating cost of the plant can be reduced, e.g., by installing more trays and reducing reflux.

4. DESCRIBE THE CHALLENGES IN INSTALLATION.

Andrew: The biggest challenge in installation is attention to detail. Installation is often done by contract workers who may have little experience in the field. The company involved is often the lowest-cost bidder. The inspecting engineer has a great responsibility to make sure the installation is performed correctly. He or she also has the last opportunity to catch errors. Inspect with more than just the eyes. Yes, make sure that equipment is mechanically installed as designed. But also look at the equipment and ask, “Does this installation make sense?” If it does not, do not arbitrarily change things, but get an explanation of why it is the way it is, and make sure that specialists have checked the situation.

Mike: Welding inside towers “always” needs to be followed by heat treating. Large-diameter towers require large, and heavy, parts that need to be able to fit through manholes.

Karl: One of the challenges of installation is that the rings are installed by the tower vendor, and they may or may not match the trays and tower vapor / liquid draws. In one installation we did not install the bottom tray, so we could match the liquid to the correct side of the reboiler baffle. It required some field modifications to the trays.

Christian: Depending on the size, installations can be huge and complex projects that require detailed planning and project management. Manpower, equipment, and hardware need to be at the right place at the right time, boxes need to be clearly labelled, etc. Sounds easy, but often creates problems.

At the same time there is a trend to reduce costs by using low-wage contract workers, vessels manufactured in low-cost countries, etc.

This result, for example, in much larger than usual column tolerances. In the last few years, we have developed (or had to develop) a variety of “adapters” and modification solutions to adjust hardware to accommodate unlevel support rings and large deviations from roundness tolerances.

It has also proven very successful to include us, your hardware supplier, in vessel inspections at the column manufacturer’s workshop before delivery on site. We are able to make detailed documentation of the real dimensions. This helps to reduce errors and installation time on site.

5. DESCRIBE THE CHALLENGES IN COMMISSIONING.

Andrew: Specific services may create unique problems at startup, but the biggest single issue in many services is bottoms level control. The problem tends to be worse in towers with trays. Level spans on tower bottoms are typically set by operational control ranges or by inventory requirements for downstream processes or equipment (such as pumps). Rarely, if ever, does liquid inventory in the level range take into account the liquid inventory on trays or packing. As vapor rates rise, liquid inventory builds up on trays, reducing the liquid flow to the tower bottoms. Control response is to either reduce bottoms product rate or to allow liquid level to drop. If vapor rate drops during startup, trays lose liquid to the bottoms. This creates a sudden increase in bottoms rate or bottoms level, similar to inverse response.

If the bottoms liquid level span is too small, the sudden liquid level changes can cause the control range to drop to zero or rise to over 100 in noticeably short periods. I have seen some towers with many trays in which the control range on the bottoms liquid inventory was as low as 25% of the tray liquid inventory.

Low liquid level has the potential to damage downstream equipment such as pumps or fired heaters. High liquid level can cause the reboiler return to massively entrain liquid up the tower. The massive entrainment can knock trays out or cause other damage.

Startup is a busy time. Many control loops are set to manual, and operators cannot monitor everything closely all the time. They tend to keep liquid levels high to protect downstream equipment. When an upset happens, a sudden additional increase in liquid level damages the internals of the tower. In my experience across all distillation towers, this is one of the most common causes of equipment damage during startup.

Mike: Technician/engineer shift changes often bring different start-up philosophies and communications between the shifts are often too incomplete.

There was a unit start-up where a senior engineer on the night shift had his own philosophy regarding how the new unit should operate. His instructions to the operators differed very significantly from those that were given to the day shift operators. The unit operation went very astray. Two days were lost getting the start-up back on track. Because of his actions, the senior engineer was fired after the unit finally reached equilibrium.

(I was the day shift engineer who complained about the performance of the night shift engineer, who was not my friend. The unit manager took me outside of the control room where I honestly thought that he was going to punch me or tell me to shut up or both. Instead, he told me that he had been afraid that the night shift senior engineer was going to lead the start-up astray. He thanked me for my future patience. He hinted that the night shift engineer was soon going to be "a goner.")

Karl: One of the commissioning challenges is that pumps need to run at least 60 to 70% percent of design and many feed pumps do not have minimum flow loops. Therefore, the feed to the tower goes from 0 to 60% quickly. When I design a feed loop I consider this and try to plan for this start up issue. Having a larger feed pipe and feed internals might be a good idea.

6. DISCUSS YOUR MOST DIFFICULT TOWER TROUBLESHOOTING.

Mike:

A. Alongside other engineers, I visited a certain liquid-liquid extractor 6 times over a 2-year period. Its capacity was lower-than-expected by 20%. Our team could not figure out what was wrong? Eventually, we provided a new extractor that had 20% more cross-sectional area than the original extractor. That new extractor worked well.

In a liquid-liquid extractor, there is a downward liquid velocity and an upward liquid velocity. In reviewing the pilot data, the correlations were based on hole velocities and not cross sectional tower area velocity, therefore the scale up was not correctly predicted. We corrected the correlations.

B. I climbed the outside of a tower on a cold and snowy night to read some gauges at the top of the tower. All of the vertical ladders were surrounded by cages. I wore all appropriate safety gear. When I shined my flashlight at one particular pressure gauge, a "black thing" came flying and screeching at my head. I fell several feet until my groin impacted one of the cage's horizontal support struts. I did not know that during the previous 24 hours, a crow had decided to build a nest behind the subject pressure gauge. To this day, I "hate" crows.

C. A liquid-liquid extractor was experiencing an abnormally high pressure drop. The top manhole was opened, and a flashlight revealed that the top two trays appeared to be "good as new." When we entered the tower, the trays crumbled beneath our feet. Acids had made the trays paper thin. The trays in the bottom of the tower proved to be plugged with the by-products of the acid attacks.

Karl: We simulated and designed a tower for a Thailand Chemical Plant. We put more than adequate safety in the design. Tower was then built and erected. Tower was commissioned, would make design capacity, but would not make design purities. For the next three days we tried every combination of temperature, pressures and flows to achieve the design purity. It was a very difficult three days.

I then choose to reinspect the tower, the startup seemed smooth, but we may have damaged some trays in commissioning, and we started planning to shut down the tower. The tower bottoms was the one with the purity and there was a feed / bottoms heat exchanger. We decided before we shut the tower down to sample the bottoms before the heat exchanger. It was on test. The feed / bottoms heat exchanger was leaking. We repair the heat exchanger and tower pass the performance test.

Henry: Copyright © by Henry Z. Kister, reprinted by permission.

In 1988, I was leading a team troubleshooting a chemical tower that was revamped to achieve a small increase in capacity (0%-4%) by increasing the trays hole areas from 8.5% to 13% of the active area. In addition, radiuses were added at the bottom of the downcomers to smoothen exit of liquid and reduce downcomer backup. Strangely, instead of gaining capacity, it lost 5% capacity.

Before 1992, the mechanism of vapor cross flow channeling (VCFC, Kister, H. Z., K. F. Larson, and P. E. Madsen "Vapor Cross Flow Channeling on Sieve Trays: Fact or Myth?" Chem. Eng. Prog., November 1992) on sieve and valve trays was not recognized by the industry. I inquired around, and everyone was telling me that the capacity reduction should not have happened. There was one lead, as yet considered unlikely, that suggested that VCFC could be taking place, a mechanism well-known in bubble-cap trays. Good troubleshooting investigations follow all leads, likely or unlikely. So, I did.

In VCFC (Figure 1) the hydraulic gradient on the tray induces preferential rise of vapor near the outlet and middle of the tray and forms a vapor-deficient region near the inlet to the tray. The high vapor velocities near the tray outlet step up entrainment, while the low vapor velocities near the tray inlet induce weeping. Interaction between adjacent trays (Figure 1) accelerates both the outlet entrainment and the inlet weeping. The result is excessive entrainment and premature flooding at the tray middle and outlet, simultaneous with weeping from the tray inlet, accompanied by a loss of efficiency and turndown. The theory was that increasing the tray open area on the trays led to the onset of VCFC and consequent premature flooding on the trays.

The challenge was to test this theory. Gamma scanning was the most promising technique, but no one had experience with applying gamma scans for diagnosing VCFC, nor an idea how to do it. So, it was decided to brainstorm together with Tru-Tec who had extensive gamma scanning expertise. Understanding the challenge, Tru-Tec lined up their three top experts, and together with the client's top experts and the author met in Tru-Tec LaPorte's office. The meeting started about 3 PM, lasted for a few tedious hours. A lot of ideas and stimulating brainstorming were tossed around, but no breakthroughs. People began to wonder if any resolution will come out.

Just then, one of the Tru-Tec experts spoke. "Looks like we have brainstormed for a few hours and got nowhere. Maybe we are doing it all wrong. So, let me suggest another way". Then he continued "Not far from here there is an excellent Mexican Restaurant called Don Key. They have fabulous food, but what is more important, is they have an excellent beverage called Donkarita. It is similar to Margarita, except that it kicks like a donkey. After one or two of these, we may get more creative and come up with a winning idea". His motion was unanimously and enthusiastically accepted. Not only did we enjoy a delicious meal and donkaritas, but we left with a plan that everyone was happy with – and laid the foundations to the quantitative scanning that is used to troubleshoot channeling until today (Kister, H. Z., "Use Quantitative Gamma Scans to Troubleshoot Maldistribution on Trays", Chem. Eng. Progr., February 2013).

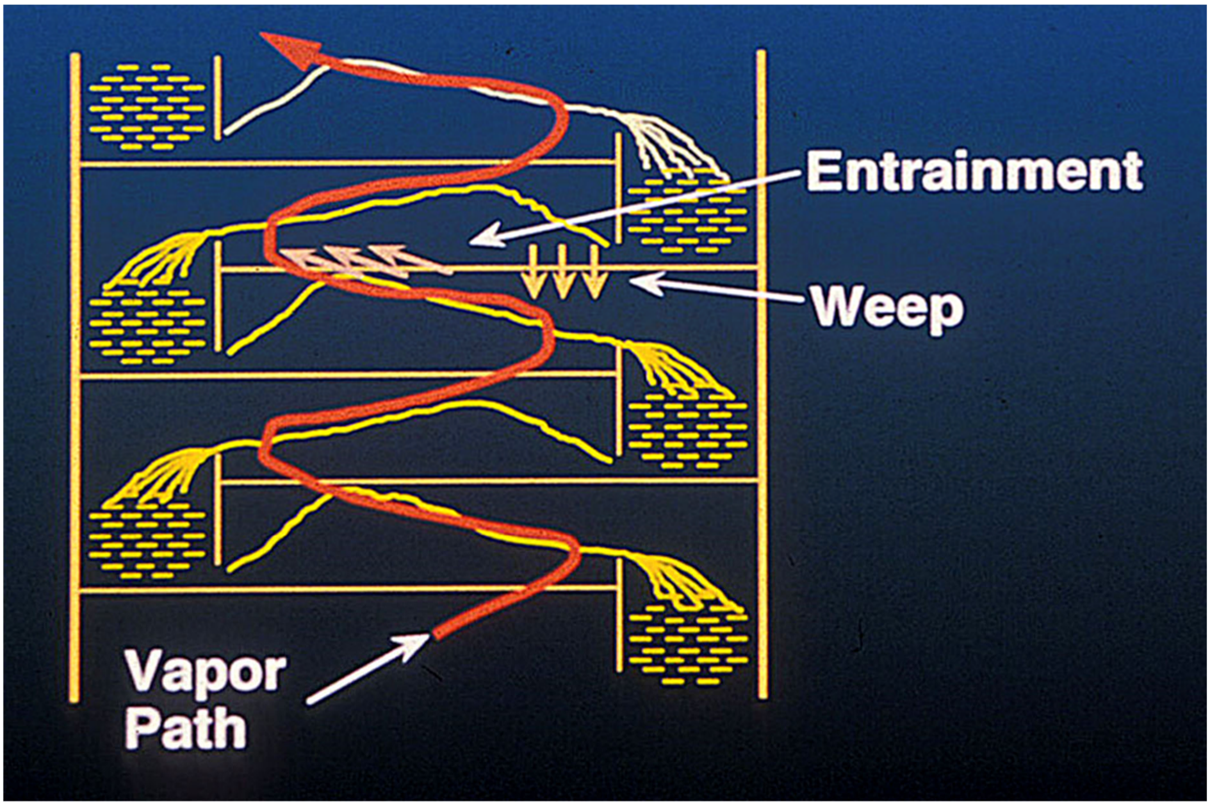

Figure 1. Vapor cross flow channeling

The winning idea was to scan three different chords along the flow path length both under flooded and unflooded conditions, and to apply quantitative analysis to derive froth heights, froth densities, and liquid heads. From the liquid heads hydraulic gradients can be inferred. The details are described in the above reference. The results are shown on Figure 2. In Figure 2, all the tray dimensions are drawn to scale. Tray liquid head values derived from quantitative analysis of gamma scans are plotted so that a liquid head of half the tray spacing is plotted as a point on the tray above. Zero liquid head is plotted as a point on the tray floor.

In the unflooded scans (Figure 2a), hydraulic liquid gradients are flat or slight, sloping from tray inlet to outlet. Upon flood initiation (Figure 2b), the hydraulic liquid gradients on the odd trays become very steep, especially between the middle and outlet. The even trays (trays 2, 4, and 6), show large uniform hydraulic gradients stretching from inlet to outlet at flood initiation.

This intensification of the hydraulic gradients strongly supported VCFC as the root cause of the observed premature flood. From an unlikely hypothesis, VCFC became the leading theory. Based on this diagnosis, minor modifications were made to tower auxiliaries that permitted raising tower pressure. With VCFC inducing a premature entrainment flood, raising tower pressure reinstated the lost capacity. Prior to the modifications, raising the pressure did not improve capacity because the trays were bottlenecked simultaneously by downcomer backup (which was debottlenecked by adding the downcomer radiuses) and by entrainment flood.

The takeaway, when dealing with a challenging problem, a change of environment from a four-wall meeting room to a more pleasant surrounding may help usher in a creative solution.

Christian: In the beginning, it is often very difficult to get a complete picture of the situation, especially if you are not on site. If this is not possible, a detailed discussion with engineers and operators is often very helpful to get more insight. Photos can also be invaluable.

Expect the unexpected. I have seen lost shoes and helmets inside columns, but also undetected problems with measurements can be challenging.

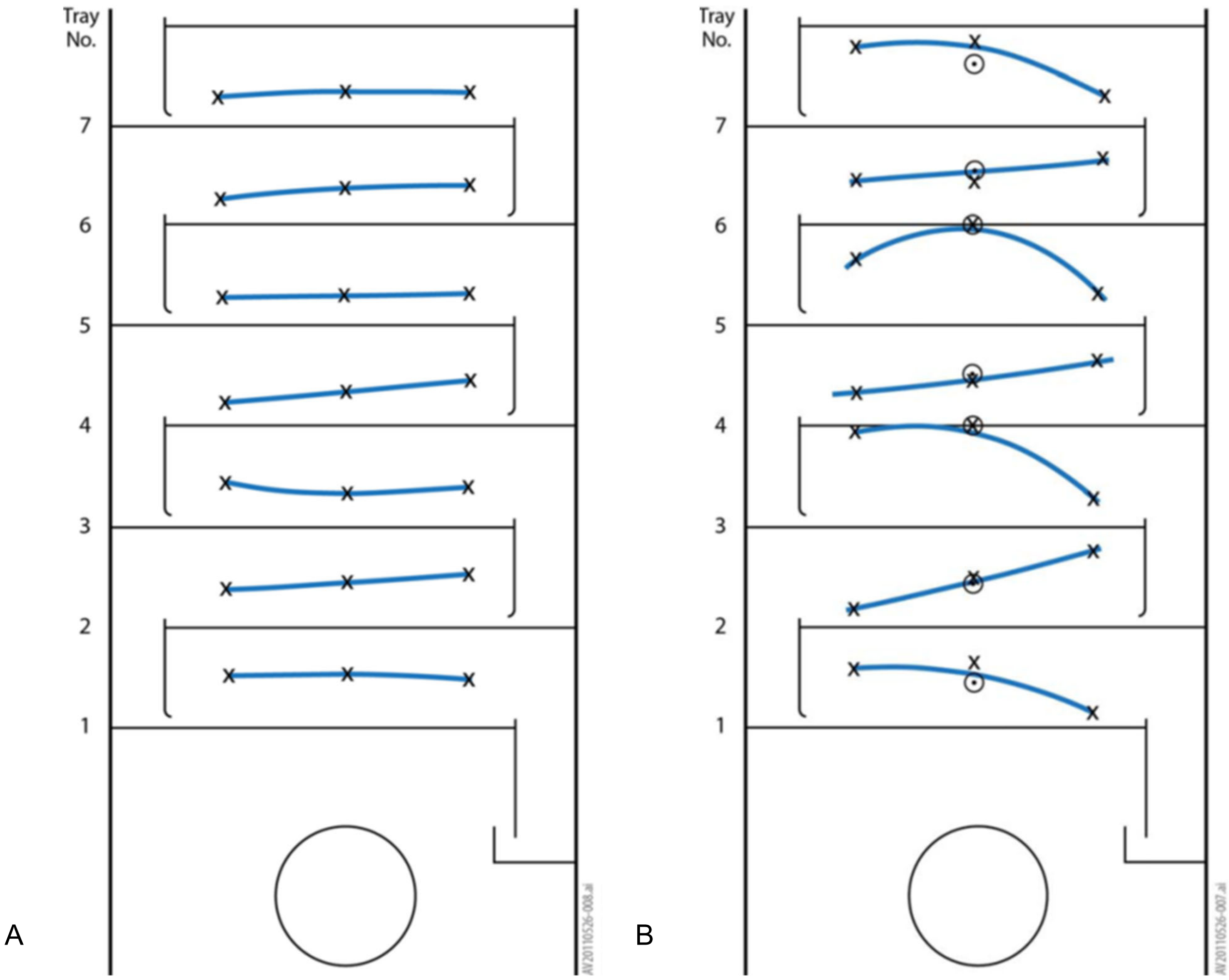

Figure 2. Clear liquid heights derived from three gamma scan chords along the flow path length plotted on a to-scale column sketch. Tray liquid head values derived from quantitative analysis of gamma scans are plotted so that a liquid head of half the tray spacing is plotted as a point on the tray above. Zero liquid head is plotted as a point on the tray floor. Hydraulic gradients can be inferred from the diagram (a) Unflooded (b) Flooded.

7. DISCUSS YOUR MOST UNUSUAL TOWER TROUBLESHOOTING FINDING?

Andrew: On one troubleshooting assignment, I received P&IDs the day before the plant visit. The service was a specialty product made in an exceptionally large batch distillation column that operated at deep vacuum (<1 mmHg, <0.133 kPa). After some time looking at the P&IDs I was completely stunned. They were so odd I took them to a colleague to verify I was reading them correctly. The next day in the field I confirmed the P&IDs were accurate. The column had no reflux. In fact, the reflux line did not exist at all.

(Years later I learned that some industrial units in the late 1800s and early 1900s were built without reflux. They depended upon heat loss to provide the internal liquid in the tower. For an example, look at the Roger's still in his 1871 patent. The top section of the tower gets all its reflux from heat loss.)

The unit requiring troubleshooting was built in the 1950s and converted to its current use in the 1980s. While it was older, it was not that old. It should have had reflux.

Adding a reflux line, modifying the internals in the top of the tower to handle that reflux, and adding flow control led to substantially improved separation, to the amazement (and satisfaction) of the plant.

Mike: In one troubleshoot in Asia, new hardware was being employed in a tower to increase the tower's capacity. Shortly after start-up, the Asian company complained that the tower was not working. After flying immediately to Asia and when I arrived at the plant, I determined that the customer had never even tried to increase the feed rate to the new feed rate. With me present, they raised the feed rate and the tower performed well.

Karl: There are many great stories in the industry of unusual tower troubleshooting.

A. A new small packed column would not meet specifications. The tower was shut down and an inspector was brought in to inspect the column. The small amount to random packing was delivered in kitchen sized garbage bags. The installation team had carefully installed each garbage bag of packing inside the column. The inspector instructed the team to take out the garbage bags and just install the packing in the column. The column then met specifications.

B. A Low Pressure DeMethanizer was upgraded with a larger new stainless steel reboiler. When welding stainless steel there is an argon gas blanket requirement. The welders will install plugs to reduce the amount of argon gas required. Welding was completed and unit commissioned but reboiler would not function properly.

Tower was shut down and inspected. Welders did not remove the plugs as required. Plugs were removed, tower recommissioned and then functioned properly.

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As a Startup Superintendent at ICI Australia in the late 1970's, I was in charge of starting up a new unit in the Olefins plant that produced polymer grade propylene. The main equipment was a C3 splitter column, which at the time was the biggest tower in Australia, is in operation until today, and can be seen as one lands in Sydney airport. The auxiliaries were a reboiler heated by waste heat from the olefins quench water system, a water-cooled condenser, and a reflux drum. Due to the very high cooling load, a new cooling water system was added to service the new unit alone, mostly the C3 splitter condenser. The cooling tower was some distance away from the unit, and the large cooling water pipes were mostly underground to avoid the need of a pipe rack. The equipment was installed, and I was in the process of commissioning and testing the unit, well before hydrocarbons were introduced.

The cooling water system startup was smooth and uneventful. Everything was fine, water flowed over the cooling tower, the pump was pumping with a discharge pressure close to design. As the process side was not commissioned yet, no heat was exchanged.

While taking a walk to closely watch the system, I noticed the cooling water flowmeter (Figure 3, the flowmeter was in a horizontal section of pipe, with ample pipe diameters before and after to give a reliable reading) reading zero. The dial was actually hitting the stop pin below the zero mark. I called the plant instrument foreman, who checked the transmitter.

"This worthless transmitter is kaput." He said (using another word for "kaput"). All that Engineering have been giving us in this plant is junk (again, using another word for "junk"). Why don't you tell Engineering to provide us with instruments that work?"

This statement was true. Many of the instruments that were supplied were inoperative. I passed the message to Alan, the Engineering instrument engineer.

"Again? Our apologies. Let me look into it"

A couple of hours later he returned.

"I checked the transmitter. There is nothing wrong with it. It works".

"So why is it reading zero? The pump is pumping flat out, the water flowing, and had the transmitter been any good, it would not read zero".

"Are you sure your cooling water is not going backwards?" he said.

I almost choked, then stared right at him.

"Alan, if anybody else came up with this nonsense, I would have thrown him out of this door. But in your case, I have a lot of respect for your expertise, which incidentally, with this kind of comments I am about to lose. But just for old times' sake, I will get this nonsense out of your head".

"How?" he asked

"Let's go". We climbed up the condenser platform. I pointed at the butterfly valve at the condenser outlet. "I will shut the valve and open the 1" vent valve (Figure 3). What you will see is a water jet shooting up 30 ft in the air, which will disprove your nonsense idea".

"Good deal" he said.

I shut the butterfly then opened the vent. A water jet shot up 30 feet in the air, just like I expected. As I was getting ready to shut the valve the jet plunged, and plunged, until it stopped. I shook my head in disbelief.

"This is why the needle was hitting the stop pin. It was reading negative and tried to tell us something", he said.

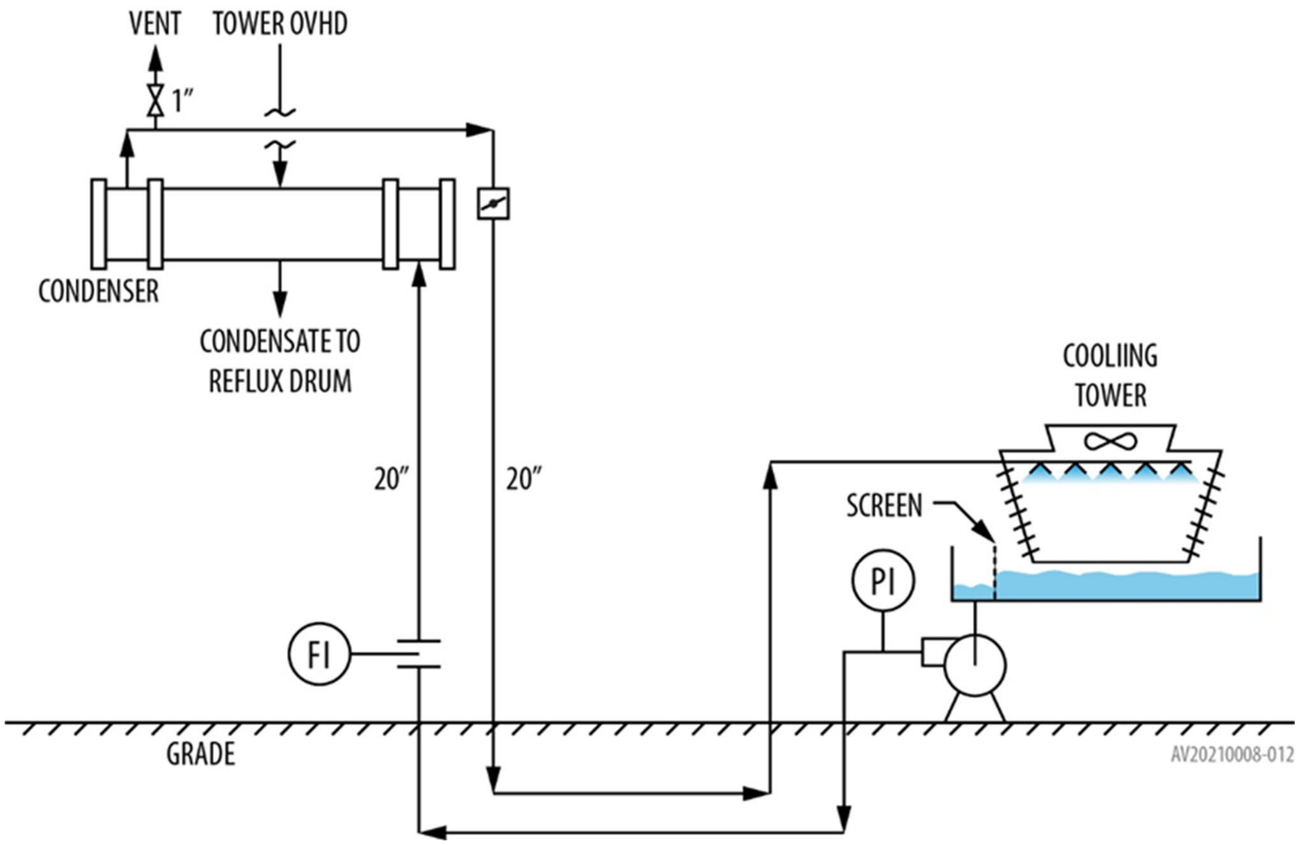
I reopened the butterfly, the repeated the test. Just in case I was dreaming. The same happened again. It was not a dream.

"I am taking it back, Alan. My respect to your expertise work has doubled. This is an amazing catch".

I instructed removing the dirt that covered the underground pipes. Sure enough, there was an incorrect pipe connection underground (Figure 4) that was overlooked by the construction inspectors. As a result, the cooling water flowed backwards through the condensers and the meter. Being underground, no one suspected. Fortunately, this was identified before hydrocarbons were introduced. It was the one flow meter and a top notch instrument engineer that made all the difference.

The takeaway, never disbelieve an instrument. It may be trying to tell you something. You do not need to trust it but listen to it. Always thoroughly check out suspicious readings.

Figure 3. Cooling water system in propylene purification unit



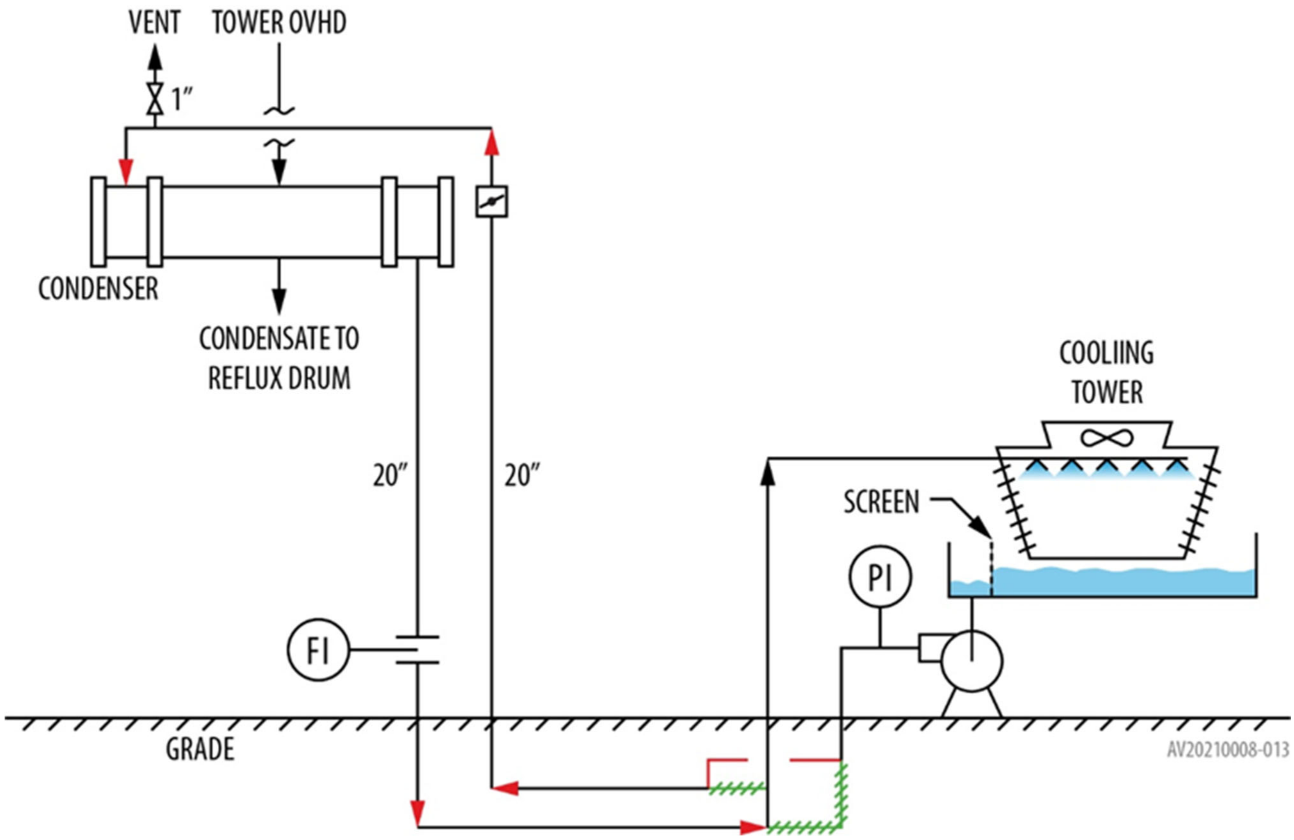


Figure 4. Cooling water system in propylene purification unit as installed.

8. WE HAVE 1ST GENERATION, 2ND GENERATION AND NOW 3RD GENERATION RANDOM PACKING. WHAT ARE THE STRENGTHS AND WEAKNESS OF EACH?

Andrew: Cost drives the selection between these packing types. All work well; design and operating characteristics of all types are well understood. As a general rule, the later the generation, the higher the capacity or efficiency for the same column cross section. There are subtle differences that apply to specific cases, but these must be examined on an individual basis.

For an existing column, buy the cheapest that meets your combination of required stages and capacity. For a new column, 3rd generation types typically reduce column diameter and total capital, making them the preferred option despite the premium price. However, the design team should always check. In some situations that premium price can make the overall plant more expensive and sway the decision towards 1st generation packing.

Karl: Like most things in life, there is no one size fits all. Distillation is similar. There is the best choice for each application. Many times, the vendor will try to propose their highest profit margin packing, when it is not the best option for the application.

1st Generation

Typically, thicker metal of construction leading to higher crush strength and corrosion resistance. For quench tower that are heat transfer limited, 1st Generation Cascade Mini Rings may be the best choice.

Later Generations

They typically will be thinner metal of construction leading to lower crush strength and corrosion resistance. They will have lower pressure drops, but higher capacities and efficiencies.

Christian: 1st and 2nd Generation Packing are more or less industrial standards:

- easy to replace,

- variety of suppliers,

- well-known (implemented in several software packages),

- larger wall thickness (corrosion resistance).

Disadvantages are normally high pressure drop and low capacity.

3rd or 4th Generation Packing are optimized for

- low pressure drop,

- high capacity and

- sufficient mass transfer (compared to 1st or 2nd generation type packing with same surface area) by open structures and thin coil materials, resulting in wide operation limits. Disadvantages are thin material, which is more sensitive to corrosion, and higher investments costs due to difficulty in getting “equivalent” products. Typically, these packing types also provide less holdup. In some rare cases this might be a problem, such as when the process asks for higher liquid residence times.

9. WHAT ARE SOME OF THE MORE DIFFICULT APPLICATIONS TO DESIGN?

Andrew: Applications that include a mix of challenging process conditions and device hydraulics can be difficult to design. Small errors in estimates of vapor-liquid equilibrium can have large consequences for design or performance of super fractionators (100+ stages required) or systems requiring extremely high purities. Reactive distillation can be particularly challenging. Reaction pathways and kinetics interact with composition profiles, making it difficult to predict holdups that will provide an effective space velocity. Making changes to systems with extreme changes in vapor and liquid rates can be challenging, since different equipment designs must be placed in placed in different parts of the column. You need to be able to predict these locations correctly. One more: when working with exceptionally large diameter towers, one needs to pay particular attention to both fluid flow and distribution as well as the equipment’s structural details.

Mike: The correct efficiencies associated with alcohol separations are difficult to ascertain and to utilize. Refinery columns often require numerous transitions, feeds, draws, and side strippers requiring attention to detail.

Karl: In a new column one can put in adequate safety margins. In revamps, often the safety margins are challenged. In one C3 Splitter revamp, which is very well understood VLE, the vendor was very aggressive in guaranteeing a capacity and the capacity was not reached in the performance test run. Even in well understood systems performance test runs can fail.

In polar and foaming systems, the design of the actual capacity can be challenging. There have been many failures in foaming systems reaching design capacities.

10. WHAT ARE THE ADVANTAGES AND DISADVANTAGES TO DIVIDED WALL COLUMNS?

Andrew: The potential advantages of a divided-wall column are lower capital and operating costs. The major disadvantage is that the column is more complex, and both design and control are more challenging. The entire possible range of feed and product compositions must be well understood. The equipment for controlling the liquid-vapor splits must be designed to accommodate the spectrum of composition and rate ranges. Alternatively, a design margin can be incorporated to deal with feed and product variations. However, allowing for extra reflux or extra boil up will reduce the capital and energy savings that made the option attractive in the first place.

One other minor disadvantage should be mentioned. In a divided-wall column, the total reboiler duty will drop, but the duty at the highest temperature level may go up. This is because 100% of the reboiler duty is now at the equilibrium temperature of the heaviest product instead of being split across multiple products. In some heat integrated systems this may negate much of the potential energy savings.

Christian: Typical advantages are better energy efficiency and lower investment because multiple separations are done in one column. The disadvantages are that control and design are more complex. As a consequence, troubleshooting may become more complicated, just because more parameters are involved.

11. WHAT ARE SOME OF DISTILLATION PROCESS CONTROL ISSUES?

Andrew: The biggest issue in distillation control is interaction between control loops and loop delay time. The simplest loops, overhead liquid level control and bottoms liquid level control are typically, but not always, straightforward. Beyond that, composition controls can be overly complex and create the largest interactions. Controlling composition really means controlling the entire composition profile throughout the column. However, the composition controls get feedback from only one point on a column. As a result, all SISO (single-input, single-output) control configurations suffer from loop interaction and lag. It is possible to set up controls to meet composition specifications on all product streams. This is known as two-point control, or two-product control, and is such a difficult control problem that extremely few industrial units even attempt it.

Advanced control logic can reduce control interaction and delay, but rarely eliminates it. Selecting a good basic control configuration greatly eases the task of an advanced controller. At the other extreme, selecting a poor basic configuration will make the task of an advanced controller nearly impossible.

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Experience is the best teacher and learning from the past is the best way for solving today's problems and avoiding tomorrow's. To this end, we conducted a survey of the case studies of malfunctioning towers (Kister, H. Z. "What Caused Tower Malfunctions in the Last 50 Years", Trans. IChemE Vol. 81, Part A, p.5, January 2003).

This survey was updated in Kister's book "*Distillation Troubleshooting*", (John Wiley & Sons, 2006) and recently the control issue in Kister's IChemE-sponsored on-line course "Troubleshooting Distillation Controls" https://www.kisterdistillationseminars.com or https://www.icheme.org/career/training/online. Below is an overview of the prime distillation control issues based on this survey.

Our survey put control issues in the 3rd spot on the distillation malfunction list, to be surpassed only by fouling/plugging and problems in the column bottom sump. The prime control issues can be categorized into three classes: control system assembly issues, temperature and composition control issues, and condenser, pressure, and reboiler control issues.

A related category is misleading instruments. A control system is only as good as its instrumentation. This category is discussed at length in "*Distillation Troubleshooting*" and excluded here due to space limitations.

Most control system assembly issues case histories come from chemicals and olefins/gas towers, where splits are usually much tighter than between petroleum cuts in refinery towers. Over half of the reported control system assembly issues stem from violation of three basic principles. First, violation of the material balance control principle. Special difficulties have been encountered when adopting the material balance control to towers with side draws. The good practices for this situation were described by Luyben's classic article ("10 Schemes to Control Distillation Columns with Sidestream Draw offs", ISA J., 13 (7), p. 37, 1966.). The second is violation of what is known as "Richardson's rule" which states "It is not a good idea to control a level on a small stream." The third is attempting to simultaneously control two compositions in a two-product column without decoupling the interference between them.

Some case histories address the drawbacks of some of the common material balance control schemes. Examples include the slow dynamic response of a scheme that controls tray temperature by manipulating reflux in large tray towers, or the inverse response experienced with the scheme that controls the bottom level by manipulating the reboiler steam.

Two approaches have been successful in improving control system assembly: the traditional approach diagnoses deficiencies and eliminates them by judicious changes to the control system. The alternative, more modern way is to replace the conventional control scheme by advanced controls, using models and statistical process controls.

Most of the temperature/composition control malfunctions come from chemicals and olefins/gas towers, where splits are usually much tighter than between petroleum cuts in refinery towers. The major composition control issue has been correctly identifying the best temperature control tray. Application of the excellent method by Tolliver and McCune ("Finding the Optimum Temperature Control Trays for Distillation Columns", InTech 27(9), p.75, 1980) has effectively dealt with this issue. Still, there are some situations where finding a satisfactory temperature control can be elusive. Other key issues have been achieving successful analyzer control and obtaining adequate pressure compensation for temperature control. With analyzers, the main problems have been measurement lags and on-line time. Modern analyzer controls are often associated with advanced controls and have grown in significance in recent years.

Two approaches have been successful in improving temperature and composition controls. The traditional approach uses solutions such as defining the best temperature control tray and cascading analyzers onto temperature controls. A more modern way is to use virtual analyzers, based on model calculations from tower measurements, and using statistical process controls. Useful tricks, such as pressure correction to the temperature, or using an average temperature, have been incorporated with both approaches.

More pressure, condenser and reboiler control case histories come from refinery than from chemical towers. One reason for this is refiners' extensive use of hot vapor bypasses. About one third of the reported pressure and condenser control case histories are associated with hot vapor bypasses, practically all in refineries. There is little doubt that this is potentially the most troublesome pressure control method, mostly due to poor configuration of hot vapor bypass piping, evolving from poor understanding of its principles. These principles have been detailed in recent papers by Kister and Hanson ("Control Column Pressure via Hot Vapor Bypass", Chemical Engineering Progress, February, p. 35, 2015) and Kister ("Flooded Condensers Controls: Principles and Troubleshooting", Chemical Engineering, January, p.37, 2016). When configured correctly, our experience is that hot vapor bypasses are seldom troublesome.

Another troublesome pressure/condenser control is by cooling water throttling. It has induced low cooling water velocities and high outlet temperatures, leading to fouling, corrosion, and instability, even boiling of cooling water. This is more of a problem in chemical towers, when venting an inerts stream from the reflux drum requires a temperature control at the drum to avoid product loss or product contamination. A third issue is low points that accumulate condensate in the pressure-controlled vapor product lines with the condensate backpressuring the column.

Reboiler and preheater controls have been troublesome in both refinery, chemical, and olefins/gas towers. Temperature control problems with preheaters are common, often due to disturbances in the heating medium or due to vaporization in the feed lines. Almost all the reboiler case histories reported involved a latent-heat heating medium. Hydraulic problems such as "stall" are common when the control valve is in the steam/vapor line to the reboiler, while loss of reboiler condensate seal is common when the control valve is in the condensate lines out of the reboiler. Such control problems often lead to column capacity limitation, hammering, fouling, and corrosion. The variety of solutions, well-illustrated in "*Distillation Troubleshooting*", is a tribute to the ingenuity and resourcefulness of engineers, supervisors, and operators.

Advanced controls have their own problems. One issue is updating multivariable controls (MVC), which can be troublesome when the process train changes, especially if the MVC simultaneously optimizes an entire unit rather than individual towers. Another issue is the response to bad measurements. So far, the number of reported case histories of troublesome advanced controls has been low.

12. WHY ARE SOME CONDENSERS PLACED AT GROUND LEVEL, AND DO YOU RECOMMEND THIS DESIGN?

Andrew: Consistent with good equipment design and minimizing overall plant cost, there is no problem with having condensers at ground level. The biggest reason for putting them there is to reduce their capital and maintenance costs. Less structure is required when the condenser is placed close to the ground. Access for maintenance is easier.

There is one operational consequence to consider. Putting the condenser close to the ground normally necessitates a vertical rise to the overhead reflux drum. The drum is installed at a higher elevation to make the installation of pumps cheaper and easier. Therefore, the condenser operates at a higher pressure than the drum because of the static head between them.

For most applications with total condensers, the liquid will be sufficiently subcooled to prevent vaporization in the line to the reflux drum. Partial condensers, on the other hand, may have problems with unstable two-phase flow regimes in the outlet line to the drum.

Having a condenser at ground level is not a good choice when the higher pressure created by the static head increases upstream costs. One such application is a fluid catalytic cracker (FCC) main fractionator overhead system. The extra liquid height can create significant cost or capacity problems for the upstream FCC air blower. On larger, modern units the air blower is an axial compressor. These compressors have limited head capability and small changes in the downstream pressure profile caused by the static head between the exchanger and the reflux drum could have large consequences.

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Money and environment. The hot vapor bypass scheme permits mounting water-cooled total condensers at ground level instead of on a platform above the reflux drum. Locating large cooling water condensers at ground level eliminates the need for massive condenser support structure and for piping cooling water to high elevations and provides easy access for maintenance. Keep in mind that for a 7-meters long elevated condenser, a 20-meters long platform is needed because the tubes need to be pulled out at the turnaround. With the hot vapor bypass scheme and ground-level condensers the piping is simple, the control valve is small, the response is fast, and there are no concerns about drawing vacuum in the cooling water return pipes at the high elevations. These advantages can translate into handsome savings in steelwork, platforms, trolleys, and maintenance. These savings can be major in large installations, especially where a battery of condensers rather than a single exchanger is used. Further, with the increasing environmental awareness, the elimination of the massive platforms, steelwork, and piping reduces the carbon footprints of the plant.

As stated in my reply to 11 above, about one third of the reported pressure and condenser control case histories are associated with incorrectly configured ground-level condensers controlled by hot vapor bypasses, evolving from poor understanding of its principles. These principles have been detailed in recent papers by Kister and Hanson ("Control Column Pressure via Hot Vapor Bypass", Chemical Engineering Progress, February, p. 35, 2015) and Kister ("Flooded Condensers Controls: Principles and Troubleshooting", Chemical Engineering, January, p.37, 2016). When configured correctly, our experience is that hot vapor bypasses are seldom troublesome. Figure 5 shows samples of incorrectly and correctly-configured arrangements. Some key considerations are:

This method is only suitable for total condensers, i.e., no vapor product. It should not be used for partial condensers, as it depends on subcooling.

Correct piping is mandatory. Liquid leaving the condenser is subcooled, and dew point vapor will collapse onto it causing instability and hammering. Figure 5a shows a configuration that I have seen more than two dozen times, none of which worked. The bypass vapor must enter the vapor space of the reflux drum (Figure 5b) with no previous contact with the subcooled liquid. The bypass must be free of pockets where liquid can accumulate; any horizontal runs should drain into the reflux drum. Most important, liquid from the condenser, as well as any other subcooled liquid streams like the reflux pump minimum flow recycle, must enter the reflux drum near the bottom of the drum (Figure 5b).

Operation is likely to be troublesome if the drum liquid surface is agitated. Agitation may be due to high-velocity impingement of the hot vapor jet on the liquid surface, due to upward-directed subcooled liquid jet reaching the liquid surface, as well as other causes. Agitation of the liquid surface can be avoided by judicious baffling (Figure 5b).

Because of the liquid leg between the condenser and the drum, sudden reduction in drum pressure can rapidly suck the liquid right out of the condenser, causing a major upset. There is also the possibility of U-tube oscillations. Both issues can be mitigated by adding a throttling valve in the liquid leg between the condenser and the drum (Figure 5b).

This technique can suffer from interaction between the drum and the condenser liquid levels. To minimize the interaction, the pressure controller should be tuned much tighter than the drum level controller. This can be an issue if the reflux drum is small, and the level controller needs to be tuned fast to avoid overflow or loss of level. This scenario is uncommon; we have encountered this situation only once, but another case was reported in the literature.

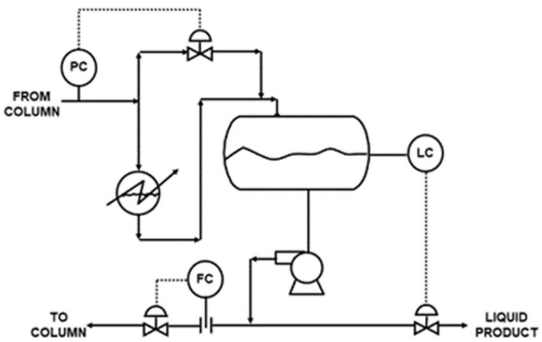
Because of the liquid leg between the condenser and the drum, non-condensables accumulate in the condenser and need venting from their accumulation points. If a vent line is absent, instability and capacity bottlenecks are likely to result.

Leakage of vapor through the bypass valve at the closed position can substantially reduce condenser capacity. Under sizing of the by pass control valve may lead to inability to keep the tower pressure up in cold winter days when the drum is not insulated. In some cases, the reflux drum vapor space may need to be insulated to minimize interference from rain and snowstorms.

We have seen cases when the hot vapor bypass control valve is manipulated by the drum pressure instead of the tower pressure. We do not recommend this. Dynamically, this control is inferior because the vapor volume in the drum is much smaller than in the tower and more variable in response to ambient changes.

There are other less common issues described in the cited articles above.

A



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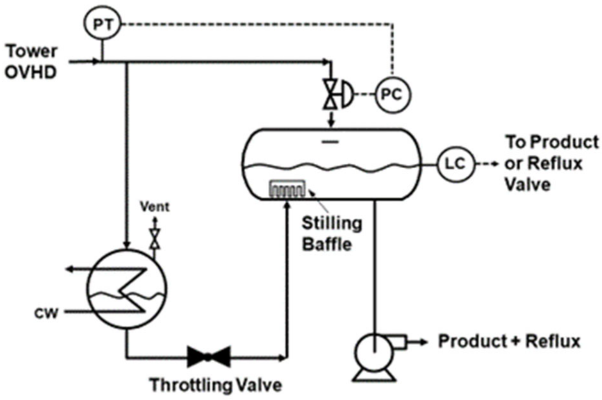

Figure 5. Hot vapor bypass controls with ground-level condensers (a) Incorrect arrangement, hot bypass vapor collapsing onto subcooled liquid, leads to instability, poor control, hammering (b) Correct arrangement, vapor going to vapor, liquid to liquid, no contact between hot vapor and subcooled liquid.

In summary, this method has major cost and environmental advantages in large water-cooled total condensers, and we would highly recommend it in the correct applications. To make it work it is imperative to understand its principles and configure it properly. When configured correctly, our experience is that ground level condensers with hot vapor bypass controls are seldom troublesome.

13. WHAT ARE THE ADVANTAGES AND DISADVANTAGES OF MULTIPLE DOWNCOMER TYPE TRAYS?

Andrew: I will only address generic trays with multiple downcomers here, not specialized designs used in proprietary tray configurations.

Trays with multiple flow paths add extra downcomers to extend the weir length and reduce liquid load per length of weir. This reduces tray pressure drop and is the most important reason for using them—their ability to handle high liquid loads. The reduced pressure drop also increases their vapor handling capacity.

The main disadvantage of multiple-pass trays is that they increase the minimum liquid rate as well. Blowing, which reduces efficiency, can occur at low liquid rates. As weir length increases, so does the liquid rate at which blowing starts. Multiple-pass trays generally have less capacity range than single-pass trays.

Two-pass trays are relatively straightforward to understand and design. Trays with three passes or more include non-symmetrical sections, requiring a more detailed understanding of hydraulics to get constant vapor-to-liquid ratios in all sections. Design and analysis of trays with three or more passes should be left to specialists.

Multiple-pass trays are also used to reduce the flow path length across the tray. However, if the only purpose is to reduce flow path length, a stepped tray with an intermediate weir is often more effective.

Mike: Multiple downcomer type trays have been employed very successfully for over 50 years in many applications and at exceptionally large diameters. Efficiencies are sometimes slightly less than with crossflow trays. Multiple downcomer trays employ significantly more metal and therefore cost more. Their capacities, however, are often 20% higher

Karl: The challenge of these types of trays is estimating the correct efficiencies. They can be at least 10% less efficient due to the low path flow length. But they can be installed on lower tray spacing and end up with higher overall tower efficiency, with the advantage of higher capacity. Rarely do you get higher capacity and efficiency.

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These are excellent high-capacity trays for the correct applications. Their capacity enhancement is a result of three factors:

1. The addition of downcomers increases weir lengths, thus lowering the weir loads (the weir load is the quantity of liquid flowing per unit weir length). This shifts the operating point to the left on Figure 6. Towers that have high weir loads with conventional trays (especially > 80 m3/h/m of outlet weir) can significantly gain capacity by the additional weir length. The designs usually aim at bringing the towers to the maximum region (20-30 m3/h/m of outlet weir). Towers that operate at low weir loads (< 50 m3/h/m of outlet weir) have little to gain from this type of trays.

2. These trays use truncated downcomers, terminating just over about halfway through the tray spacing. Liquid issues from holes or slots at the bottom of these downcomers. This allows perforating what normally would be the dead seal areas under the downcomers and gains active area.

3. These trays usually use small perforations, typically 5 mm, and seldom more than 7. Smaller holes gain capacity, with 5 mm holes typically giving 7% higher capacity compared to 13 mm holes at weir loads of 20-30 m3/h/m of outlet weir.

4. The capacity can be further enhanced by incorporating slots that impart a forward push to the liquid, together with anti-jump baffles that catch this liquid and divert it into the downcomers.

The disadvantages of these trays are:

1. The additional downcomers reduce the flow path length, which lowers tray efficiency. The efficiency of this type of tray is typically of the order of 10% less than conventional trays. This would normally mean a stage loss; however, the capacity gains are so high, that the multi downcomer type trays can be placed at tighter spacing and still give substantial capacity gains. Alternatively, the tray spacing can be optimized to give more capacity, more stages, or both.

2. The small holes make such trays prone to plugging. Such trays are not a good application for fouling services.

3. There are some troubleshooting issues with multiple downcomer trays, mostly at lower pressures (< 10 barg). Since the downcomers have no static seals, they are prone to vapor to breaking in ("blowby"). In addition, with the large number of downcomers, metering the correct amount of liquid to the various panels becomes an issue, and can lead to maldistribution and dry regions. Dry regions often lead to or aggravate blowby issues. Good initial liquid distribution is critical, and we have seen problems with feed and reflux entry. These troubleshooting issues can cause large reductions in tray efficiency, forcing operator to use excess reflux and reboil, and often run into a capacity limit. These issues are discussed with tips on how they can be diagnosed using gamma scans in Kister's article ("Gamma Scan Quantitative Analysis Can Diagnose and Mitigate Channeling in High-Capacity Trays with Truncated Downcomers", Chem. Eng. Progr., April, p. 45, 2013).

4. Due to the ease of losing the downcomer seal, the turndown of multiple downcomer trays is relatively low, typically about 1.5, compared to 2 for sieve trays, 2.5 for fixed valves, and 4-5 for moving valve trays.

5. Multiple downcomer trays are very difficult to inspect at the turnaround. This again emphasizes the need to avoid them in fouling services. When metallurgists need to inspect the tower wall for corrosion at turnarounds, this can be prohibitive.

6. They are expensive, especially when there is a need to change tray spacing, which may require special techniques to avoid or minimize welding to the tower shell. For a new tower, the expense can be offset by reducing column height or diameter.

7. Multiple downcomer type trays are far more sensitive to out-of-levelness than conventional trays. Many towers have experienced severe efficiency loss because

the trays were not installed level to the tight vendor specs. Sticking to these specs can be challenging in tall towers, especially during windy conditions when the tower sways, and waiting for the wind to drop may prolong the turnaround.

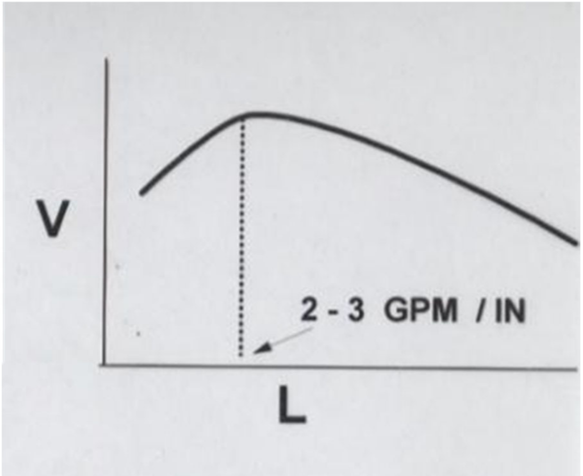

Figure 6. A typical tray capacity diagram, vapor (V) versus liquid (L), with the liquid load expressed as weir loads (liquid flow rate, m3/h, divided by outlet weir length, m). The upper curve is a typical jet flood curve, slightly steeper than in reality to illustrate the principle. The diagram shows a maximum vapor-handling capacity at a weir load of 2-3 gpm/in (20-30 m3/h /m of weir length) .

Christian

First of all, we have to distinguish between multi-pass or multi downcomer trays. Both designs have their application in columns with high liquid loads. Multi pass type trays, like 2 , 4 or 6 pass trays, deliver a stable efficiency, when properly designed. Some have a reportedly lower efficiency due to reduced flow path length and different liquid guiding on the tray panels. Furthermore, all of them come with a dynamically sealed downcomer, which limits the operating range of the tray. If not properly designed, this tray type shows a high risk of gas by passing. Additionally, some of them do not have a manway, which excludes the possibility of inspection during turnarounds or for trouble shooting.

14. WHEN SHOULD TOWER SCANNING BE UTILIZED?

Mike: The best reason to scan is when liquid cannot get down the tower. Several years ago, a paper was given at a Chicago Engineering Conference describing a trayed tower that was having capacity problems. Gamma scanning showed that one downcomer on one particular tray was blocking liquid flow appreciably. Once the tower was idled and entered, the engineers found that a tower revamp crew had left a vacuum cleaner (yes, a vacuum cleaner) in one of the tray's downcomers.

Karl: The tower should be scanned it the tower is have challenges. If a tower is having challenges the simple checks should be performed first. Confirm the levels, take a temperature profile, and pressure profile. These simple checks will then lead to an answer of when a tower should be scanned.

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Commercial gamma scanning has been successfully applied to troubleshoot tens of thousands of towers for over five decades. Scan data have traditionally been interpreted by visual examination of the scans to detect changes or trends in entrainment, froth (or spray) heights, and liquid holdup.

Most gamma scan applications are qualitative, even many of those that may claim to be quantitative. Qualitative gamma scans can readily detect gross abnormalities such as flooding, missing trays, collapsed trays, excessive liquid level in the bottom sump, or heavy foaming. This technique can also help diagnose a flood mechanism and shed light on more subtle abnormalities such as high or low tray loadings, excessive entrainment, excessive weeping, blockage, and multipass liquid maldistribution. Gamma scans performed on a routine basis can also be used to monitor deterioration in column performance due to fouling, corrosion, and other factors.

In packed towers, a "grid" of four equal chords is often shot (Figure 7), one chord after the other. For each chord, the source and detector are moved simultaneously down the bed, taking shots every 2 to 4 inches. This "grid" gamma scan looks for maldistribution and channeling, which is by far the main cause of packed tower efficiency loss. When liquid distribution is good the four chords give the same detector readings. Differences between the chords are interpreted as bed maldistribution.

Judicious setting of the chords can also provide a measurement of liquid height and frothiness in collectors, parting boxes, and distributors, and identify trough or pan overflows or level unevenness causing liquid maldistribution or premature floods. Overflows and plugging (which too will cause overflows) were identified as major troubleshooting issues in packed towers (Kister's book "Distillation Troubleshooting", John Wiley & Sons, 2006)).

In addition, grid scans can normally detect the position of the bed, the disappearance of some of the packings (e.g., by being corroded away), significant blockages and local flooded regions.

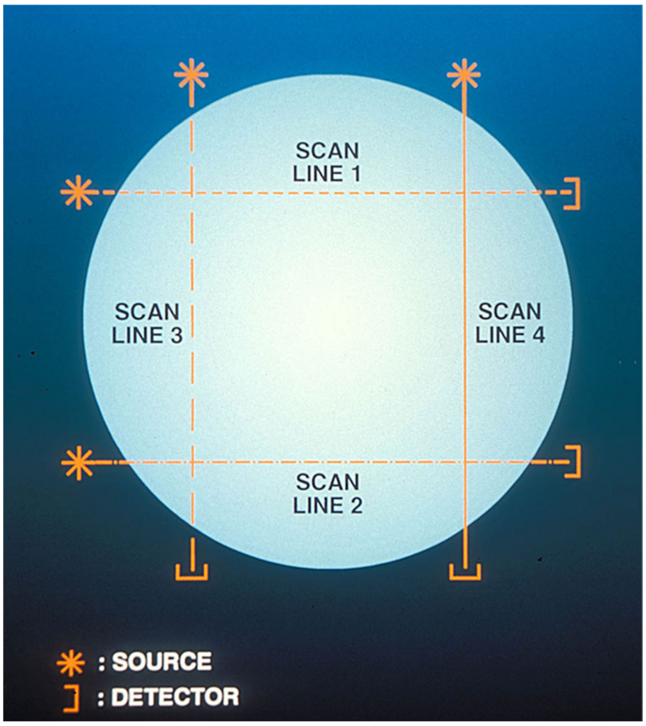


Figure 7. Packed tower grid scan

In CAT (Computer Aided Tomography) scans, a gamma scan source is placed, and a number of detectors (typically about nine) are set around the bed at evenly spread marked radial positions, all at the same elevation. Once done, the source is moved to the position of the nearest detector, the detector to the position of the source, and the scan is repeated. This continues until the source is placed in all the radial positions around the bed. The profiles obtained are then integrated to give the two-dimensional absorption density profile, which identifies liquid-rich regions (high density) and drier regions (low density). The CAT scan can be repeated at additional elevations along the bed, but this runs up the costs. This technique is used primarily to identify the nature of maldistribution in packed beds. It is also invaluable in showing a center-to-periphery liquid maldistribution that grid scans cannot identify.

Stationary monitoring ("time studies") places a number of stationary gamma ray sources with detectors right across at strategically-placed locations along the column shell. Starting in the unflooded condition, the rates (reflux, boilup, feed, or whatever variable is studied) are raised until the amount of radiation transmitted sharply falls in one of the vapor spaces, indicating liquid accumulation and therefore flooding. Stationary monitoring can diagnose where the flooding starts, calibrate liquid level instruments, detect entrainment, and overflow from packing distributors, and detect floods due to high bottoms level. Stationary monitoring is also valuable for accurately determining the flood rates, permitting engineers and operators devise a strategy for pushing the tower to its limits or revamping.

Gamma scans can detect plugging, presence of vapor in liquid lines, and presence of entrained liquid in vapor lines. This technique can give reliable quantitative numbers when performed properly.

Capacity-enhancing features of high-capacity trays such as push valves, truncated downcomers, and multiple downcomers, bring with them unique troubleshooting challenges. Excessive forward push may generate excessive froth gradients, truncated downcomers do not have a static liquid seal, which may permit vapor passage up the downcomers ("blowby"), and multiple downcomers lead to non-symmetrical active areas which may promote channeling and maldistribution. These are difficult, often impossible, to diagnose using conventional troubleshooting techniques such as vendor software, ΔP measurement, and qualitative gamma scans. Judicious multi-chordal gamma scans with quantitative analysis (Kister, H. Z., "Use Quantitative Gamma Scans to Troubleshoot Maldistribution on Trays", Chem. Eng. Progr., February, p.33, and April, p.45, 2013) is the best tool for diagnosing these issues. With this technique, several parallel chords are shot along the flow path (Figure 2). Froth heights, froth densities, liquid heads, and entrainment index data are determined for each tray at each chord and are plotted to-scale on a tray diagram. From these plots, channeling patterns can be inferred.

Recently, Tracerco introduced quantitative packed tower gamma scans. The liquid holdup of the packing is calculated as the difference between the density inferred from the scan and the dry packing density available from the packing vendor. A large difference between the two indicates that there could be some liquid accumulation or solids there, suggesting flooding or plugging. This type of scan is effective for evaluating the extent of maldistribution and bed plugging.

Christian: Tower scanning gives you more information about distribution of the liquid inside the column during operation. Also “dislocated” internals/trays can be detected. Typically, it can help to locate the problem and also to make assumptions about possible options to solve the problem. However, it does not usually allow for solutions without opening the column. So, you have to decide whether it makes sense to scan the column or to directly open it.

15. WHAT IS A GOOD TURN DOWN PERCENTAGE TO KEEP GOOD EFFICIENCY?

Andrew: Most specifications arbitrarily call for a turndown to 50% of maximum capacity.

Packing has an exceptionally large capacity range and turndown is usually not a problem. The challenge in packed towers lies with the liquid distributors. Most liquid distributors should perform well from 100% to 50% of rate. The range can be extended by using more complex, more expensive, and larger equipment. Even so, it is unusual to see a liquid distributor with a realistic operating range wider than 33% to 100%. Pushing a gravity distributor to operate below its minimum reasonable flow rate leads to a gradual, but accelerating, drop in performance. Exceeding maximum rates tends to result in a rapid performance drop. For pressure-spray distributors the trends are reversed.

As for trays, most single pass trays will operate reasonably well in the 33-100% range. This is true of sieve trays as well as valve trays. The more tray passes, the smaller the turndown range. Special high-capacity trays with dynamically sealed downcomers may have reduced operating ranges for the liquid rate. Special designs can accommodate much larger turndown ranges. Trays with ranges of 10% to 100% can be built, but the required features normally lead to higher tray pressure drops and either lower capacity for the diameter or higher tray spacing.

Mike: Plant operators can select any turndown percentage that they truly need. Tray designers can adjust the tray design to accommodate those needs. Here is an extreme example: If a trayed tower needs to be turned down to 1% of full rates, that tower can indeed achieve good efficiencies as such rates - as long as the tray spacings are 100 inches.

Karl: Most pumps and compressor have a minimum flow of 60%. Rarely is a plant running less than 60% due to these constraints. But I have seen design jet flood designed at 30% - at the normal design unit rate, which leads to low tray pressure drop and low tray efficiency. I prefer to design jet flood 50% or higher. Jet flood can go to 110% without losing efficiency, if you have good down comer capacity – so at 50% you still have 60% safety factor. Many operating companies still require downturn of 33% and vendors are required to meet this requirement even though the plant will almost never be running at 33%.

Christian: Typical turn down ratios for packings are:

•1st and 2nd Generation Packing: 1 : 3

•3rd and 4th Generation Packing (RMSR/ RMXR): 1 : 4

•Structured metal Packing (RVT RMP 250):1:6

For random and structured packing, the “turn down ratio” may not be the proper design criteria. It is more useful to discuss minimum and maximum gas factors and liquid loads. Limitations are normally fixed by the tower internals, mainly liquid distributors and collecting trays which have lower operation ranges than packing media.

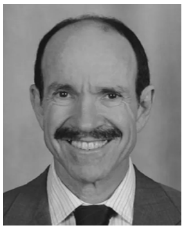
Typical turndown rates for trays are:

•Movable Valve trays (e.g., V1): 1 : 4-5

•Sieve and fixed valve trays: 1 : 2-3

We are always working to improve product performance. For example, our MRV Fixed Valve Tray shows good and constant efficiency over a wide operating range of 1 : 4, compared to the typical operating range of 1:2-3 for standard fixed valve trays (FRI test unit, total reflux).

Besides that, there are special tray designs that show much higher turndown ratios (e.g., bubble cup trays) but have other disadvantages (cost, pressure drop). Dual flow trays are used in some special applications; they only have a very low turndown ratio.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Resetarits


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Andrew W. Sloley is a Principal Consultant at Advisian. His current responsibilities include consulting, conceptual design, and front-end design for the refining and petrochemical industries. His experience includes detailed process and mechanical design and troubleshooting for distillation systems. Previously he worked at Exxon Chemicals, Glitsch, PCS, The Distillation Group, and CH2M HILL. His specialty is the area of distillation, product recovery, and heat integration. This has included conventional distillation, extractive distillation, liquid-liquid extraction, and complex heat-integrated processes in the refining, petrochemicals, and chemicals industries.

Christian Geipel received his Master's in Mechanical and Process Engineering from TU Darmstadt/Germany in 2000 followed by his PhD in thermodynamics from TU Darmstadt/ Germany (Prof. Stephan) in 2006. From 2004-2013 he held various positions in the field of thermodynamics and process design for AUDI and LINDE ENGINEERING in Germany. Since 2013, he has been one of the managing directors of RVT Process Equipment GmbH.

After receiving BS and MS degrees in chemical engineering, Mr. **Mike Resetarits'** career began in 1974, at Union Carbide Corporation. While there, Mike worked on the development of the MD distillation tray and the MU liquid-liquid extraction tray. The ECMD tray was invented with Dr. Mike Lockett. Several Union Carbide businesses were acquired by UOP in 1988, at which point Mike became a UOP R&D Manager. In 1999, Mike joined Koch-Glitsch as their R&D Director and later as their Global Tray Product Manager. Mike helped to lead the development efforts associated with Ultra-Frac trays and Flexipac HC structured packing. Mike spent appreciable time in the field, running pilot plant tests, starting units, performing troubleshooting work and collecting data. He is coauthor of 17 patents and over 60 presentations and articles.

Karl Kolmetz is a Senior Technical Engineering Professional / Senior Manager at KLM Technology Group. He is the Editor for Engineering Practice Magazine and the Handbook of Process Equipment Design. He has authored more than 140 publications on a variety of subjects for product recovery, troubleshooting, training, project management, process design, process safety management and simulation with safety and environmental focus. His research interest focuses on how to apply the fundamentals of engineering to practical applications. The Handbook of Process Equipment Design has over 100 chapters on a variety of process equipment including, line sizing, pumps, compressors, relief valve, flares, separators, distillation and others. Karl is a Certified Practicing Engineer (CPE) from the International Association of Certified Practicing Engineers (www.iapce.com).

Catalysts | Heart of Hydroprocessing Units

Rajesh Sivadasan

INTRODUCTION

Refining industry is going through many challenges. Chief among them being the use of technologies and catalysts which reduces the carbon footprint of products they produce while at the same time raising the performance of products.

With increasing frequency, refiners driven by economics, are processing heavier and more difficult crudes, while at the same time have to meet the increasingly stringent product specifications. Attaining these goals require alterations in either process and/or catalyst design. Hydroprocessing technologies allow production of cleaner and better performing premium products. It has become an integral part of industry and practically it is impossible to attain product specifications without passing through at least one hydroprocessing step. Catalysts are “the heart” of hydroprocessing technologies and its performance is critical to refinery’s profitability. Going forward, its importance will increase in the coming decades to combat climate change.

This paper will provide a brief introduction on hydrotreating and hydrocracking catalysts commonly called as hydroprocessing catalysts.

CATALYSTS

Catalysts are materials which when added to a chemical reaction increases the rate or the speed at which the reaction is occurring. Every reaction proceeds via a path/mechanism called the reaction mechanism with a particular activation energy associated with it. This activation energy is the minimum amount of energy required to take the reactants to the condition in which they will start reacting with each other to carry out the chemical reaction. If this amount of energy is not available, no reaction will happen. As shown in Fig 1, introduction of catalyst results in a different path/ mechanism whose associated activation energy is much smaller than that without the catalyst and results in a higher reaction rate at the same temperature.

HYDROTREATING CATALYSTS

Hydrotreating process is applied either as a finishing step of final products or as an intermediate step to prepare feed for downstream processes like reforming, fluid catalytic cracking or hydrocracking. The process involves chemical reactions between organic compounds containing the contaminants and hydrogen in presence of a catalyst. If these

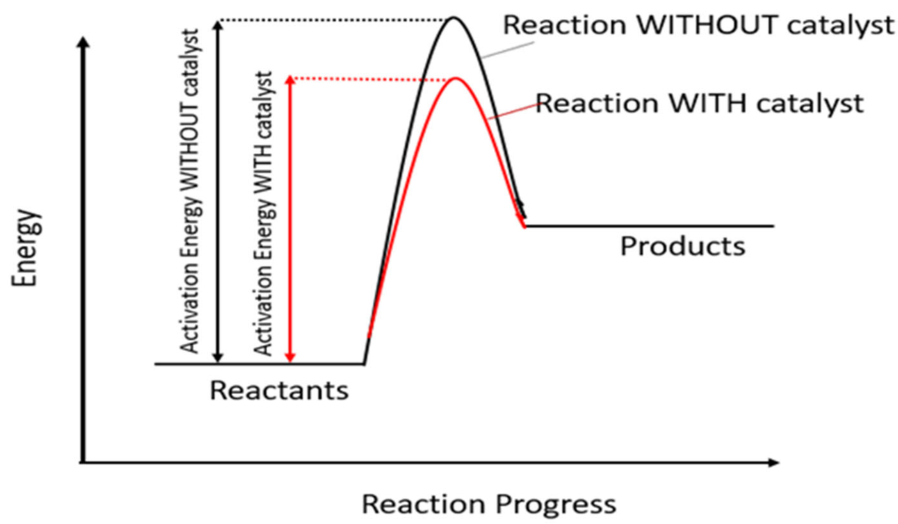

Fig 1: Energy diagram

contaminates are not removed, it may result in emissions and cause environmental problems, damaging ecosystems as well as human health. Based on the types of contaminants removed, the reactions can be called as hydrodesulfurization or HDS (removing sulfur), hydrodenitrogenation or HDN (removing nitrogen), hydrodeoxygenation or HDO (removing oxygen), hydrodemetallization or HDM (removing metals) and hydrodearomatization or HDA (saturating aromatics). Process flow schemes are pretty much similar for most hydroprocessing processes; however, the operating severity varies, determined by variables like type of contaminants, depth of contaminant conversion, hydrogen partial pressure and catalyst. Typically, the reactions are conducted in fixed trickle bed reactor(s) with catalyst pellets stacked in a packed bed and gas (hydrogen) and liquid (oil) flowing co-currently from the top of the reactor.

Catalyst activity is determined by several factors including the type/quantity of metals, dispersion of metals and metal-support interaction. Though metals with different groupings have been researched, typically two groups of metals are used in the production of commercial catalysts. The active phase normally comprising metal from Group VI (Molybdenum-Mo) and the promoter metals from Group VIII (Nickel-Ni and Cobalt-Co). Other promoter metals can also be added to improve the performance, for e.g., Phosphorous is added to NiMo catalysts with the objective to improve HDN activity. Traditional hydrotreating processes mainly employs CoMo and NiMo combinations. One of the key components optimised during catalyst manufacturing process is the quantity of metals or metals loading. As shown in Fig 2, activity of catalyst increases with higher metals loading but after a peak, activity drops off. There is an optimal loading for each support, which varies from 1-6 wt% as

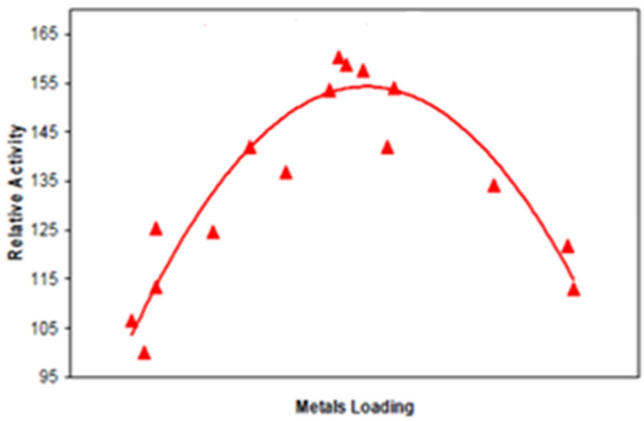

Fig 2: Relationship between metals and activity (ART, 2015)

Ni or Co oxides and from 8-24 wt% as Mo oxides. Active forms of these metals are their sulfide forms and good sulfiding is essential for achieving full performance potential.

The interaction between metal and support also has a big impact on activity. Before 1995, most catalysts were of “Type I” kind in which the metals had a strong interaction with the support. Since then, “Type II” catalysts have become the industry standard which relies on careful tuning of the metal-support interaction and have found wide acceptance in a variety of hydrotreating applications especially in ULSD production and hydrocracking pretreat. Active phase in Type II catalyst has weak interaction with the support and has more stacking of MoS2 slabs than Type I giving it more intrinsic activity. As shown in Fig 3, further improvements in Type II catalysts have enabled them to provide significantly higher activities. This has been made possible with better raw materials, improved manufacturing techniques including use of chelates, optimisation of metal dispersion on support, enhanced promoter effectiveness and advanced analytical tech-niques giving more insight into catalyst morphology.

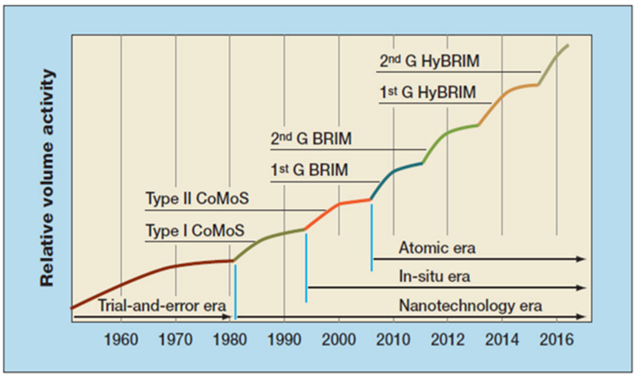

Fig 3: Development progress – from Type I through Type II (Haldor Topsoe, 2016)

The metals are supported on a carrier (support). The nature of support plays a key role in the morphology, dispersion, and catalytic activity of the prepared catalysts. The support also provides the mechanical strength to the catalyst as well as high surface area to maximize metals dispersion. Among the available and developed supports for hydrotreating catalysts, gamma alumina (γ -Al2O3) has been widely applied due to its reasonably high surface area, porosity and thermal stability, excellent mechanical strength, very good morphology, easy availability and low cost.

To process a particular feed, a universal catalyst or a catalytic system does not exist and catalyst selection must be tailor-made. In low pressure units that need low reaction severity, a CoMo combination is typically applied. For high pressure units were the reaction severity is high, a NiMo combination is normally preferred. To optimise for particular reactions within "zones" in reactor, a stacked combination of CoMo and NiMo catalysts is sometimes chosen to provide the best performance rather than just using one combination alone. Fig 4 illustrates an example on balancing HDS activity while minimising hydrogen consumption using stacked combination. As NiMo catalyst is added to the system, there is a significant boost in HDS activity and the product sulfur goes down before eventually hitting a minimum (max HDS activity). With increase in percentage of NiMo catalyst, hydrogen consumption also increases. In the region where the system shows the best HDS activity, hydrogen consumption is only slightly greater than that of all CoMo system, and well below that for all NiMo system.

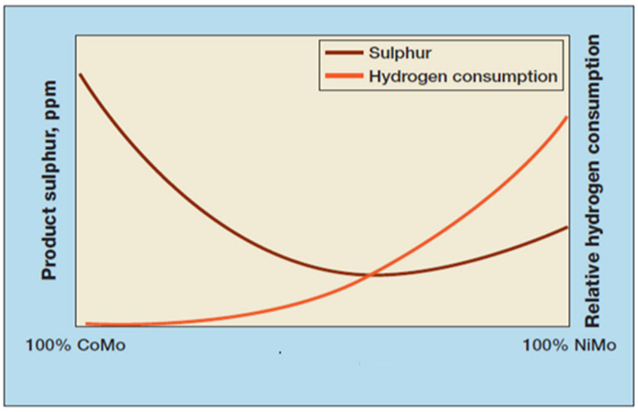

Fig 4: Balancing HDS activity while minimising H2 consumption (ART, 2014)

Good HDN activity is the primary function of a hydrotreating catalyst in a hydrocracking unit as the organic nitrogen compounds are detrimental to the performance of cracking catalyst. The rate limiting step in HDN reaction pathway is aromatic ring saturation because in most refractory nitrogen compounds the nitrogen atom is incorporated in the aromatic ring which needs to be saturated first before the nitrogen can be extracted. As a result, NiMo catalyst are used as hydrocracking pretreat catalysts.

HYDROCRACKING CATALYSTS

Compared to hydrotreating catalyst, hydrocracking catalyst performs a dual-functional role where it cracks the high molecular weight hydrocarbon and then simultaneously hydrogenates the unsaturates that are formed during the cracking step. The main composition of any hydrocracking catalyst contains an acid support for cracking/ isomerization functions and metals for hydrogenation/ dehydrogenation functions.

For cracking catalysts, hydrogenation function is provided by Group VI (Mo, Tungsten-W) and VIII base metals (Co, Ni) or noble metals (Platinum-Pt or Palladium-Pd), with extensive use of base metals as these are relatively cheap. Most commonly used combinations in industrial hydrocracking catalysts are NiMo and NiW in their sulfided form. Metal pairs such as CoMo and CoW have limited use because of their lower hydrogenation activities. Though not widely used as base metals, noble metals exhibit much higher activities than the sulfided base metals in a clean reaction environment and are typically used in the second stage of a two-stage hydrocracking process. In the first stage operation or in a single-stage process, noble metals tend to lose their advantage over the sulfided base metals due to the presence of contaminants, primarily H2S. Without significant performance advantage, it is not economical to operate with noble metal catalyst under these process conditions. The metal loading for base metal catalyst varies from 1-6 wt% for Ni oxides and from 8-20 wt% for Mo or W oxides. In addition to metals loading, atomic ratio of metals is also important as it impacts the hydrogenation activity of the catalyst. A study of the hydrogenation activity of Group VI and VIII using toluene as a model compound in the presence of H2S, concluded the optimum atomic ratio to be about 0.25 as shown in Fig 5.

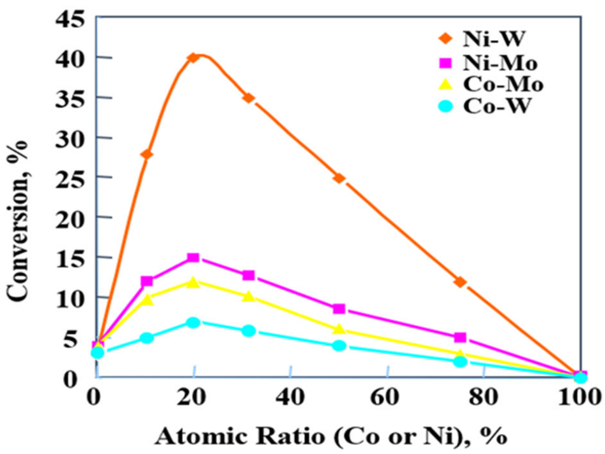

Fig 5: Correlation between conversion and atomic ratio (UOP, 2004)

Since the acid function is responsible for isomerization/cracking reactions, cracking catalyst activity is primarily determined by the total acidity of the catalyst. Acidity is a function of silica to alumina ratio and is provided either by amorphous (alumina and silica-alumina) or crystalline (zeolites) components which also acts as the support. Of several solid oxide materials used as amorphous supports, silica-alumina is most widely used due its high acidity and low cost. Cracking catalysts made with amorphous solid oxide have high selectivity towards distillate products. Zeolites are microporous, crystalline aluminosilicates with tetrahedrally coordinated framework aluminium. Among various zeolite materials, modified Y zeolite is most successfully applied in industrial hydrocracking. Most zeolites are synthesised from a slurry consisting of silica, alumina and caustic. The synthesised zeolite is modified by ionic exchange and thermal or chemical treatment called dealumination to obtain an active catalyst; an ultra-stable Y (USY) zeolite with appropriate unit cell size (UCS). Catalyst manufacturing operating severity determines the degree of dealumination. As the degree of dealumination increases, silica to alumina ratio increases and UCS decreases as shown in Fig 6. Since the properties of dealuminated Y zeolite can be varied so widely during manufacturing process, USY is the most versatile material for hydrocracking. A high UCS USY zeolite will have higher activity and are typically used in naphtha production while a low UCS USY zeolite will have lower activity and are used for distillate production.

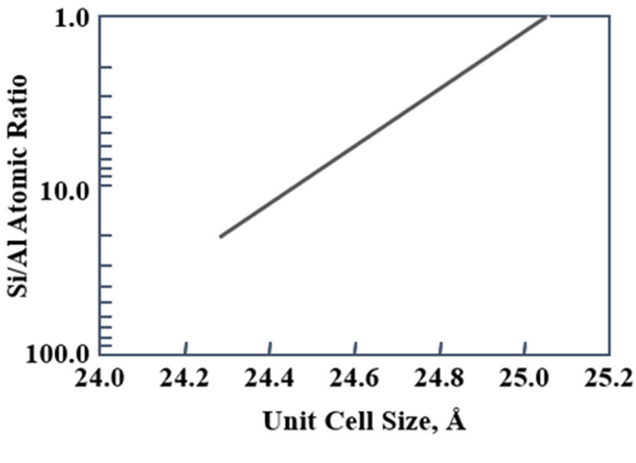

Fig 6: Correlation between zeolitic unit cell size and Si/Al ratio (UOP, 2004)

Catalyst pore structure plays an important role in how they convert the feed. Fig 7 shows the difference in the pore structure of an amorphous catalyst having more larger pores than the zeolite catalyst. Larger pore structure of amorphous catalyst support preferentially converts heavy molecules and combined with its lower activity result in higher yields of the heavier products.

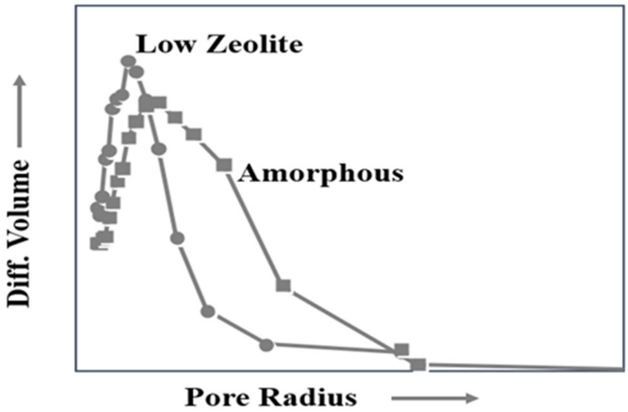

Fig 7: Comparison of pore structure (UOP, 2004)

The product yield depends on metal to acid activity ratio of the catalyst in use and increases with increasing ratio. The yield could approach maximum when a catalyst can exhibit ideal cracking behaviour. The product distribution characteristics are quite different between amorphous catalysts and those containing Y zeolite. Fig 8 shows the incremental yield as a function of the distillation temperature of the product slates for both types of catalyst. At the same conversion level to 700° F, amorphous SiO2·Al2O3 catalyst shows a much higher distillates selectivity whereas the product of Y zeolite catalyst is heavily skewed toward naphtha. The Y zeolite catalyst being significantly more active than the amorphous catalyst lacks the proper metal/acid balance. As a result, excessive secondary cracking occurs over the Y zeolite catalyst shifting the product to lighter boiling range.

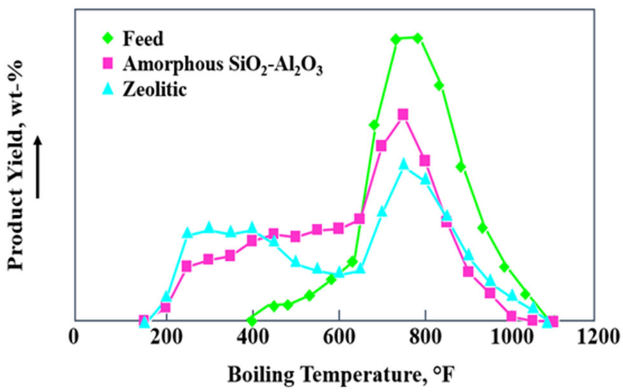

Fig 8: Comparison of product profiles (UOP, 2004)

CONCLUSIONS

The pivotal role of hydroprocessing units in a modern refinery hinges on its unique ability to produce a wide range of premium quality products with catalyst technology at its heart. Pushed by more stringent fuel quality legislations, catalyst vendors are combining state-of-the-art characterization and manufacturing techniques to come up better catalysts and innovative catalyst selection techniques to unlock opportunities which can significantly enhance refinery profit in a sustainable way and cause less damage to the environment.

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De-scaling Application: Why the Pump Fails?

Abhijeet Keer

1.0 INTRODUCTION

It is essential for the steel rolling mills to improve efficiency and product quality. One of the key factors in determining the quality of the final product is the scale removal process; scale which isn't properly removed can be combined with the steel billet as it passes through the mill, resulting in defective or lower quality steel. This process of removal of Scales is called Descaling.

Steel slab or billet heated to approx. 1200˚C in reheating furnace is transferred to roller tables on its onward journey to rolling mill stand where it will be rolled in multiple passes to achieve lower thickness or desired shape. Slight oxidizing atmosphere is maintained in reheating furnace. That is why, it results in the formation of Iron oxide scales on the Steel billet. Scale formed in the reheating furnace need to be washed before sending the material to mill stands.

2.0 PROCESS DESCRIPTION

The Process of descaling i.e. Striking of Water Jet with High impact Energy can be accomplished through different ways. Different methods such as direct descaling or indirect descaling have been incorporated. Let us understand how this works.

Direct descaling involves using a High pressure usually a Multistage Centrifugal Pump which will deliver the required amount of flow through the Nozzles attached to header in the Descaling Box. In case of direct method, one thing that must be noted is that, the total demand of the flow for descaling is served by Pump alone.

Whereas in In-direct method, the Pump along with the accumulator is used. The accumulator is a pressure vessel which stores the fluid at high pressure and supplements the pump flow as and when required.

Now, the question may arise, what is the need for such different methods of direct and indirect.

If we look at the Process of descaling in detail, we will realize the pain points associated with the system. The Descaling operation is an intermittent operation. The Billet or slab that rolls over the production line has a cycle time. It passes through various stations for certain period of time. Descaling section being one of the first stations that the slab encounters, once it comes out of the reheating furnace. Hence the process of Descaling is a cyclic one. As shown in fig.1, when the material reaches Descaling section called as descaling box, the descaling valve opens and the specially designed spray nozzles installed on top and bottom header, spray the jet of water on the material. The descaling valve remains open typically for 20 to 60 seconds of cycle time depending upon the length of the steel slab and closes again. The Steel billet/slab, in order to attain required thickness has to go through multiple passes through the rollers. Hence, in some cases, in one cycle lasting for 5 to 7 minutes, descaling valves open and close for 5 to 7 times, sometimes at the interval of 30 seconds or less.

Since the system has a cyclic operation with fluctuating demands, this is likely candidate for the use of Accumulators. The energy efficiency of the operation, as can be seen, is low. Of course, the measure contributing factor for the energy consumption being the Pump. As the efficiency of the descaling process depends on the impact of water jet, it is quiet certain that the system will need the High pressure pump with moderate flow. Hence this calls for the usually high speed Multistage Pumps, since there is clear advantage of efficiency in case of centrifugal pumps. However, the large amount of Bypass flow time during the life time of the pump, the energy wastage is considerable and hence, it is essential to reduce the size of the pump which will in turn means reducing the flow.

So use of accumulators gives this liberty for reduction in size of the pump. The accumulators supplement the Pump by adding the required amount of flow when necessary. Without the accumulator, the pump size would have been larger to meet the demand of the operation.

This shows the first and foremost pain point of the System i.e. intermittent operation.

Secondly, if we refer to fig.1 again, we will see that there is recirculation valve placed near to Reservoir from where the Pump is taking the Suction. As we have seen, the operation of descaling is intermittent one; there is a dwell time in the operation. Till the next cycle starts, the pump has to keep running and hence, the flow of the pump is routed through recirculation valve, back to the reservoir. The operating point for the pump usually in this condition is Minimum allowable continuous flow, which shall be sufficient so that, there will not be any undesired Temperature increase of the water being pumped and secondly the vibrations of the equipment shall not go beyond the limit.

So our second pain point highlighted here is the running of pump towards left of the performance curve.

3.0 VARIOUS PROBLEMS ASSOCIATED WITH THE PROCESS

Now, as we understand the system, let us list out the problems associated with the pumping of the water required for the operation.

1. Frequent Change of operating point
2. Running the pump on bypass flow
3. Varying system resistances
4. Quality of water being Pumped
5. High speed application
6. High pressure meaning high system resistances

These are some of the causes of problems associated with the Pump, which lead to frequent breakdown of the equipment.

The typical failures encountered by the Pumps are high vibration, high bearing temperatures, seizure, even shaft breakage is not uncommon.

The frequent changes in operating condition lead to transient operations and also lead to fatigue loading of the pump. If we observe the cross section of one of these typical multistage pumps, we see that the shaft of the pump is carrying a radial and huge amount of Axial thrust due to the high pressure operation. The shaft being slender and with number of impellers mounted will deflect. Of course, it will be designed to keep the deflection within limit but the simultaneous action of Axial thrust, and that too fluctuating, present the conditions, which are needed to be analysed in detail during the design stage and will be

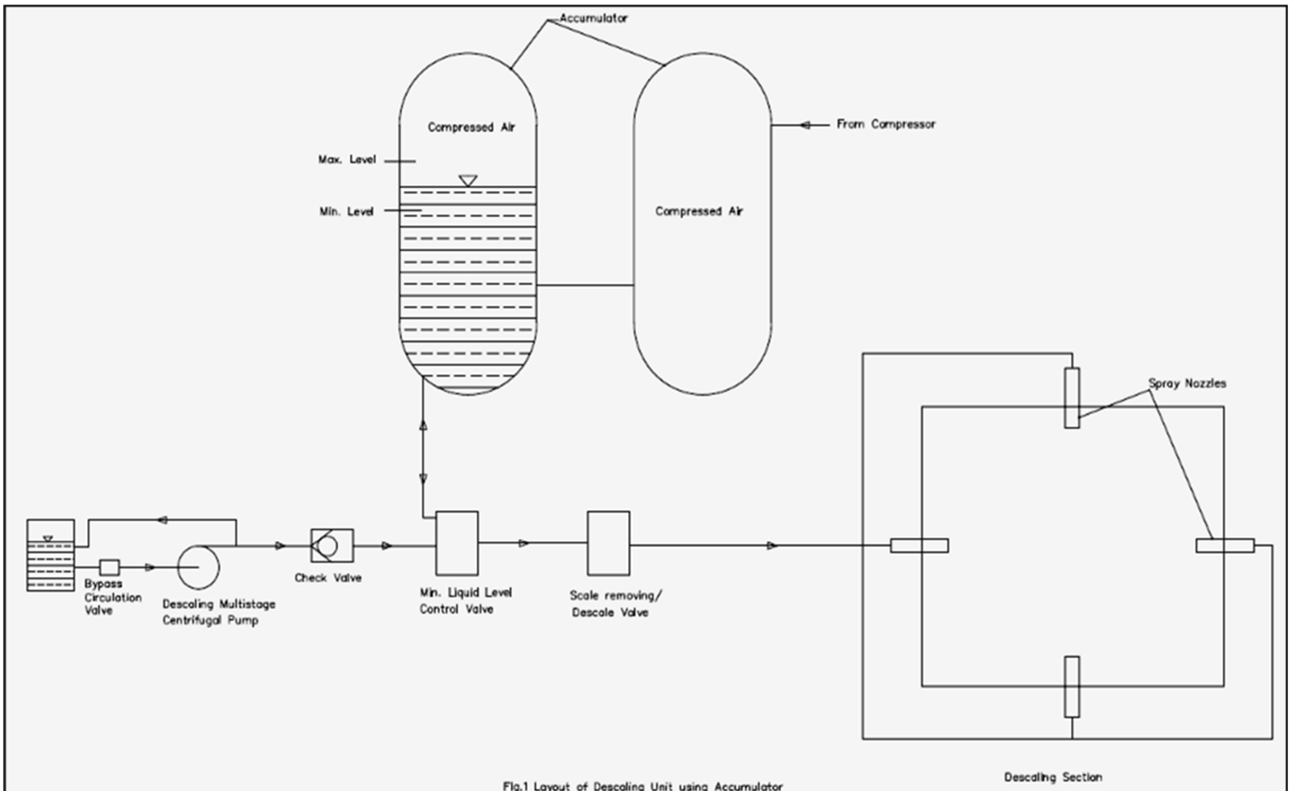

Fig.1 Layout of Descaling Unit using Accumulator

Fig.1

different for each application. Many of the times, these transient conditions are overlooked rather not even considered during the enquiry stage. The result being the frequent failure of the pump on field.

Most commonly observed pain point is the vibration of the equipment. One thing to note here is the design of Rotor of the multistage pump. Here is the quick simple calculation to illustrate, how one might go wrong, if the submitted operating conditions by the customers are not assessed properly.

Let's say the operating speed of the equipment under consideration is around 4500 rpm.

Now, at this operating speed if we want to keep the vibrations below 2.5 mm/s then it is imperative to balance the Rotor dynamically at grade G2.5 as per ISO 1940. Now let's do some basic calculation;

V = r x w ; where v = vibration velocity mm/s

r = eccentricity, mm

w = angular speed, rad/s

now,

$$r = \frac{2.5 \times 60}{2\pi(4500)}$$

Hence, r = 5.3 microns

Which means that, the eccentricity allowable for the centre of mass at given operating speed is only 5 microns. Now, if we are assembling the impellers on the Shaft, there must be some clearance for the assembly of shaft with the impeller, surely now, if we choose the sliding fit between shaft and impeller which might result in minimum limit higher than this 10 microns (radially 5 microns). This will surely lead to higher vibrations.

Now, this all what we discussed is for the Best efficiency point of the pump. One can guess, what will happen when the pump is running on the left of the curve where the vibrations are inherently higher due to flow transients.

Another point of interest could be checking the natural frequencies of the baseplates as well as complete assembled pump. One should make sure; the pump is not in resonance while operating at given operating speed.

In case of Multistage Pumps, the axial thrust that the Shaft needs to carry is huge. There are various means by which the Axial thrust can be compensated. For multistage Pumps, balancing disk or balancing drum are usually employed methods to counter Axial thrust. Balancing disk completely balances the axial thrust and hence does not require an additional thrust bearing. Balancing drum on the other hand, requires additional thrust bearing to carry residual axial thrust

For intermittent operation and high axial thrust, such as this, of course balancing drum is the obvious choice. One of the key things here worth noticing is that, the design of the balancing drum is for single duty point, which is Best efficiency point of the Pump. Hence for all other operating points, especially at lower flows, the Axial thrust will change and will be huge due to high pressure at lower flow.

These fluctuations in the Axial thrust could even lead to relative movement between the Bearings thrust plate and the Shaft. And this likely to result in the Fretting damage as well. Fretting occurs when there is an oscillating movement between two surfaces which are having almost zero clearance. In case of thrust plate, it is mounted on the shaft with the Interference fit, but due to huge amount of thrust and that too fluctuating; the fretting damage is likely to happen.

Also during design stage, one needs to take care of the elongation of the shaft due to loading and should select the material which will remain stiff without any excessive deflection.

One more problem associated with intermittent operation that might be overlooked during the selection of the equipment is, the changes in nozzle loads. This leads to the misalignment of the equipment, which leads to higher vibrations, uneven wear, bearing temperature rise, etc.

Along with that, if the pump is running at lower flow, means at higher system resistance, there are chances of flow separation on the suction side of the vane. This in turn disturbs the head generation and results in flow pulsations in the fluid and also in the pump components. Another problem is the suction or discharge recirculation due to low flow conditions. Mostly, in case of low specific speed pumps such as these, discharge recirculation could be more pronounced which could even lead to damage at the Impeller shroud outside diameter and vane tip area.

Another pain point is the descaling water used accumulates with solid particles. Despite prefiltration, not all particles can be removed from the water. Solid particle accumulation in the process water can be harmful for the sealing in the liquid end and other wetted parts. This also results in uneven wear of the hydraulic components, especially close clearance areas such as mechanical seals. The wear at impeller outside diameter is also not uncommon.

4.0 CONCLUSION

The descaling operation being a cyclic one is critical for the pump and associated accessories and piping as well. Hence careful evaluation and exact formulation of specifications for the equipment is very essential. During design stage, one need to address at least the points listed above, in order to have trouble free operation of the equipment.

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Abhijeet Keer is a young passionate engineer, who has been working in the Design field of the Centrifugal Pumps for over 6 Years. The mechanical Construction and Materials are his strong skills. With over 6 years of his Career working as a Design Engineer with major players in Pump Industry, KSB limited and Kirloskar Brothers Limited, he has developed strong affinity towards this field. He completed his Bachelor’s degree in Mechanical Engineering from University of Mumbai, India. He enjoys writing articles related to Pumps and applications. His professional experience covers new Product design and developments, material selection and application engineering and complete mechanical constructions.

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How to...Downcomers | Part 5
All About Downcomers and Weirs

Dr. -Ing. Volker Engel

Tower trays and internals are the heart of all distillation columns. Their design is an essential part of a process engineer's task and determines the process reliability and economy.

This article is the part of a series on different kinds of trays and internals.

In almost all tray towers, the liquid flows horizontally from the inlet, gets in contact with the vertical streaming gas, generates a two-phase layer on the active area, leaves the tray at the outlet weir and degases in the downcomer while passing to the next tray below.

The main focus in tray towers is often only the active area (type of tray, pressure drop, froth height, efficiency, ...), where the mass transfer takes place. To understand all the complexity of trays, it is necessary to get an overview of the various shapes and layouts of downcomers.

DOWNCOMERS (DC)

The main function of the downcomer is to collect all liquid from the active area, degas the liquid, lead the liquid to the next tray and seal the downcomer against gas bypass (Fig. 1).

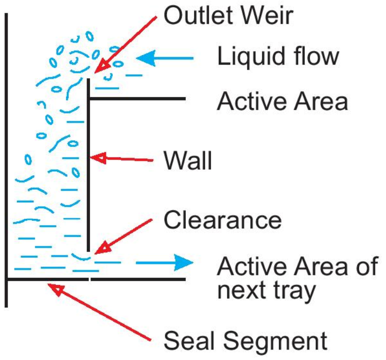

Fig. 1: Parts of a downcomer

NOTE: The downcomer belongs to the tray, where the liquid **leaves** the tray. Its numbering is therefore the same as the tray. Any downcomer above the top tray (e.g. for inlet) is called "False Downcomer" (ref. *Fig. 33*).

The size and shape of the downcomer is specified by the liquid flow rate, whereas the size of the active area is determined by the gas load.

One of the main tasks at designing a proper tray is to choose suitable values for these two areas. This sets the tower diameter!

FLOW PASSES

One of the very first steps in the design of a tray is to deal with the liquid load: It shows, how many downcomers per tray are required and how the downcomer(s) have to be designed. To handle the liquid load, you have to supply the appro-priate downcomer area as well as enough weir length.

To achieve this goal, there are different layouts with one or more downcomers and special down-comer shapes. (Some other designs like Dual-flow trays, baffle trays, shower decks, ... have no down-comers.)

In the easiest case, there is only one downcomer (Fig. 2). The liquid is streaming from one side of the tower to the other. There is one active area and one flow path. This (common) design is called 1-pass or Single-pass tray. The tray design is turned by 180° at each stage. It is used for tower diameters up to 3m. (For some special applica-tions you will even find 1-pass trays up to 8m.)

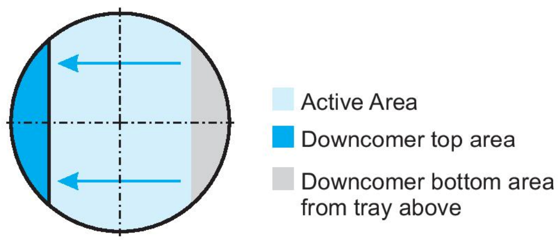

Fig. 2: 1-pass tray layout

For higher liquid loads and larger diameters, two liquid paths are needed (Fig. 3). This leads to two different tray designs, which are used alternately for the stages (therefore the odd tray numbers belong to one design, the inboard design has a center downcomer, the other (outboard design) has two side down-comers. The liquid is strea-ming from the cen-ter down-comer to the side downcomers and on the next stage from the sides to the center downcomer. The active area is symmetrical on each tray. 2-pass designs are often used in practice.

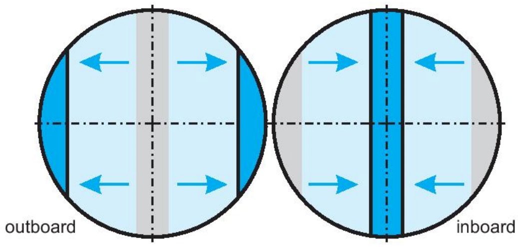

Fig. 3: 2-pass tray layout

The next logical step is the 3-pass tray design (Fig. 4). There is a side downcomer and an off -center downcomer. The design is rotated by 180° per stage. There is only one design for all trays (as for the 1-pass design). This is an ad-vantage in terms of the investment costs compared to the 2- and 4-pass tray design. On the other hand there are three different shaped active areas per stage and a very dif-ferent geometry of the downcomers. There-fore it is not easy to design this 3-pass tray for a wide operation range and you will not find this type very often in practice (but you will!).

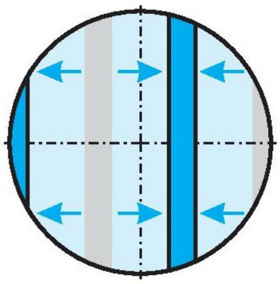

Fig. 4: 3-pass tray layout

The 4-pass tray has two tray layouts (Fig. 5): One stage is equipped with two side down-comers and one center downcomer (outboard), the other with two off-center down-comers (inboard). As each layout is symmet-rical to the tower center line, there are only two different active area shapes. Therefore it is easier to calculate than the 3-pass-design – but as the liquid splits up in two different ac-tive areas (with different froth heights, pres-sure drop, ...), the calculation of 4-pass trays needs a complex iteration for the individual load of each active area for each design case (min / design / max).

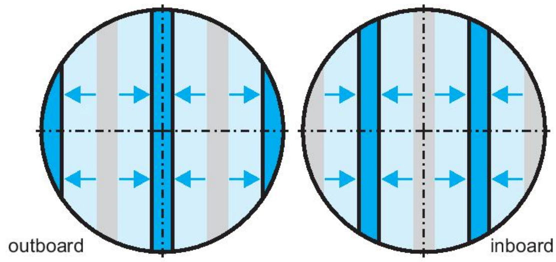

Fig. 5: 4-pass tray layout

The next logical number of downcomers would be the 5-pass tray. You will find it in lit-erature, but rarely in practice.

For large tower diameters the 6-pass tray is used (Fig. 6). Since it is symmetrical to the tower center line, there are "only" three differ-ent active areas per tray and four different down-comer shapes per tray. Therefore the calculation is similar to that of the 3-pass tray. The outboard layout has two side down-comers and two off-center downcomers, the inboard layout has one center downcomer and two off-center downcomers.

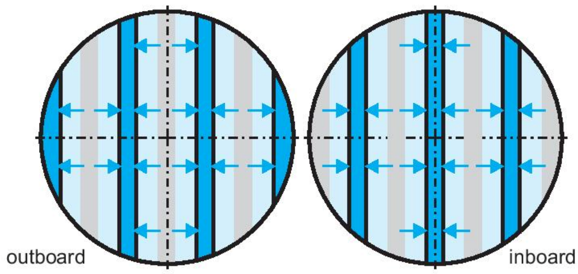

Fig. 6: 6-pass tray layout

The choice of a certain design (number of passes) and setting the width of the down-comer results in a certain length of the flow paths. This flow path length is relevant in terms of hydrau-lics (contact time of the gas-liquid mix-ture) as well as in terms of practice (the flow path length is the maximum width of one dimen-sion of the manway). Therefore you have to check these hydraulic and securi-ty aspects, too.

MULTI-DOWNCOMERS

The designs described above are called con-ventional multi-pass trays and they are "bound" to the circle geometry of towers. In-stead of using the tower shell as part of the downcomer wall, you can place downcomer boxes in the active area. These designs are called Multi-Downcomer designs (”MD trays”).

There are different principles:

You can place the downcomer boxes over the entire diameter of the tower and rotate the design by 90° per stage (Fig. 7). The crossing points of the boxes are blocked to have no liquid shortcuts.

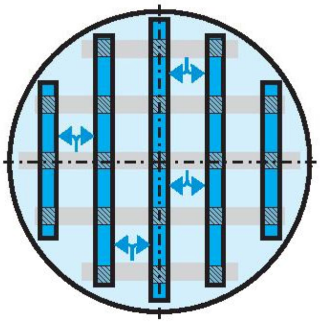

Fig. 7: MD layout (uop design)

Another option is to place the downcomer boxes only on one half of the cross section area and rotate the designs by 180° per stage (Fig. 8). These designs are also known as Calming Section trays.

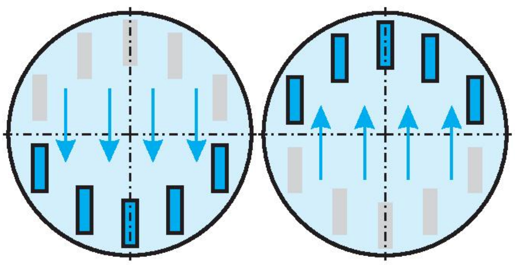

Fig. 8: MD layout (“Calming Section”)

Another type is shown in Fig. 9. It is also known as HiFi-tray (Shell).

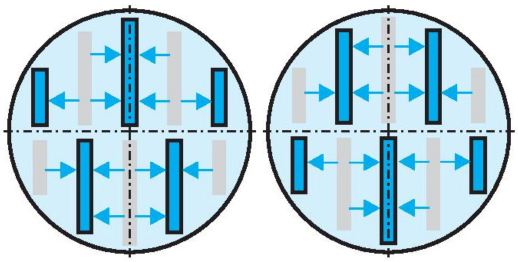

Fig. 9: MD layout (“HiFi-tray”)

These MD trays are normally used for large liquid load when volumetric ratio between vapor and liquid rate is low (medium to high-pressure systems).

REVERSE-FLOW

Another design for the flow path is the so-called Reverse-Flow tray (Fig. 10). It is used to achieve long flow paths and long contact times between gas and liquid.

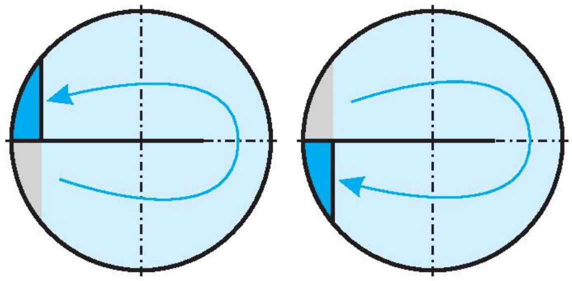

Fig. 10: Reverse-Flow layout

DOWNCOMER SHAPES

The side downcomer in a classic design is nor-mally chordal (Fig. 11). This design is comparably easy in construction and fabrication for tower attachments and tray parts.

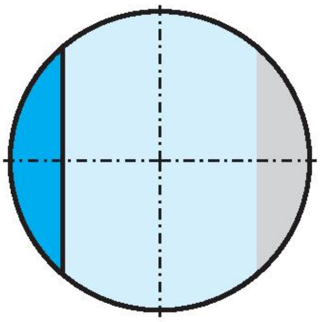

Fig. 11: Chordal downcomer

For high liquid loads you can have multi-chordal designs (Fig. 12). Especially at the side down-comers you are struggling with the geometry of round towers: changing the width of the side downcomer has great effect on the area, but small on the weir length. With a multi-chordal design you can achieve a long(er) weir length without increasing the corresponding chordal downcomer area.

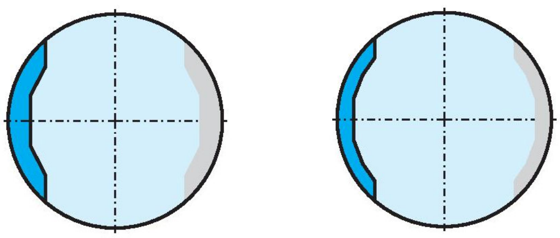

Fig. 12: Multi-chordal downcomers

At the design of multi-chordal downcomers, you have some degrees of freedom. Normally you will find 5 segments as multi-chordal, some-times 7 segments. The construction and fabrication of a multi-chordal tray is more ex-pensive than a chordal design.

SLOPED DOWNCOMER

The gas-liquid mixture entering the downcom-er has the average density of the phases. During the degassing process, the density increases and ideally becomes the liquid den-sity at the bottom of the downcomer. You can take benefit from this change in density (i.e. change in volume) by reducing the cross sec-tional area of the down-comer over its height. This adjustment is called sloped or stepped downcomer (Fig. 13).

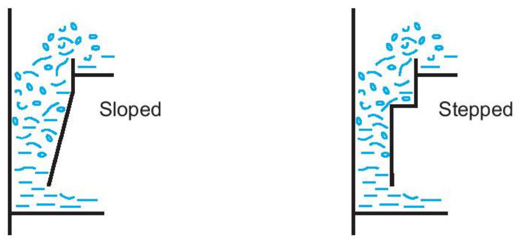

Fig. 13: Sloped and stepped downcomers

It can be applied on all types of downcomers (side, center, off-center, multi-downcomers, ...). The benefit of this design is the gain of active area on the next tray.

TRUNCATED DOWNCOMER

Another possibility to maximize the active ar-ea is to use truncated downcomers (Fig. 14). In this design the floor of the next tray is not the bottom of the downcomer – the downcom-er ends above the active area. For Multi-Down --comers this design is standard. The area below the downcomer can be used (partly) as active area.

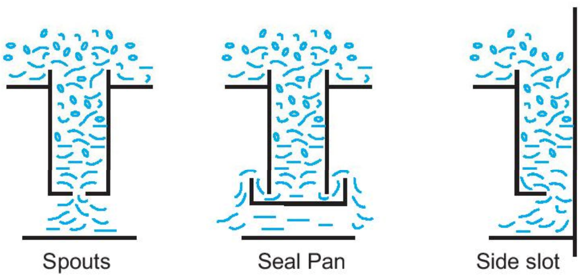

Fig. 14: Truncated downcomers

Truncated downcomers bring us to some fun-damental hydraulic considerations.

DOWNCOMER FLOODING

There are two flooding effects related to down -comers. One deals with the maximum through-put, the other with the liquid level in the down-comer.

The first one is called Choke Flood and is re-lated to the cross-sectional area of the down-comer and the physical properties of the liquid and the gas. If there is not enough area for the degassing (upflowing gas and downflow-ing liquid) the downcomer "chokes". In result, no liquid will pass trough the down-comer and the tower starts flooding.

The standard model for calculating the max. volume flow rate through a downcomer was published by Glitsch in 1993 (here in SI-units):

$$\dot{V}_{DC,Lmax,1} = 0.1698 \cdot SF \cdot A_{DC,top}$$

$$\dot{V}_{DC,Lmax,2} = 0.006955 \cdot \sqrt{\rho_L - \rho_G} \cdot SF \cdot A_{DC,top}$$

$$\dot{V}_{DC,Lmax,3} = 0.0079824 \cdot \sqrt{TS} \cdot \sqrt{\rho_L - \rho_G} \cdot SF \cdot A_{DC,top}$$

$$\dot{V}_{DC,Lmax} = \min(V_{DC,Lmax,1}, V_{DC,Lmax,2}, V_{DC,Lmax,3})$$

The equations are based on gas density rG, liquid density rL, tray spacing TS and system factor SF. This last parameter describes the difficulty of degassing. You can find lists of values for different appli-cations in literature (ref. to References at the end of article).

The second flooding effect is the so-called Aerated Downcomer Flood (also called Down-comer Backup Flood): In this case the liquid level in the downcomer exceeds the total down-comer height. The level in the down-comer is the result of "border effects" (Fig. 15):

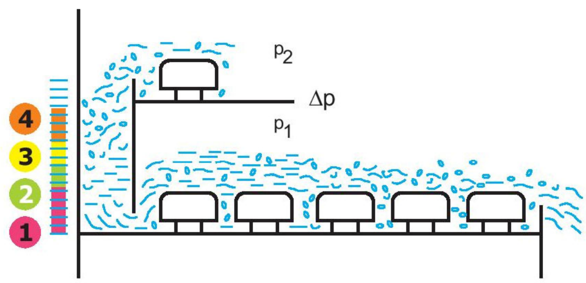

Fig. 15: Liquid level in downcomer

1. The two-phase layer on the next tray seals the downcomer. It is calculated by the physi-cal height of the weir plus the weir crest height.

2. The hydraulic gradient of a tray produces an additional liquid head in the downcomer. (Depends on contact elements and flow path length.)

3. The liquid is accelerated by leaving through the clearance. This rise in kinetic energy is taken from potential energy and results in an additional liquid head.

4. The pressure p1 of the gas is higher than the pressure p2 at the tray above. This difference in pressure results in liquid head in the down-comer.

All these values are calculated as “clear liquid”. In praxis the liquid in the downcomer contains gas (calculated as so-called Aeration Factor) and leads to a higher level in the downcomer than the clear liquid height. If this level exceeds the downcomer height plus weir height, the down-comer will not be able to handle the load and floods.

> NOTE: The effect of Aerated Downcomer Flood is *not* correlated with the downcomer area. Therefore the change of downcomer area will not affect this value!

There are some calculation models for the Aeration Factor in literature (ref. to References at the end of article).

A short glance back: We started the discussion with truncated downcomers. Since the height of the downcomer is relevant for the Aerated Downcomer Flood (and a truncated downcomer is less in height), truncated downcomers are only applicable, if there is no problem with Aerated Downcomer Flood.

OUTLET WEIRS

The liquid-gas mixture enters the downcomer at the outlet weir. The standard weir is a plain bar of about 50 mm height. In liquid limited systems or on trays running in the froth regime, the weir height might be more. The specific weir load should be more than 4.5 m3/m/h (or 5 mm weir crest height). Why?

If the weir crest height is very low, any leveling issues of the tower (due to tower attachments, tower installation or even wind) can cause problems: If a part of the weir is not in use, this part of the active area is stagnant – no mass transfer will happen (Fig. 16).

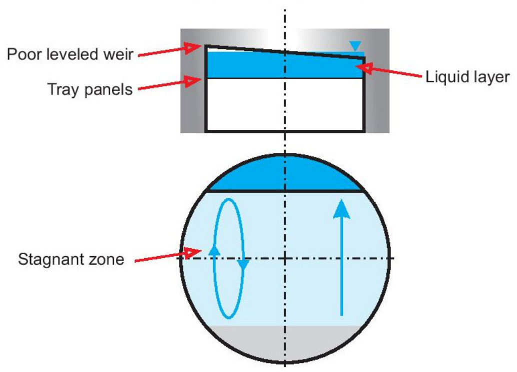

Fig. 16: Malfunction of weir

To achieve a uniform overflow at the entire length of the downcomer, you can use Notched Weirs (Fig. 17). At low loads the liquid uses the bottom parts of the notches. At higher loads the entire weir is used by the liquid.

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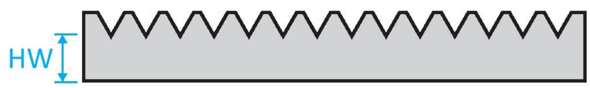

Fig. 17: Notched Weir

For safety reasons (e.g. during installation or inspection) it is strongly recommended to have plain tops at the notches.

Another possibility to adapt long weirs for small liquid loads is to block the weir (Fig. 18). The weir length is reduced to the openings. An additional effect of a blocked weir: it keeps the two-phase layer on the active area.

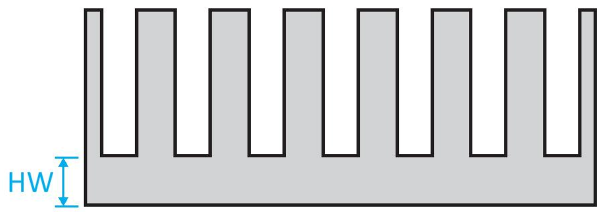

Fig. 18: Blocked Weir

The number and width of the blocks depend on the weir crest height you have to achieve. The method of blocking is not only relevant for small liquid loads, but also for balancing the different weir lengths of 2-and-more-pass layouts. The block should be at least as high as the two- phase layer of the active area.

Blocked weirs prevent workers/inspectors from climbing over the center/off-center downcomers to the other pass! If there is no manhole in each pass, there should be manways in the blocked weir.

Sometimes you will find weirs with adjustable height. These are built in when it is known that the load will change in the future.

ANTI-JUMP BAFFLES

At center, off-center and multi-downcomers, the liquid enters the downcomer from both sides. At high liquid loads and small downcomer widths, you have to install so-called Anti-Jump Baffles (Fig. 19). These baffles ensure, that only half of the area is used by each side.

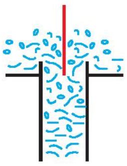

Fig. 19: Anti-Jump Baffle

Just like for the weir blocks, Anti-Jump Baffles have to be as high as the two-phase layer on the active area. Therefore they block the passage during inspections. It is good practice to have bolted manways in the baffles.

SWEPT-BACK WEIR

To achieve a long weir, you can use a so-called Swept-back Weir (Fig. 20). The downcomer stays chordal, but the weir is – similar to the shape of multi-chordal downcomers – longer than the chordal length.

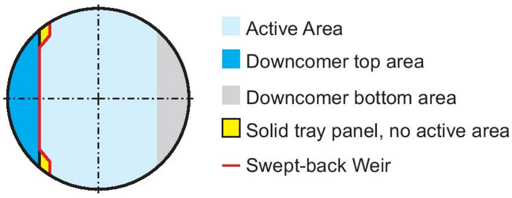

Fig. 20: Swept-back Weir

WEIRS FOR STARTUP

For high capacity trays you will find so-called Weir Spouts (Fig. 21): At tray floor level, there are rectangular openings.

These holes help to fill the downcomer initially at startup of the tower. During operation they help to reduce the weir crest height. They are only used, if the tower is always running at high liquid loads and there is no minimum load issue.

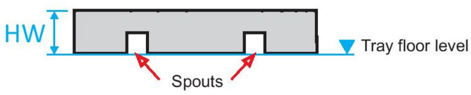

Fig. 21: Weir Spouts

CLEARANCE

The last aspect of downcomers is the design of the liquid outlet (= inlet to the next tray).

The liquid is leaving the downcomer through the so-called Clearance. The outlet velocity of the liquid should be less than 0.45 m/s. If it is higher, the liquid will splash and overrun the first rows of contact elements (see later).

The height of the clearance HCL is normally less than the outlet weir height HW to have a Static Seal

$$\text{Seal}_{\text{static}} = HW - HCL.$$

For liquid-tight trays (e.g. bubble cap or tunnel trays) this seal is present, whenever liquid is on the tray. For all other tray types (e.g. sieve, float valves, fixed valves) the seal is only working, if there is enough gas for no weeping.

If there is no Static Seal (due to design, poor hydraulics or at startup of the tower) there is the risk of gas bypassing the active area through the downcomer.

During operation the weir crest generates additional liquid height (How) (Fig. 22) and generates a Dynamic Seal

$$\text{Seal}_{\text{dynamic}} = How + HW - HCL.$$

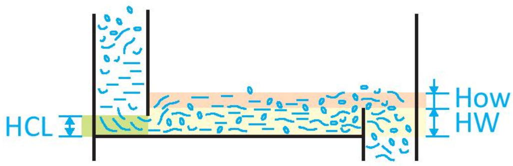

Fig. 22: Static and Dynamic Seal

So-called high capacity designs are often running with no static but only dynamic sealing.

RADIUS LIP

The acceleration of the liquid in the clearance results in liquid head (ref. to Fig. 15, Aerated Down-comer Flood 3). To reduce this effect, you can add a Radius Lip (Fig. 23). This shape lowers the outlet orifice coefficient.

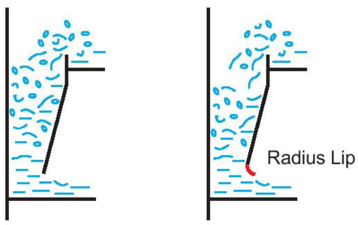

Fig. 23: Radius Lip

INLET WEIR

To ensure sealing of downcomers you can place an Inlet Weir in front of the clearance (Fig. 24). Even at a very small liquid load, there will be no gas bypass through the downcomer. The inlet weir can only function if it is higher than the clearance. It can also be notched (for low liquid loads).

The inlet weir is helpful for the startup of the tower and at low liquid load – but the downcomer tends to blocking in fouling systems.

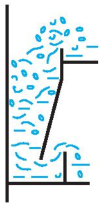

Fig. 24: Inlet Weir

As discussed at the Aerated Downcomer Flood, one part of the liquid level in the downcomer depends on the liquid level on the next tray. For part 1 (Fig. 15) you have to take into account the inlet weir height plus the inlet weir crest height. Therefore an inlet weir "costs" down-comer height in respect to downcomer flooding (especially at low tray spacings).

RECESSED SEAL PAN

The effect of sealing can also be achieved by the so-called Recessed Seal Pan (also called Inlet Pot, Fig. 25). The main advantage over the inlet weir is, that it "costs" no downcomer height. But of course it is more complex in construction and fabrication. It is normally combined with a sloped downcomer.

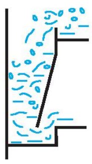

Fig. 25: Recessed Seal Pan

INTERRUPTER BAR

At float valve trays you may sometimes find elements, that are looking similar to inlet weirs. They are not for sealing the downcomer but to keep the first valve row functional (Fig. 26). The bar is about 13mm high and is called Interrupter Bar.

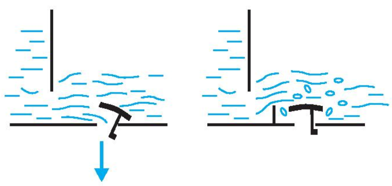

Fig. 26: Interrupter Bar

BUBBLING INITIATORS

At high liquid outlet velocity, the liquid overruns the first rows (of fixed valves, sieve holes, ...). To break the impulse at high capacity trays there are Bubbling Initiators (also called Bubbling Promoters, Fig. 27). They are placed instead of the first row of "normal" contact elements. The gas outlet openings of the Bubbling Initiators are not oriented towards the downcomer (to prevent gas entry through the clearance).

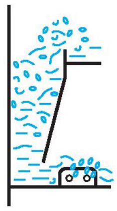

Fig. 27: Bubbling Initiator

INLET PUSHER

A special design ("NYE-tray") for the inlet is to push gas near the clearance through holes to the active area (Fig. 28). It also helps to initiate the two-phase layer. It is not easy to design this layout for a large operating region.

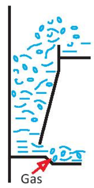

Fig. 28: Inlet Pusher ("NYE-tray")

SEAL PAN

The last downcomer of a tower/section is sealed by a special Seal Pan (Fig. 29) or submerged in the bottom liquid (Fig. 30).

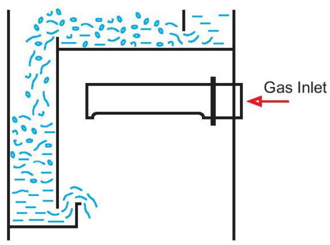

Fig. 29: Last downcomer with Seal Pan

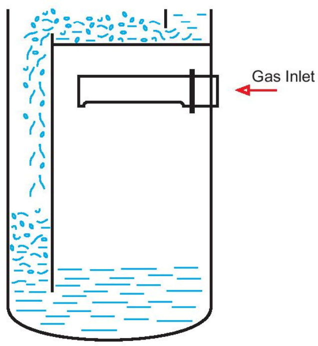

Fig. 30: Submerged downcomer

In any case, the gas inlet should not affect the outlet of the liquid. It is good practice to make the last downcomer long enough to bypass the gas inlet.

DRAW-OFF (DRAW)

You will find seal pans not only at the very last downcomer of a tower. Whenever there is a change in flow path number and/or tray orientation ("transitions"), the liquid is transferred to the next tray (or to a liquid distributor) by pipes. Fig. 31 shows some examples for Draw-Offs from a seal pan.

In all cases you have to ensure that the seal is not affected by the Draw-Off.

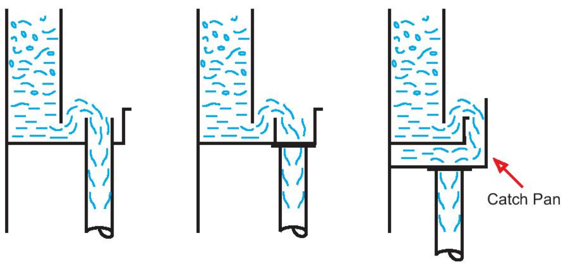

Fig. 31: Draw-Offs at Seal Pan

One of the advantages of tray towers is the possibility to draw off liquid at each stage. These intermediate Draw-Offs are normally done with the help of Recessed Seal Pans (Fig. 32).

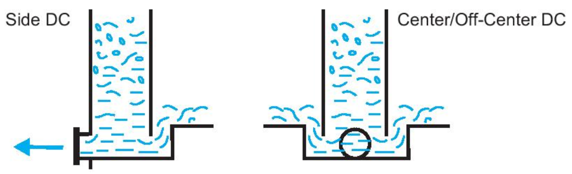

Fig. 32: Draw-Offs at Recessed Seal Pans

FEED TO DOWNCOMER

As mentioned before, you will install so-called False Downcomers (FDC) to feed liquid on the top tray of a section (Fig. 33). They are designed accordingly to the downcomers of the following trays.

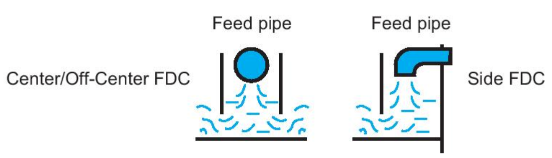

Fig. 33: False Downcomer

To feed liquid to a tray within a section, it is good practice to feed near the downcomer clearance (feed pipe with holes directed to the area above the clearance, Fig. 34). If there is the danger of two-phase or super-heated liquid, you will shield the downcomer wall by a so-called Impingement Baffle.

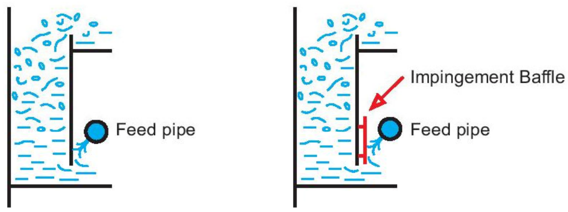

Fig. 34: Feed to tray

MECHANICAL ASPECTS

The active area is carried by the support ring, the downcomer (at the outlet weir side) and the downcomer seal segment (at the inlet). There-fore these downcomer elements have to be designed to withstand the dead load plus the liquid load.

For large tower diameters, the upper part of

the downcomers (so-called Downcomer Truss) will be fabricated in a higher material thickness (Fig. 35).

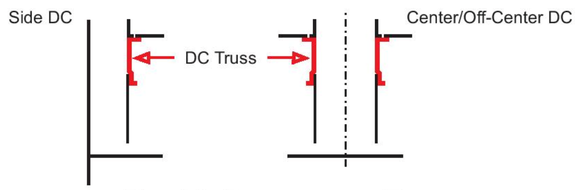

Fig. 35: Downcomer Truss

Whenever the downcomer elements would not fit through the manhole, the elements have to be divided, too. (A benefit of a downcomer wall consisting of two parts is the opportunity to adjust the clearance height during installation.)

In downcomers there are used so-called Down-comer Brackets (Fig. 36). They support the downcomer seal segment and prevent the walls from vibration. In center/off-center downcomers the brackets additionally connect the walls and keep them in place – in best case even at a pressure surge. (A well-known failure pattern is when the downcomer walls are pushed inwards after a pressure surge and the panels of the active area therefore slip and fall.)

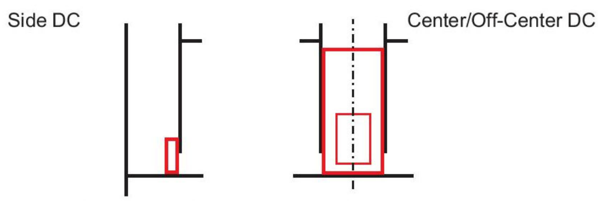

Fig. 36: Downcomer Brackets

CALCULATED PARAMETERS

The following parameters have to be calculated and checked for all downcomers at each load. This can hardly be done by hand, but with suitable software. Any software that does not show all of these parameters carries the risk of overlooking relevant values. Since the supplier's free software does not do this, commercial software (ref. TrayHeart) should be used.

Parameters of the downcomer:

- Residence Time
- Clear liquid level in downcomer
- Aeration Factor of downcomer
- Aerated liquid level in downcomer
- Liquid head caused by clearance
- Liquid outlet velocity
- Max. liquid volume flow through downcomer

For the outlet weir the following parameters have to be calculated:

- Specific liquid load of weir
- Weir crest height
- Throw width over weir
- Liquid ratio through spouts

If there is an inlet weir:

- Specific liquid load of weir
- Weir crest height

CONCLUSION

There are many layouts and designs for downcomers and weirs. The effort for the engineer is to combine the best options to achieve an optimum for the very different aspects as startup, statics, hydraulics and interaction with the active area. The hydraulic design of downcomers has to be as accurate as the design of the active area.

ABOUT THE AUTHOR

Volker Engel studied process engineering at the Technical University of Munich and did his Ph.D. thesis on packed columns with Prof. Johann G. Stichlmair. Since 1998 he has been the managing director of WelChem Process Technology GmbH and head of the TrayHeart software. TrayHeart has developed into a state-of-the-art design tool for trays and internals in process technology.

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Internet: www.welchem.com; Info: service@welchem.com


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Cold Eye Review of Project Progress or Any Situation

Lalit Mohan Nainwal, C Eng. (India), MBA

Since everything is person to person, everything counts or nothing counts at all. Many times whether it is in projects, events, incidents or a survey one has to go through circumstances where he/she has been asked or he/she is required to report/state exactly what he has seen/reviewed. One is expected to state facts without adding anything in it due to his/her previous experience, existing knowledge, or added knowledge by others while viewing/reviewing the situation. One is neither expected to exaggerate nor understate. A statement of facts or report which states how the progress/situation is as of that moment of visit/review without being judgemental or emotionally attached at all is a cold eye review report.

1. History

Historical book Mahabharata in India has a record of first cold eye review. A battle, which was being narrated by Sanjay to a blind king, namely Dhritarashtra, without being attached to any of both parties indulged in battle. Sanjay was the person with the capability to see what is going on at the battleground and he has to state the facts to the king. In short cold eye review means an emotionless review of any situation or any work in progress.

2. Why it's termed Cold Eye Review

The review report author or originator is expected to be an experienced professional/ subject matter specialist in whatever field the review being performed, his vision without attachment is expected in the final submission, his review namely cold eye review ignores advice and search facts beyond appeasement. Cold eye here is something to tell the reader of that report that this is what I saw exactly against given scope, specs, drawings, or details in any form.

3. How Cold Eye Review is a necessity in Projects

An unfortunate fact that false reporting always occurs, there are several reasons, excuses and misunderstandings why it occurs, let's not Management seeks the exact report of any project's progress at various stages/phases let it be Engineering, Procurement, Execution/Construction, and Completion.

Let's go in details of progress on each phase,

3a. Engineering Progress Reporting

An expert is asked to go to an engineering contractor's office or at a consultant's office, the first thing will come to anybody's mind that how can someone review the actual progress when most of the things are in soft form in computers of various designer or engineering function heads?

answer is, the expert will have a schedule and expected deliverables list from engineering with him and he will have a clear understanding of the job before he proceed to review. Once the review started, he will note down the facts against schedule & deliverables not expectations or statements of team members who are working on that job.

3b. Procurement Progress Reporting

When an expert is about to review procurement progress, he will be informed by so many sources, we had MRs ready, this was floated to vendors, that was EQ or CQ from vendors, but an expert is expected not to get involved in any opinions and check against agreed procurement schedule baselines whether the progress is made as required and where exactly the procurement stands on the day of review.

3c. Execution/Construction Progress Reporting

This is the most complicated work in case someone is requested to ascertain the correct progress in huge projects or someone is expected to provide cold eye review of a huge event. The expert will find many works are ongoing and will be dragged by many on the percentage progress they want to report. The cold eye review involved experienced professional/subject matter expert will have to isolate himself from influences and check reality on the ground against agreed schedule or time frame, quality standards, or other relevant documents.

3d. Completion Progress Reporting

Unfortunate that many works in projects or huge events marked verbally complete but so many things remain to be done, I am not going into punch listing here but an expert shall keep in mind that complete means complete in all sense. In cold eye review the EP/SME will check it against required inspection records vs signed off records, verifying things physically by himself.

4. How can management ask Cold Eye Review on the prevailing situation

Experienced Professional / Subject Matter Expert is always asked to help on finding facts and to the best understanding of facts, management uses it to compare with what is reported. Normally from any event or work there will be many sources providing information to many heads at the same time. An execution head will get reports from his project involved team that so and so activity is complete but the quality team is not coming to inspect, at the same instance quality team's head will be appraised that execution/construction has not yet completed the work and raised request to inspect. There will be ambiguity and cross fights among departments and professionals. Management needs to carefully choose a person who can have a viewpoint not inclined to any team when he is on his mission to produce a cold eye review report.

5. Cold eye review is not punch listing, why it's so & how?

Cold eye review is a detailed document that provides insight into the current situation and has no influence on upcoming works or preparations. The Cold eye review for a work if seen incomplete will mark it incomplete even if it's argued that whole preparation is done and work will be completed within a short time. Its not punch list which can mark work as partially complete with few comments open.

6. Outputs of Cold Eye Reviews

Cold eye review is expected to deliver a crystal clear picture of a project progress or ongoing event where does it stand when the study/review was done.

7. Discussion / Reporting findings of cold eye review

To avoid conflict and in an attempt to make things work as per plan again, the cold eye review report will be taken up for discussion with senior level team direct related to or involved in execution. The management put forward findings to have a responsibility matrix developed and targets set again to close the findings. Management can have an insight where were the lapses and can correct from their end too.

8. Action plan for implementation of cold eye review findings

Once discussion with all concerned completed, the detailed implementation plan will be generated and all responsible will have a say before the action plan is finalized. All lapses or grey areas where no one seems taking the work can be redistributed to team members and the action plan is finalized and implemented.

9. Conclusion

Cold Eye Review is key to success for the complicated work which has huge volumes and huge teams working. It brings the efforts toward a single goal. Cold eye review is a hot work to be completed when nothing else seems to give correct reports.

ABOUT THE AUTHOR

Lalit Mohan Nainwal has always strived for knowledge, Lalit continued his education with his career as a shipwright apprentice and achieved his AMIE(Mech), C Eng (India) & MBA. Apart from Indian Navy's Docks, Lalit worked with several renowned EPC companies in India & abroad such as Tata Projects, SKE&C, and KBR Inc. in major projects such as HRRL, KPPC Aromatics, Pearl GTL, and Ichthys LNG Project.

Evaluating Pipeline Operational Integrity | Sand Production

Jayanthi Vijay Sarathy

Piping systems associated with production, transporting oil & gas, water/gas injection into reservoirs, experience wear & tear with time & operations. There would be metal loss due to erosion, erosion-corrosion and cavitation to name a few. The presence of corrosion defects provides a means for localized fractures to propagate causing pipe ruptures & leakages. This also reduces the pipe/pipeline maximum allowable operating pressure [MAOP].

The following article covers methods by DNV standards to quantitatively estimate the erosion rate for ductile pipes and bends due to the presence of sand. It is to be noted that corrosion can occur in many other scenarios such as pipe dimensioning, flow rate limitations, pipe performance such as pressure drop, vibrations, noise, insulation, hydrate formation and removal, severe slug flow, terrain slugging and also upheaval buckling. However all these aspects are not covered in this document.

Based on the erosional rates of pipes and bends, the Maximum Safe Pressure/Revised MAOP is evaluated based on a Level 1 Assessment procedure for the remaining strength of the pipeline. The Level 1 procedures taken up in this article are RSTRENG 085dL method, DNVGL RP F-101 (Part-B) and PETROBRAS's PB Equation.

GENERAL NOTES & ASSUMPTIONS

1. In evaluating corrosion defects, the generally accepted or traditional approach is the ASME B31G code which gives overly conservative results in terms of lower burst pressures with which operators repair/replace the corroded pipe/pipeline segments. This represents higher maintenance costs necessitating the need to follow a procedure that meets pipeline integrity requirements while also lowering maintenance, repair & replacement costs.
2. To assess pipeline integrity, standard corrosion assessment procedures are classified on three levels – Level 1, Level 2 and Level 3. Level 1 procedure represents longitudinal area of metal loss based on the maximum defect depth and overall defect length. The ASME B31G, RSTRENG 085dL, and DNVGL RP F-101 method for single defect can be classified as Level 1 methods. Level 2 procedure represents longitudinal area of metal loss based on the defect depth profile. The RSTRENG Effective Area method and DNVGL RP F-101 method for complex shaped defects can be classified as Level 2 methods. Level 3 assessment methods involve using Finite element methods (FEM) provided the FEM model is validated against experimental results.
3. Corrosion failures are caused by two main mechanisms – Leakage resulting in a relatively small loss of product and Rupture causing a sudden release of pressure which propagates in isolation.
4. To understand corrosion assessment procedures, two terms come into play – Folias Bulging factor [MT] and flow stress [sf]. Folais factor represents the bulging effect of a shell surface that is thinner in wall thickness [WT] than the surrounding shell. It takes into account the work-hardening effect, i.e., the increase in the stress concentration levels as the corrosion defect begins to bulge before eventually causing a failure. The flow stress is the stress at which the corrosion defect is predicted to cause a failure.
5. In pipeline assessment literature, SMYS & yield strength are used differently. Specific Minimum Yield Strength (SMYS) is the absolute minimum yield strength for a particular material grade specified by ASTM standards. Whereas, yield strength is obtained from mill conducted tensile tests. In cases, where the yield strength value is not available, SMYS can be used instead.

6. When a corrosion defect occurs inside a pipe/pipeline, the defect tends to propagate longitudinally. ASME B31G mandates a maximum allowable longitudinal length [LM] for a given defect depth [d]. As per Modified ASME B31G method i.e., 085dL method, defects are classified as Long defect and short defect based on the condition, LS2/Dt = 50, Where D = Pipeline Outer diameter (OD) and t = pipeline nominal wall thickness. When field measured defect's longitudinal length, L < LS, the defect is termed as short defect. When L > LS, the defect is termed as long defect. The DNVGL RP F-101 method does not classify defects in relation to their longitudinal length. The pressure strength of long defects is a function of the longitudinal defect length [L]. The Longer the defect, lower is the failure pressure However a limit exists in the value of L, beyond which any large increase in the longitudinal defect length, L produces very little reduction in the failure pressure.

7. Long Internal defects are one of the various causes for geometry corrosion induced damage that occur in oil & gas pipelines. These occur on the pipe/pipeline bottom due to accumulation of liquids including water. Whereas long external defects are caused on the pipeline's outer surface due to loss of protective coatings.

8. ASME B31G assumes a parabolic profile across the area of the defect, i.e., Area of defect = 2/3´d´L, where, d = Defect depth and L = Defect longitudinal length. Whereas with the RSTRENG 085dL method, the defect area is approximated as 85% of the peak depth, i.e., by using a factor of 0.85, i.e., Defect Area = 0.85´d´L.

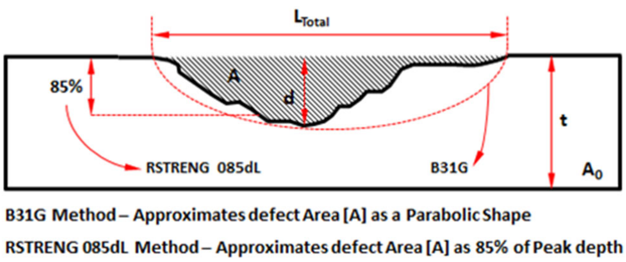


Figure 1. Corrosion Shape Approximation

9. The potential for sand particles to get carried from the formation to well bore in oil & gas wells is subjected to the reservoir geology. With the onset of water formation or rapid change in well conditions, there is sand formation. Employing a zero rate of sand production would be economically infeasible. Therefore sand management programmes are put in place whereby upstream facilities are equipped with sand traps with necessary safeguards that aid in achieving an acceptable sand rate. The standard used for this article is DNVGL RP O501 which provides empirical models that cover plain erosion & not the combined effects of corrosion-erosion, droplet erosion & cavitation. The article therefore considers plain erosion which leads to corrosion pits in the pipeline & the associated MAOP is computed using the standard corrosion assessment methods.

10. When applying the original ASME B31G method in simplified form (Appendix L of ASME B31.8), the Safe Operating Pressure given as P' must first be calculated using the pressure corresponding to a hoop stress equal to 100% of SMYS for the operating pressure, P. The resulting P' is the estimated failure pressure, which must then be divided by the design factor/desired factor of safety to obtain the correct P'.

CASE STUDY

A case study is made for a case of 30 MMscfd of well fluids transported through an 8" DN carbon steel flowline from the well head to a trunk line. The design details are,

Parameter	Value	Unit
Operational Life of Pipeline	25	Years
Pipeline Diameter [DN]	8.625	in
Pipeline Wall Thickness [WT]	3.18	mm
API 5L Spec	PSL1 X65	-
Ultimate Tensile Strength [s_u]	530	MPa
SMYS [S]	448	MPa
Design Pressure [DP]	93.5	bara
Design Temperature [DT]	100	^{0}C
Gas Flow Rate [m_g]	31,657	kg/h
Liquid Flow Rate [m_l]	14,928	kg/h
Gas Density [r_g]	42.0	kg/m^3
Liquid Density [r_l]	713.2	kg/m^3
Gas Viscosity [μ_g]	1.34E-05	kg/m.s
Liquid Viscosity [μ_l]	4.72E-04	kg/m.s
Mixture Viscosity [μ_m]	2.58E-05	kg/m.s
Gas Velocity [V_g]	5.9	m/s
Liquid Velocity [V_l]	0.16	m/s
Sand Content [ppmW]	50.0	ppmW
Average Sand Particle Size	300	µm
No. of Pipe Diameter [90^0 Long Elbow]	1.5	[-]
Inclined Pipe Impact angle [a]	30^0	degrees

WELL FLUIDS MIXTURE PROPERTIES

The mixture density and mixture viscosity of the well fluids can be determined as follows.

$$\rho_m = \frac{\rho_g V_g + \rho_l V_l}{V_g + V_l} \quad (1)$$

$$\mu_m = \frac{\mu_g V_g + \mu_l V_l}{V_g + V_l} \quad (2)$$

Therefore applying the above correlations,

$$\rho_m = \frac{[42 \times 5.9] + [713.2 \times 0.16]}{5.9 + 0.16} = 60.2\,\frac{kg}{m^3}$$

(3)

$$\mu_m = \frac{[0.0000134 \times 5.9] + [0.000472 \times 0.16]}{5.9 + 0.16} = 0.0000258\,\frac{kg}{m.s}$$

(4)

INCLINED PIPE EROSION RATE – DNV RP O501

The flowline profile over the terrain would have inclined sections. With the onset of water production from the wells, quartz sand particles from wells [50 ppmW, 300 µm, 2,650 kg/m3] impinge at an impact angle of 300. As per DNVGL RP O501 [Rev. 2015], for ductile materials, the maximum erosion occurs for impact angles in the range of 150 to 300, whereas brittle materials experience maximum erosion at normal impact angle. The erosive wear can be estimated as,

$$E = \frac{m_p \times K \times U_p^n \times F(\alpha)}{\rho_t \times A_t} \times [3.15 \times 10^{10}]$$

(5)

Where,

m_p = Sand Flow rate [kg/s]

K=Material Constant (2´10^{-9} for Steel Grades)

n =Material Constant (2.6 for Steel Grades)

U_p = Particle Velocity [m/s] (V_g + V_l)

r_t = Pipeline density [kg/m^3]

A_t = Pipeline Area exposed to Erosion [m^2]

F(a) = Function characteristic of ductility [-]

The value of F(a) is calculated as,

$$F(\alpha) = 0.6 \times \left[Sin(\alpha) + 7.2\left(Sin(\alpha) - Sin^2(\alpha)\right)\right]^{0.6} \times [1 - e^{-20\alpha}]$$

(6)

For the condition, F(a)Î [0, 1] for a Î [0, p/2]

Note: 1 mil = 1/1000th of an inch

The sand flow rate based on ppmW is calculated as,

$$m_p = \frac{[m_g + m_l] \times ppmW}{10^6} \quad (7)$$

The erosion rate can be calculated beginning with estimating the function characterizing pipeline ductility, F(a) as follows,

For 30^0, $\frac{\alpha\pi}{180} = \frac{30\pi}{180} = 0.5236$ (8)

$$F(\alpha) = 0.6 \times [Sin(0.5236) + 7.2(Sin(0.5236) - Sin^2(0.5236))]^{0.6} \times [1 - e^{-20\times 0.5236}] = 0.9890$$
(9)

The sand flow rate based on ppmW is,

$$m_p = \frac{[31{,}657 + 14{,}928] \times 50}{10^6} = 2.33\ kg/h \quad (10)$$

The pipeline area exposed to erosion is,

$$A_t = \frac{A_p}{Sin(\alpha)} = \frac{0.0355}{Sin(30)} = 0.0711\ m^2 \quad (11)$$

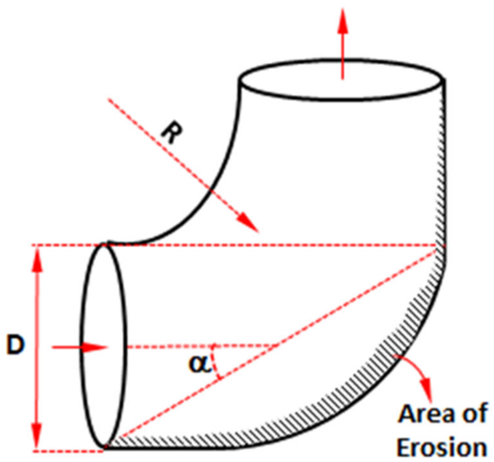
The erosion rate is therefore calculated as,

$$E = \frac{2.33 \times [2\times 10^{-9}] \times [5.9 + 0.16]^{2.6} \times 0.989 \times [3.15 \times 10^{10}]}{3600 \times 7{,}800 \times 0.0711}$$
(12)

$$E = 0.0078\ mm/y\ (or)\ 0.31\ mils/y \quad (13)$$

PIPE BEND EROSION RATE – DNVGL RP O501

Pipeline bends are prone to erosional wear. When the flow direction in the bend changes, sand particles crash against the bend wall, instead of following the flow direction. Assuming a straight length [10D] before the bend, the erosion rate is estimated as,



The characteristic impact angle, a for the pipe bend geometry is calculated from the radius of curvature. The radius of curvature, R_c [i.e., bend radius] for a bend is expressed as number of pipe diameters. Considering a 90^0 long elbow, the bend radius in terms of number of pipe diameters is 1.5, i.e., R_c = 1.5.

$$\alpha = Tan^{-1}\left[\frac{1}{\sqrt{2R_c}}\right]$$
(14)

$$\alpha = Tan^{-1}\left[\frac{1}{\sqrt{2\times 1.5}}\right] = 0.5236\ rad\ or\ 30°$$
(15)

The length of the 90^0 bend is estimated as,

$$L_{Bend} = \frac{\theta}{360} \times 2\pi R_c, Where\ \theta = 90°$$
(16)

$$L_{Bend} = \frac{90}{360} \times 2\pi \times 1.5 \times 8.625 = 20.3in$$
(17)

As per DNVGL RP O0501, the erosional rate [E] for pipe bends is computed as,

$$E = \frac{m_p \times K \times U_p^n \times F(\alpha) \times G \times C_1 \times GF}{\rho_t \times A_t} \times [3.15 \times 10^{10}]$$
(18)

Where,

E = Erosion Rate [mm/year]

m_p = Sand Flow rate [kg/s]

K=Material Constant (2´10^{-9} for Steel Grades)

n =Material Constant (2.6 for Steel Grades)

U_p = Particle Velocity [m/s] (V_g + V_l)

r_t = Pipeline density [kg/m^3]

A_t = Bend Area exposed to erosion [m^2]

$F(a)$ = Function characteristic of ductility [-]

G = Particle Size correction function [-]

C_1 = Constant accounting for multiple impact of sand particles at the bend's outer end [2.5]

GF = Geometry Factor

The geometry factor [GF], is taken to be 1.0 based on the assumption that the straight line section, upstream of the bend is greater than 10D. For straight line section less than 10D, the GF increases to 2 or 3. To arrive at the particle size correction term, G, the dimensionless parameter A is calculated first.

$$A = \frac{\rho_m^2 \times Tan(\alpha) \times U_p \times D}{\rho_p \times \mu_m} \tag{19}$$

The diameter relation [g] and critical diameter relation [g_g] is calculated as,

$$\gamma = \frac{d_p}{ID} \tag{20}$$

Where, d_p = Average Particle diameter

$$\frac{d_{p,c}}{ID} = \gamma_c = \begin{cases} \frac{\rho_m}{\rho_p[1.88ln(A) - 6.04]} & if\ 0 < \gamma_c < 1 \\ 0.1 & if\ \gamma_c \leq 0\ or\ \gamma_c \geq 0.1 \end{cases}$$

(21)

The particle size correction function [G] is,

$$G = \begin{cases} \frac{\gamma}{\gamma_c} & if\ \gamma < \gamma_c \\ 1 & if\ \gamma \geq \gamma_c \end{cases} \tag{22}$$

The pipeline Bend Area exposed to erosion is,

$$A_t = \frac{A_t}{Sin(\alpha)} \tag{23}$$

Applying the expressions to the case study,

$$A = \frac{60.2^2 \times Tan(30) \times 6.1 \times 0.2127}{2{,}650 \times 0.0000258} = 39{,}315 \tag{24}$$

$$\gamma = \frac{300}{0.2127 \times 10^6} = 0.0014 \tag{25}$$

$$\gamma_g = \frac{60.2}{2{,}650 \times [1.88ln(39315) - 6.04]} = 0.0016 \tag{26}$$

Therefore since g < g_g, the particle size correction function is,

$$G = \frac{\gamma}{\gamma_g} = \frac{0.0014}{0.0016} = 0.8601 \tag{27}$$

The critical particle diameter [$d_{p,c}$] is calculated as,

$$d_{p,c} = ID \times \gamma_c = \frac{0.2127 \times 0.0016}{10^6} = 349\mu m \tag{28}$$

The pipeline area exposed to erosion is,

$$A_t = \frac{A_p}{Sin(\alpha)} = \frac{0.0355}{Sin(30)} = 0.0711\ m^2$$

(29)

The function characterizing pipeline ductility, F(a) as follows,

For 30^0, $\frac{\alpha\pi}{180} = \frac{30\pi}{180} = 0.5236$ (30)

$$F(\alpha) = 0.6 \times \left[Sin(0.5236) + 7.2\left(Sin(0.5236) - Sin^2(0.5236)\right)\right]^{0.6} \times [1 - e^{-20 \times 0.5236}] = 0.9890$$

(31)

Therefore the erosion rate is computed as,

$$E = \frac{2.33 \times [2 \times 10^{-9}] \times [6.1]^{2.6} \times 0.989 \times 1 \times 2.5 \times 0.86 \times [3.15 \times 10^{10}]}{3600 \times 7{,}800 \times 0.0711}$$

(32)

$$E = 0.0496\ mm/y\ (or)\ 1.95\ mils/y \tag{33}$$

MAX SAFE PRESSURE IN CORRODED AREA

With sand erosion occurring due to sand flow, defects begin to form on the pipeline inner surface. These defects have a certain depth of penetration [d] for a given wall thickness of the pipe [t]. The following section provides calculations for the maximum safe pressure for operation based on RSTRENG 085dL method, DNVGL RP F101 Single defect method and PETROBRAS PB method.

As per ASME B31G, for a pit depth of up to 10%, the pipeline can be continued to be operated with the existing MAOP. For a pit depth between 10% and 80%, the pipeline needs to be operated at the revised/reduced MAOP based on the corroded wall thickness. For a pit depth greater than 80%, the pipeline would have to repaired or replaced.

As per ASME B31G, for a contiguous corroded area having a maximum depth of more than 10% but less than 80% of the nominal pipe wall thickness, L_m should not extend along the longitudinal axis of the pipe for a distance greater than calculated from the expression,

$$L_m = 1.12B\sqrt{Dt} \quad (34)$$

Where,

L_m = Maximum Allowable Longitudinal length of corroded area [in]

D = Pipeline OD [in]

T = Pipeline selected Wall thickness [in]

The constant B is estimated as,

$$B = \sqrt{\left[\frac{\left(\frac{d}{t}\right)}{1.1\left(\frac{d}{t}\right)-0.15}\right]^2 - 1} \quad (35)$$

As per ASME B31G, B cannot be > 4.0. For corrosion depth [d/t] between 10% and 17.5%, the value of B is to be limited to 4.0. For e.g., with d/t = 0.32, the value of B & L_m is,

$$B = \sqrt{\left[\frac{0.32}{1.1(0.32)-0.15}\right]^2 - 1} = 1.23 \quad (36)$$

$$L_m = 1.12 \times 1.23\sqrt{8.625 \times 0.125} = 1.43\ in \quad (37)$$

RSTRENG 085DL METHOD

The max safe pressure with RSTRENG 085dL method is determined as follows,

$$P_f = \frac{2t[SMYS+69]\times 145.04\times F\times E\times T}{D\times 14.5}\left[\frac{1-0.85\left(\frac{d}{t}\right)}{1-\left[0.85\left(\frac{d}{t}\right)M^{-1}\right]}\right] \quad (38)$$

Where,

SMYS = Specific Min Yield Strength [MPa]

D = Pipeline OD [in]

M = Folias Bulging Factor [-]

For the condition, L^2/Dt £ 50, M is,

$$M = \sqrt{1 + 0.6275\left[\frac{L^2}{Dt}\right] - 0.003375\left[\frac{L^2}{Dt}\right]^2} \quad (39)$$

For the condition, L^2/Dt > 50, M is,

$$M = 3.3 + 0.032\left[\frac{L^2}{Dt}\right] \quad (40)$$

In this case, the measured max corroded area depth [d] and measured longitudinal length [L] in the inclined pipe is 0.04” and 3” respectively. For the selected wall thickness of 3.18 mm, d/t is 0.32, i.e., 32% pit depth.

Similarly for the pipe bend, the measured max corroded area depth [d] and measured longitudinal length [L] is 0.06” and 1.3” respectively. For the selected wall thickness of 3.18 mm, d/t is 0.48, i.e., 48% pit depth.

Therefore for the inclined pipeline,

$$\frac{L^2}{Dt} = \frac{3^2}{8.625\times 0.125} = 8.35 < 50 \quad (41)$$

Since L^2/Dt < 50, the Folias bulging factor is,

$$M = \sqrt{1 + [0.6275 \times 8.35] - 0.003375[8.35]^2} \quad (42)$$

$$M = 2.45 \quad (43)$$

Therefore the max safe pressure is,

$$P_f = \frac{2\times 0.125[448+69]\times 145.04\times 0.72\times 1\times 1}{8.625\times 14.5}\left[\frac{1-(0.85\times 0.32)}{1-[0.85\times 0.32\times 2.45^{-1}]}\right] \quad (44)$$

$$P_f = 88.4\ bara \quad (45)$$

Performing similar calculations for Pipe Bend with d = 0.06” and L=1.3”, the Max safe pressure is 90.0 bara.

DNV RP F101 SINGLE DEFECT METHOD

The max safe pressure with DNVGL RP F101 single defect method is determined as,

$$P_f = \frac{2\times t\times \sigma_u\times 145.04\times F\times E\times T}{[D-t]\times 14.5}\left[\frac{1-\left(\frac{d}{t}\right)}{1-\left[\left(\frac{d}{t}\right)M^{-1}\right]}\right] \quad (46)$$

Where,

s_u = Ultimate Tensile Strength [MPa]

D = Pipeline OD [in]

M = Folias Bulging Factor [-]

$$M = \sqrt{1 + 0.31\left[\frac{L^2}{Dt}\right]} \quad (47)$$

Applying the DNVGL RP F101 Single defect method to the same inclined pipe and pipe bend data for an ultimate tensile strength of 530 MPa, the max safe pressure is,

$$M = \sqrt{1 + [0.31 \times 8.35]} = 1.89 \quad (48)$$

$$P_f = \frac{2\times0.125\times530\times145.04\times0.72\times1\times1}{[8.625-0.125]\times14.5}\left[\frac{1-1.89}{1-[1.89\times8.35^{-1}]}\right]$$

(49)

$$P_f = 91.9\ bara \quad (50)$$

Performing similar calculations for pipe bend with d=0.06” and L=1.3”, the Max safe pressure is 96.3 bara.

PETROBRAS PB METHOD

The max safe pressure with PETROBRAS PB method is determined as,

$$P_f = \frac{2\times t\times\sigma_u\times145.04\times F\times E\times T}{[D-t]\times14.5}\left[\frac{1-\left(\frac{d}{t}\right)}{1-\left[\left(\frac{d}{t}\right)M^{-1}\right]}\right] \quad (51)$$

Where,

s_u = Ultimate Tensile Strength [MPa]

M = Folias Bulging Factor [-]

$$M = \sqrt{1 + 0.217\left[\frac{L^2}{Dt}\right] - \frac{1}{1.15\times10^6}\left[\frac{L^2}{Dt}\right]^4} \quad (52)$$

Applying the PETROBRAS PB method to the same inclined pipe and pipe bend data for an ultimate tensile strength of 530 MPa, the max safe pressure is,

$$M = \sqrt{1 + [0.217 \times 0.835] - \frac{0.835^4}{1.15\times10^6}} = 1.68$$

(53)

$$P_f = \frac{2\times0.125\times530\times145.04\times0.72\times1\times1}{[8.625-0.125]\times14.5}\left[\frac{1-1.68}{1-[1.68\times8.35^{-1}]}\right]$$

(54)

$$P_f = 94.3\ bara \quad (55)$$

Performing similar calculations for Pipe Bend with d = 0.06” and L=1.3”, the Max safe pressure is 99.7 bara.

RESULTS

Summarizing, the max safe pressure is 88.4 bara for pipeline and 90 bara for pipe bend,

Method	Max Safe Pressure, P_f [bara]	
-	Inclined Pipe	Pipeline Bend
RSTRENG 085dL	88.4	90.0
DNV RP F101 Single Defect	91.9	96.3
Petrobras PB	94.3	99.7
Design Pressure [DP]	93.5	93.5
Max Safe Pressure, P_f	**88.4**	**90.0**

Based on the erosion rate for an operating period of 25 years, the pipeline WT lost is,

Parameter	Inclined Pipe	Pipeline Bend
Erosion Rate [mm/y]	0.0078	0.0496
WT Lost in 25 Years [mm]	0.20	1.24

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TrayHeart
Tower Internals Design
TrayHeart is a professional software that performs hydraulic calculations for all types of tower trays, random and structured packings and liquid distributors. The development of TrayHeart started in 1998 and was continued jointly by universities, companies of the chemical industry and tower internals suppliers. TrayHeart ...
is based on multiple calculation models and large databases of packings, float valves, fixed valves, bubble caps, and liquid distributor templates
is a supplier-independent tool. There are no preferred product placements or promoted designs
considers static dimensions, manways and fastenings
offers an interactive 3D-view for all designs
can be used for single stage, profile and data validation calculations
has a unique, logical and multi-lingual user interface, with multiple input and output options
applies hundreds of online queries to check the feasibility and limits of the calculated designs
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Safety Talk Series

Electric Shock Hazard Prevention

Chris Palmisano, MESH, IFAC

This month we are excited to be rolling out our NEW Safety Talk Series! Every other month, we will provide a safety talk for our readers on a new topic. They can be used as resource to share with your employees as a Safety Talk or Tool Box Talk, to help you maintain a safe and healthful workplace.

Today's topic is one:

ELECTRIC SHOCK HAZARD PREVENTION

There are several precautions against accidental grounding that we all should observe when using portable electric tools. Check your tools for these conditions:

- Defective or broken insulation
- Improper or poorly made connections to terminals.
- Broken or otherwise defective plugs.
- Loose or broken switches.
- Sparking brushes.

If any of these conditions exist, have the tool repaired before using it, or report it to your supervisor. Do not use the defective tool.

When it comes to electrical safety, don't attempt to repair or adjust portable electric tools while they're plugged in. Do not use portable electric tools in the presence of water, flammable vapors or gases, unless the tool is specifically designed for such use.

Some people believe that low voltage shocks can't harm them. Actually, these low voltage jolts can be fatal. The severity of a shock is measured by three factors—the quantity of current flowing through the body; the path of the current as it passes through the body; and the duration of the current.

Faulty tools can be responsible for an accident. Tools should receive proper care so they will not become faulty. They should always be returned to their proper place, should be handled with care, and should be inspected regularly.

To reduce the hazard of electric shock, third-wire grounded or approved double-insulated tools must be used. Any extension cords you use must have three-pronged plugs. When repairing extension cords make sure the polarity is correct

There are three factors involved in accidental grounding mishaps which should be recognized. All of these factors are contributed by people and they are:

- Lack of knowledge of safety precautions,
- Ignoring hazards or safety rules
- Neglect.

Again, it's important to check your tools before using them. If they appear broken, defective, or in poor condition, report it to your supervisor. Do not use the tool until it has been approved for further use.

STAY ALERT ON THE JOB AND DO NOT TAKE ANY UNNECESSARY CHANCES

Pneumatic Dilute Phase Solids Conveying

Joe Bonem

INTRODUCTION

Solids conveying is used throughout the industrial and business world. It is used from drive-through banking depositing to heavy industry. While the pneumatic design can vary from dilute phase to dense phase and from vacuum systems to pressure system, this paper deals with pressurized dilute phase system concepts. These concepts will be helpful in both design and problem solving areas.. The paper contains guidelines, conclusions and the results of calculations. These calculated values are based on proprietary correlations. These proprietary correlations have been validated in commercial operations of a pressurized transfer system with a length of 400 feet as well as limited laboratory data. Wherever possible the correlations have been confirmed by actual plant data.

The physical properties of the solid being conveyed are those of a spherically shaped polyolefin. It is not known to what extent the correlations are valid for other solids. The paper is focused on systems where the gas is either nitrogen or air. However, the concepts are valid for other gases. The dilute phase pressurized system is generally one where the solid to gas ratio is less than 5 pounds of solids per pound of gas. Although there are literature references of a solid to gas ratio being as high as 10. Pressurized systems are defined as a system where the final pressure is above atmospheric pressure. The initial pressure will then depend on pressure drop in the conveying system. An example of this system is shown schematically in Figure 1.

When considering the dilute phase pressurized system shown in Figure 1, there are 4 major components that each can contribute to the success or failure of a system. They are:

- A means to transfer the solids from a lower pressure to the higher pressure conveying system. This can be a rotary feeder or a screw conveyer. The system shown here is a rotary feeder. There are also cases where the solids are at higher pressure and a rotary feeder or valving system Is used to control the solids flow.
- A blower to provide the energy necessary to move the solids to the next processing step. The system shown here is a typical closed loop nitrogen system where the nitrogen is returned to the suction side of the blower. However, an air conveying system would be similar except the blower and return line would not be required.
- Piping to transfer the dilute phase to the next processing step. While the piping layout may seem trivial, it is very important in solids conveying as discussed later.
- The final separation of the solids from the conveying gas which is usually done by a cyclone or “dust collector”.

An engineer confronted with the design or problem solving associated with the system shown in Figure 1 struggles because relative to more traditional engineering aspects, the technology is not well defined, not easily available and not covered in the academic world. The purpose of this paper is to introduce the non-expert to terminology and calculation approaches for dilute phase pneumatic conveying. Because the calculations use proprietary correlations, only the results of the calculations are show. However, the author is available for consulting in the discussed field.

The paper presents results of calculations for important criteria such as required velocity to suspend the particles and pressure drop in the system. In addition to results of calculations, there are guidelines and values provided that will aid in the design and problem solving of these dilute phase transfer systems.

While the thrust of this paper is a solids conveying system handling air or nitrogen, the same concepts would apply to systems where the gas is a hydrocarbon. For example, in the production of polyolefins, the effluent from a reactor often flows from high pressure to a lower pressure where the motive force is provided by vaporizing liquid from a liquid phase reactor or reactor gas from a gas phase reactor.

KEY CONCEPTS

The Key Concepts consist of the following:

- Gas Rate – This is the linear velocity of the gas based on the total cross sectional area of the pipe.
- Solids to Gas Ratio-This is normally expressed in weight terms such as lbs. solids/lb. gas. However, other units are often used. The key to remember is to confirm what units are being used.
- Solids Entrance Design- The solids entrance design to the gas stream is important to get the solids moving in the direction of the gas stream. In addition, the conveying gas that fills the rotary feeder must be vented. These features are described later.
- Piping Design – The key to a successful piping design is to minimize elbows. The pressure drop in elbows is discussed later.
- Pressure Drop in piping – The presence of solids in the 2 phase flow increases the pressure drop particularly in elbows.

These areas are discussed in the following paragraphs.

GAS RATE

The gas rate must be high enough so that the solids move uniformly. While dense phase transfer is a mechanism where the solids flow rate varies with time, in dilute phase flow the solids flow rate to a receiving vessel is reasonably constant. This does not necessarily mean that the solids are traveling at the same velocity as the gas This difference between gas velocity and solids velocity is referred to as slip velocity. The most important criteria to assure that the solids flow rate to the receiving vessel is constant is gas rate.

Gas flow rate should be above the saltation velocity. Saltation velocity is defined as the minimum velocity in horizontal piping to provide the constant solids rate to the receiving vessel. This definition also implies that the solids are dispersed uniformly in the flowing gas stream. A typical calculated saltation velocity is 60-85 fps. This is for a nitrogen/air system conveying polyolefin particles with a diameter of 700-1500 microns at pressures less than 10 psig. This calculated saltation velocity is generally increased by 15-25% to insure the successful design of the conveying system. Figure 3 shows the estimated effect of particle diameter on estimated saltation velocity for the described system. Figure 4 shows the effect of solids rate on pressure drop at a fixed gas rate for the described system. Similar calculations indicate that particle density has little effect on saltation velocity. This statement seems counter intuitive. It is associated with the compensating effects that the denser single particle and lower volumetric solids flow have on the saltation velocity.

While saltation velocity is associated with horizontal flow, the comparable term for vertical flow is “choke factor”. If the gas flow rate does not provide a choke factor above the minimum, solids will tend to build up in the bottom of the vertical riser and then surge into the receiving vessel. In the cases shown here, the saltation velocity is always greater than the velocity required to avoid choking.

In an operating facility, it is often difficult to know if saltation is occurring. Figure 5 shows pressure drop in the conveying line as a function of gas rate. As shown in this figure, there is a minimum point. The minimum point in this figure is generally considered the point of **incipient** solids saltation. Incipient solids saltation is defined as the point where the lines shown in Figure 5 intersect. As shown in this figure, as the gas flow is decreased from the highest rate shown on the right of Figure 5, the pressure drop decrease until the minimum pressure drop is reached. At this point, the solids begin to settle out. As the gas rate continues to be decreased, the solids level builds in the pipe and begins to restrict the gas flow causing the pressure drop to increase. Conversely as the gas flow rate is increased from the left side of the figure, the restricted area decreases as the flow is increased and the solids slide along the bottom of the pipe. As the gas rate continues to increase, the depth of solids in the pipe decreases and the area of gas flow increases. This results in a decrease in pressure drop. At the point of incipient saltation, the solids begin to be picked up into the gas stream. However, there will still be a layer of solids sliding along the pipe. As the gas rate is increased more solid fluidization occurs and the solids are eventually uniformly disbursed in the flowing gas stream. The point where most of the solids are disbursed in the conveying gas is defined as the saltation velocity. If the gas rate is increased further, the pressure drop will continue to increase. The saltation velocities discussed in this paper are 40-50% above this point of minimum pressure drop (incipient saltation velocity).

While it might seem appropriate to be conservative and design the gas rate to be well above the saltation velocity. Besides wasting funds in overdesign of the system, there is a risk of the higher velocity causing the temperature of the solid particles to increase past the melting point. At elbows the particles approach a complete stop and the kinetic flowing energy is converted to heat. If this heat is not quickly dispersed throughout the particle, some of the particle surface may be in the melt phase. These particles with melted surface may create agglomerates or other types of irregular particles.

The particles that slow down at the elbows must then be reaccelerated to the solid's velocity. This reacceleration creates additional pressure drop. So any excess gas rate causes both an increase in pressure drop associated with reacceleration at elbows and pressure drop associated with the higher gas rates. The increased pressure drop in the piping and reacceleration at elbows will lead to higher pressure at the entrance from the rotary feeder. This increased pressure will increase the amount of conveying gas trapped in the empty pocket of the rotary feeder. Thus, enhanced facilities must be provided to allowing venting this gas.

The subject of minimum saltation velocity is so complex that experts in the field often indicate that experimental work is the only way to accurately predict it. The correlations used in this paper seem to give reasonable results and are in general agreement with work published by F. Rizk. The Rizk relationship is shown below:

Fr^a = Ø*10^b

V=Fr*(gD)^.5

a = 1.1*d+2.5

b = 1.44*d+1.96

Where:

Fr = The calculated Froude number

Ø = The solids loading in wt. solids/wt. gas

g = gravitational constant 9.81 m/sec2

D = Pipe diameter in meters

d = particle diameter in mm

V = Saltation velocity in m/sec

SOLIDS TO GAS RATIO

The amount of solids being conveyed with a fixed amount of gas can impact the pressure drop. Figure 4 shows the impact of the solids rate on the estimated pressure drop. Table 1 shows this along with estimated saltation velocity for the various solids rates. Table 1 also shows the ratio of two phase (gas and solids) pressure drop to gas only pressure drop. The

Piping Configuration	Ratio of Pressure Drop for Two Phases/ Gas only
Horizontal Pipe	1.5 to 1.9
Vertical Pipe	1.9 to 2.5
Elbows or Bends	5.7 to 8.8

This summary shows the importance of elbows or bends in a piping layout. Table 2 shows the basis for these calculations.

SOLIDS ENTRANCE DESIGN

The optimum design of solids entrance into the flowing gas stream is shown schematically in Figure 2. In a pressurized system the rotary feeder pockets will fill with gas as the solids are discharged and flow into the conveying gas stream. This gas will be vented as rotary feeder rotate and solids flow into the pocket. If the solids are small enough, this vented gas will fluidize them and they will be held up in the standpipe. Eventually they will dump into the rotary feeder. This unsteady state creates fluctuating solids flow which can cause the dilute phase system to plug. The vessel design shown in Figure 2 should be large enough to allow the gas to disengage from the solids and vent to a proper location. Figure 2 also shows a sloped solids entrance that will assist in starting the solids flow into the flowing gas stream in the same direction as the flowing gas.

PIPING DESIGN

The piping design for dilute phase solids conveying is very important. The two most important guidelines are;

- Piping runs must be either horizontal or vertical. Inclined piping runs have been shown to have 30% higher pressure drop than horizontal piping. In the case of the conveying gas being a volatile hydrocarbon from a pressure reduction step, there will be a high rate of gas and the increased pressure drop will not be a concern. In this case it may be acceptable to use inclined piping.
- Elbows should be minimized. As indicated earlier elbows cause particles to stop flowing until they are reaccelerated. The motive force for this reacceleration must come from the conveying gas which creates an increased pressure drop.

In addition to these guidelines, considerations for the design of a sampling system are often of interest. In order to obtain representative samples of the solid being conveyed, criteria are necessary. Two alternate sampling design are possible as follows:

1. In the first alternative the sample is taken from a well dispersed stream of gas and solids. The solids must have obtained uniform dispersion in the flowing gas stream. This can often require 30-50 diameters of length following the injection of solids. The second criteria is somewhat more difficult to achieve. The sample line itself should extend into the center of the conveying line and the inlet should point directly into the flowing stream. In addition, the velocity in the sampling line going to the sample cyclone (or other solids-gas separator) should be designed for the same velocity as the actual gas velocity in the conveying line.
2. A second approach is to install a sampling device on the outside curve of an elbow. At this point the solids will be sliding along the elbow. If particle size distribution is important, this approach may give false results.

LABORATORY EVALUATIONS AND SCALEUP

The correlations to determine saltation velocity are dependent on factors such as particle shape, particle density, and small particle agglomeration that have a wide degree of error. As indicated earlier this often forces experts to use small scale testing. Using transparent pipe, it is relatively easy to spot the point of uniform solids flow. If it is not possible to obtain a visualize observation of the solids flow, The technique described above and shown in Figure 5 can be utilized. As indicated earlier, to obtain the desired flow regime of fully suspended, the velocity should be increased by 50% or more. With velocities developed in a laboratory experiment the scaleup criteria for a commercial design must be developed. Based on the proprietary correlations discussed earlier, the scaleup is probably best done keeping the laboratory velocity constant in the commercial plant. This will likely result in a lower pressure drop in the commercial plant than observed in the laboratory for the identical piping configuration.

SUMMARY

While the science involved in dilute phase transfer is evolving, this paper provides some exposure for the non-expert to this critical area. I would be interested in running my model on any set of actual data on a pro bono basis. This will provide additional credibility for the model as well as some potential improvement areas for you.

TABLE 1: IMPACT OF SOLIDS RATE ON PRESSURE DROP

ITEM					
Solids Rate pph	12000	14000	16000	18000	20000
Saltation Velocity fps	43	47	51	56	60
Gas Velocity fps	97	97	97	97	97
Solids Velocity fps	88	88	88	88	88
Total Calculated Pressure Drop psi	2.9	3.2	3.5	3.8	4.1
RATIO OF TOTAL PRESSURE DROP TO GAS ONLY PRESSURE DROP IN					
Elbows	5.7	6.5	7.3	8.1	8.8
Horizontal Pipe	1.5	1.6	1.7	1.8	1.9
Vertical Pipe	1.9	2.0	2.2	2.3	2.5

TABLE 2: BASIS FOR CALCULATIONS

Operating Conditions unless variable is used as independent variable

- Gas Flow - 9000 pph
- Solids Flow – 20000pph
- Particle Size – Spherical with 700 microns average diameter
- Particle Density – 51.7 lbs/ft3
- Particle Size Distribution – The assumed particle size distribution is such that d15.9 /d50 equals 0.42. Where d15.9 = The size particle that 15.9 % of particles are smaller than and d50 represents the medium particle size.
- Particle Specific Gravity - 0.828
- Discharge Pressure - 0.2 psig

Piping System Basis

- Horizontal Length – 400 feet
- Vertical Rise – 100 feet
- Number of elbows – 10
- Pipe ID – 8 inch

FIGURE 1
SCHEMATIC FLOW SHEET
PRESSURIZED CONVEYING SYSTEM

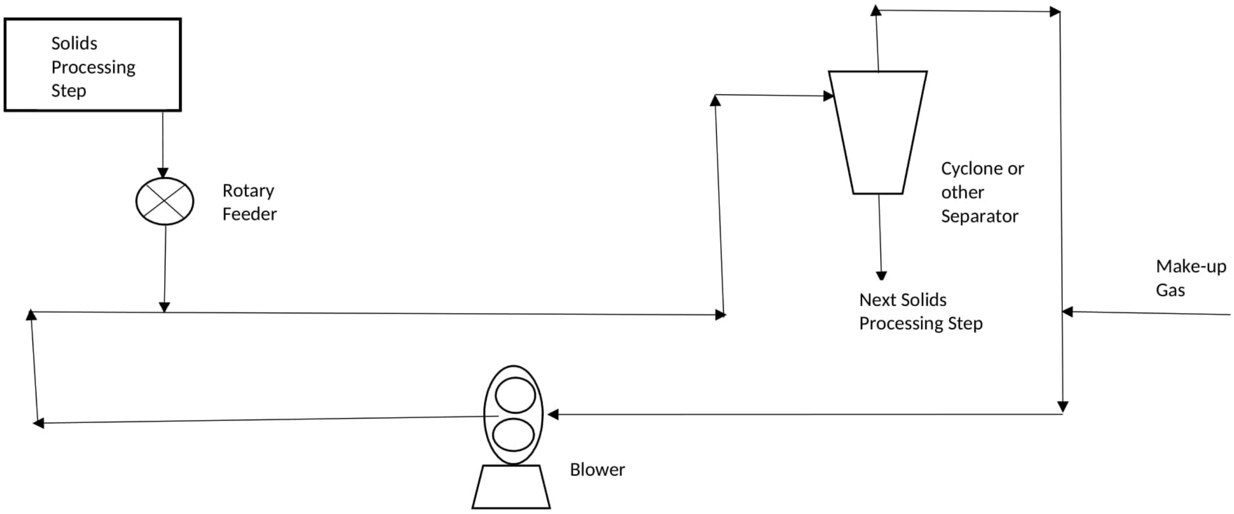


FIGURE 2
SCHEMATIC DESIGN
OF
CONVEYING LINE FEEDER

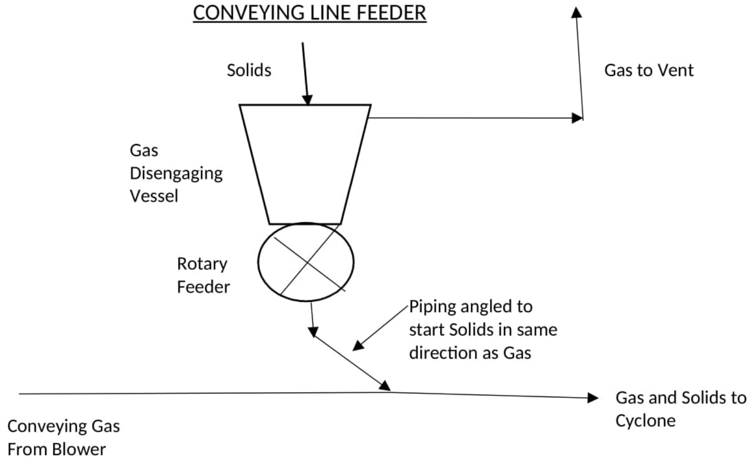
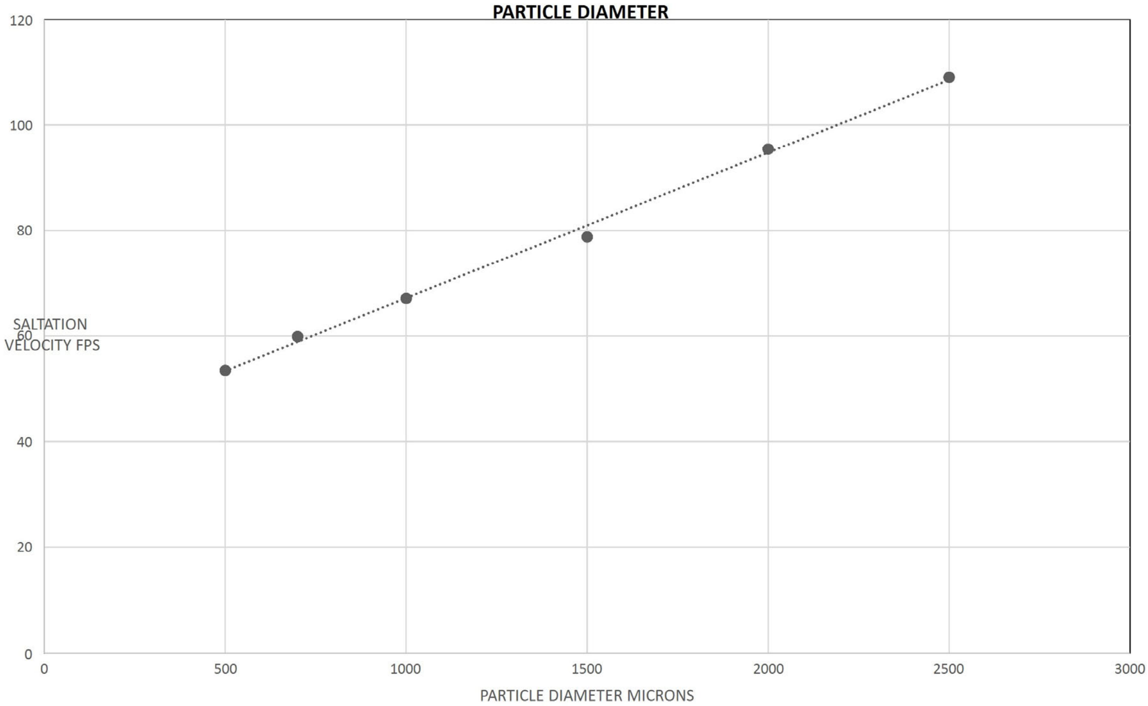


FIGURE 3
SALTATION VELOCITY
VS
PARTICLE DIAMETER



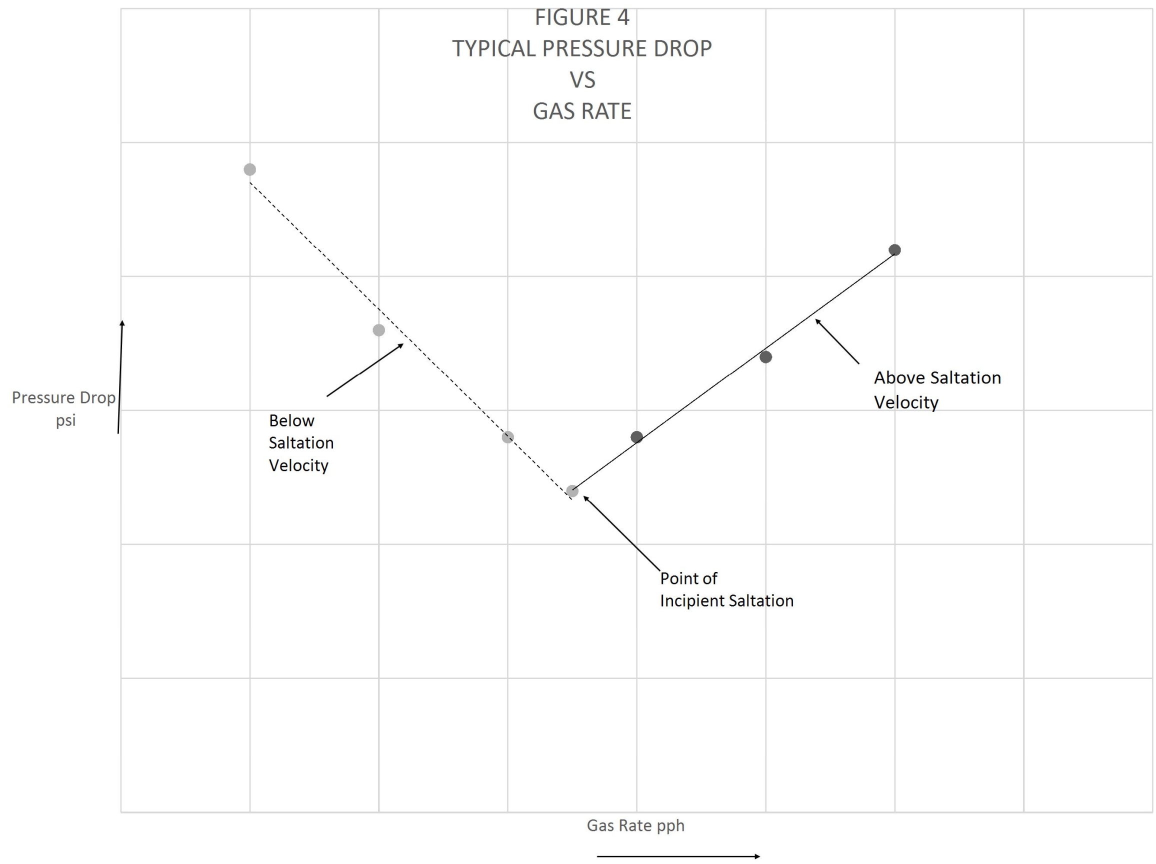


FIGURE 4
TYPICAL PRESSURE DROP
VS
GAS RATE
Pressure Drop
psi
Below
Saltation
Velocity
Above Saltation
Velocity
Point of
Incipient Saltation
Gas Rate pph

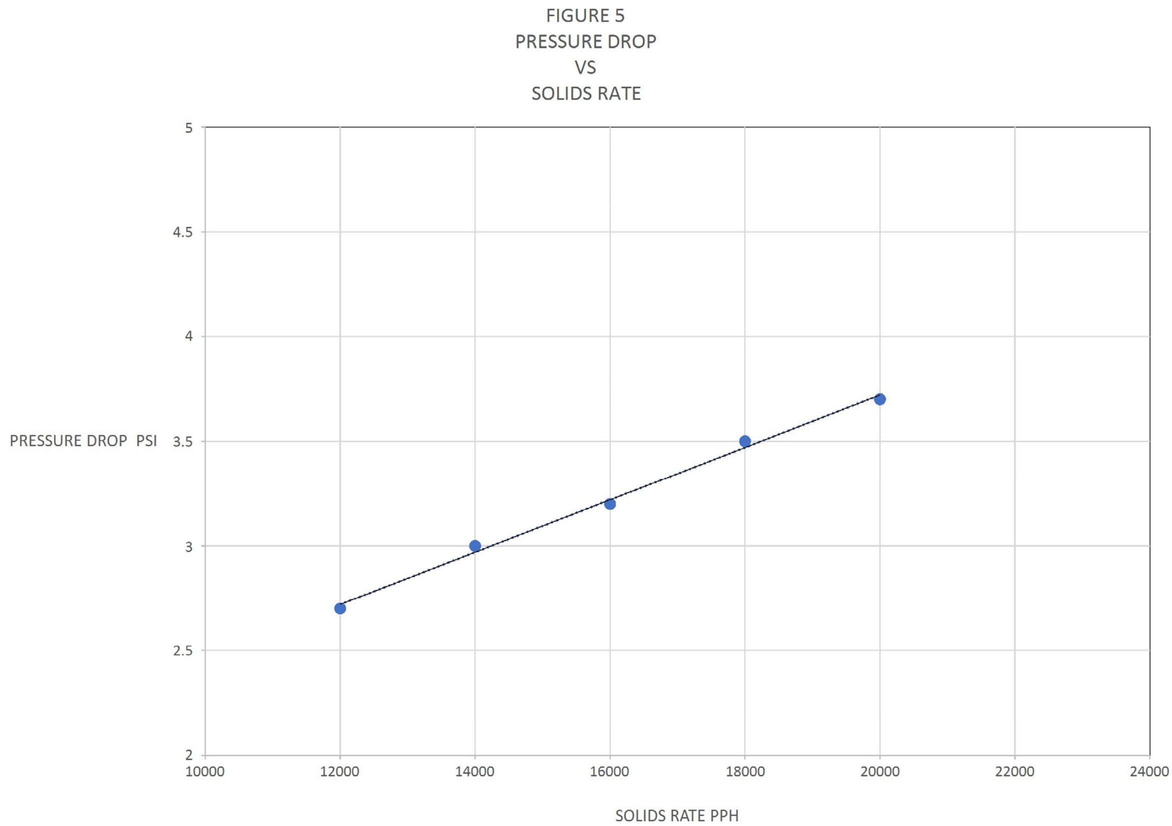


FIGURE 5
PRESSURE DROP
VS
SOLIDS RATE
PRESSURE DROP PSI
5
4.5
4
3.5
3
2.5
2
10000
12000
14000
16000
18000
20000
22000
24000
SOLIDS RATE PPH

ABOUT THE AUTHOR

Joe Bonem is a highly experienced process/ chemical engineer in areas such as oil production using fracturing techniques, oil production from the Canadian oil sands (including cleanup of tailing ponds), refining of oil and production of petrochemicals. Over the last few years, he has consulted with several Fortune 500 companies as well as some smaller companies. The areas of consulting have included process development and scaleup, process design, plant problem solving and problem solving training. He has conducted training courses in the USA and Middle East (United Arab Emirates and Bahrain) as well as worked with companies in Japan, England, France, Italy and Belgium. His specialties include: process engineering and chemical engineering, plant problem solving and training; process scaleup and development; and manufacturing and technology of olefin based polymers.

Considerations for Fabrication of Mass Transfer Products

Neil Almeida, Consulting Tool Engineer

This article points out some of the technical points in the manufacturing/fabrication of Mass Transfer products.

These do not affect the fabrication in a really significant way, but if noticed, could be very useful and make life easy from a manufacturing point of view.

In this article will use a "Tray" as an example of Mass Transfer product to illustrate the points.

TRAYS

1A- SIEVE TRAY DECK – SURFACE TENSION OF FLUID FLOW

Let us consider the Sieve tray deck (The following information equally applies to all kinds of tray deck openings.) ,

Hole opening:

There are two methods of making a hole that are predominantly used in the industry currently, they are:

- Laser cut
- Turret punch.

Both options are used widely nowadays due to their speed & accuracy.

Fig.1 shows an example of holes made by both the methods. The Top half shows a punched hole, made on a CNC Turret press & the bottom half shows a hole cut by a Laser machine. Standard Pickling has been done on both the samples. Some water was poured on the samples to simulate fluid flow to observe the difference in fluid flow on the surface of the samples. The thickness of the tray is inconsequential because we are considering the fluid flow pattern on the top surface of the tray/deck.

In a Turret punched hole the slight radius on the surface of the tray at the edge of the hole, reduces surface tension so as to allow the free flow of the fluid without pooling at the edge of the hole. This is clearly visible in the picture as the area around the hole is largely dry.

The sharp edge of a laser cut hole does not allow the free flow of fluid through the hole. This is clearly indicated because the area around the edges of the hole is quite wet.

Fig.1 Comparison of surface tension of on fluid flow in Punched & Laser cut hole.

Let us consider a fluid with more viscosity & impurity is flowing through the holes.

As impurities rise, the sharp edges of a laser cut hole offer further resistance to fluid flow which leads to fluid build up around the edges. This causes the surface tension to drop around the edges that leads to fluid build up. This fluid build up can lead to reduction in hole diameter as impurities gather around the edges of the hole. Because of the build up the fluid level rises and never actually flow on the surface of the tray.

This kind of fluid build up may be become really noticeable and affect performance in the long run. This loss of performance is generally not taken into account during process calculation.

1B FIXED VALVE – RISE OF FIXED VALVE

Fixed valves can be fabricated with a CNC Turret punch, a Normal OBC bigger neck press/ or a NC Press. CNC Turret punch is used if the production speed is important.

Fixed Valves should be produced using a

Normal press if the Valve Rise is an important consideration in the Mass Transfer process.

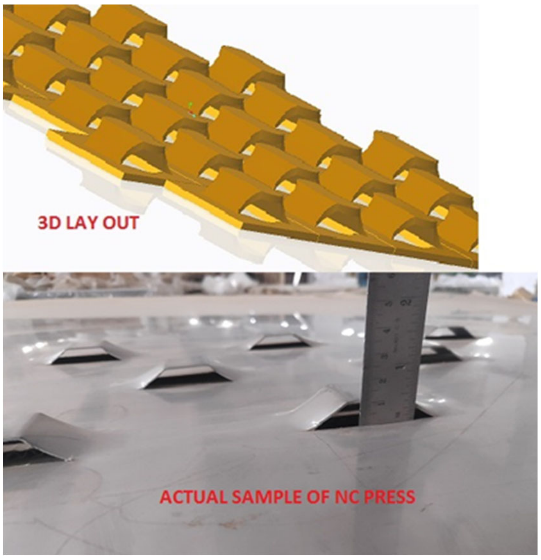


Fig.2 Typ. View of Valve Rise in 3D & actual sample of Press for Fix Valve

Fixed valves are not used in huge quantities generally. So, NC presses are used to balance the production rates and the valve rise. NC press can produce valves at speeds comparable to CNC Turret punch and produce valves with bigger rise than a Turret punch. The rise produced by a NC press is comparable to that of Normal OBC press.

The top half of Fig.2 shows a 3D image of the typical heights desired in fixed valves. The bottom half shows the actual picture of a fixed valve punched on a NC Press.

The maximum height of a fixed valve produced in a CNC Turret was 8.0mm due to the limitations of standard Turret Gap between top & bottom turrets.

NC presses can easily produce a life of 12-13mm. It could easily go up to 15-16 mm rise. We have made fixed valves with a 4.0mm thickness in Hastelloy C276 on NC Press, which cannot be done on a standard CNC Turret punch due to limitations of a CNC turret handling thicker material and loner runs.

1C-ROUND FLOATING VALVE

Round floating valves are manufactured in one of two ways: with a single progressive die or multi stage with multiple dies.

A Progressive Die can produce the valves at a significantly higher rate allowing the producer the ability to supply parts in near real time JIT, but the product yield is about 50% of the raw material thus increasing the product cost. A

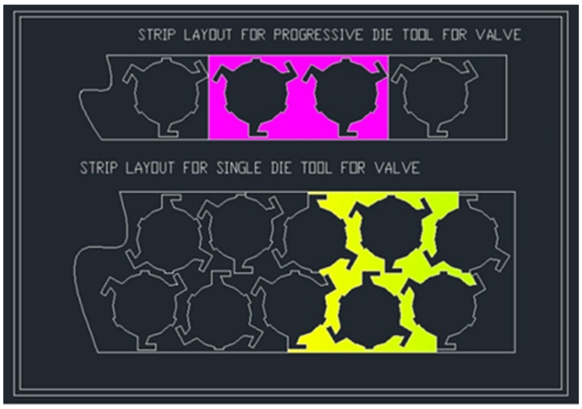


Fig.3 Strip layout for a progressive die & multistage single tool dies for manufacturing Round floating valve.

progressive die can cost about 1.5 to 2 times the combined cost of all the multistage tools, increasing the product cost significantly.

Naturally the production time for multistage tools is higher than with a progressive die, but the product yield is about 80% of the raw material which results in significant cost savings This is accomplished by tight nesting of product on a sheet. Fig 3 shows the difference in product yield for a progressive die and multistage multi tool production.

Unlike random packing valves are not used in large quantities in Mass transfer, so it might not be a good idea to choose a progressive die in this situation. It is advisable to choose the manufacturing process based on demand.

1D-STYLES OF TRAY DECK ASSEMBLY

It is important to note that the choice of tray deck design (joining different sections) is not particularly dependent on process but is heavily influenced by manufacturability and ease of installation in a column.

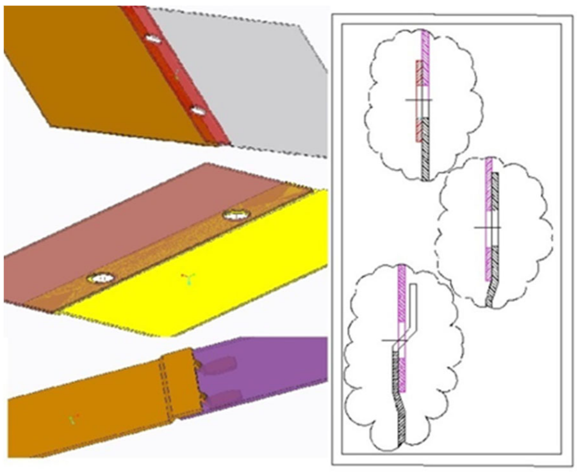


Fig.4 Different styles of Tray Assembly

The three different styles of assembling the two sections of the tray deck are illustrated in Fig4., where process design remains intact for all and change has been made mechanically considering ease for manufacturing as well as installation.

The first style being the most awkward and difficult style, where two sections are joined inside a column by bolting them together. The second style is an improvement on the first style where a jog at each end of the section allows for the overlap of adjacent sections making the bolting of the sections slightly easier than the first style and is commonly used for assembly of sections.

The third style, the latest development in joining tray deck sections which is considered most installation friendly design and which is an NBC (non-bolting construction) design that allows the installer to hold a section by its extended teeth and sliding the teeth through the slots of the next section.

We know, installation in Column is not an easy task and by choosing Third option, one can save at least 50% of deck assembly time due to NBC design.

1E-POSIBILITY OF EXTRA VALVE

The top half of Fig. 5, shows a Sieve Tray in assembly position and the bottom half shows floating valves in assembly position.

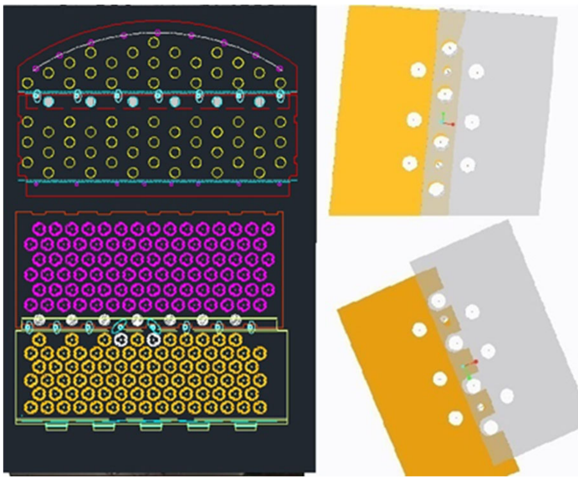


Fig.5 Typ. Proposed Place for extra valves

From a process point of view, the ideal situation is to have perforations all over the exposed (diametrical) area in a column. But mechanical limitations like joined tray decks sections reduce the working area. This lost working area cannot be considered for process calculations in valve placement design.

If we can utilise a portion in this lost working area, it will result in better performance of the mass transfer process.

On a sieve tray section if we can extend the perforation area till the area specified for clamping and fitting (illustrated in the pictures above) and can match holes on adjacent sections, when they are clamped together, additional room can be created for an extra sieve valve, opening up more process working area enabling better performance of the tray.

In a floating valve tray, we can make provision for extra valves around the area where two sections are joined together by cutting the overlapped section of the bottom section in a extra large 'U' form to allow for the legs of the valve to float up and down as illustrated by the bottom half of Fig5.

This kind of arrangement is not easy to make in a fixed valve tray unlike a floating valve tray because more material is needed at the corner of a tray to prevent the tray from being torn away from base deck. A bigger slot needs to be made on the bottom tray where it is joined to the top tray. This slot needs to be longer than the legs of the valve to allow easy flow of fluid. This kind of arrangement is only possible in trays that have some room to work around and not fit too tight in the column.

These are some items to consider to in designing better performing mass transfer products.

1F- PROVISION FOR STAND BY VALVE

A good option for the Trays to consider in some cases would be to provide for extra valves which are blocked off &are not used under normal operations.

When the original set of valves clogged with the impurities in the fluid, the blocked off valves can be opened for continued operations of the column albeit at a reduced capacity to allow for replacement to be ordered.

This can shortened downtime of column.

Special Thanks: Mr.Karl Kolmetz, Mr. Kazuo Watari, Mr. Ananth Halvi

ABOUT THE AUTHOR

Neil J. Almeida ha over 18 years of experience as consulting Tool Design Engineer in Mass Transfer and have privilege to develop new products for clients.

Test Your Process Knowledge with These 20 Questions

Norm Lieberman, Process Improvement Engineering

You can gauge the extent of your knowledge in process operations and engineering. The test assumes that you have completed high school science. The game of “Twenty Questions” was popular when I was a teenager in the 1950’s. These questions are a mixture of operational and process engineering problems. Score yourself based on the attached answer page. If you get 100% correct, I will send you a free book, “Process Design for Reliable Operations, 2nd Edition”.

1. Increasing the temperature of 1,000 ft3 of air from 60°F to 600°F, increases the air volume to about:
    - A. 2,000 ft3
    - B. 10,000 ft3
    - C. 20,000 ft3
2. The purpose of an impeller in a centrifugal pump (in non-viscous flow) is to _________?
3. A steam vacuum ejector using 150# motive steam, has an internal steam nozzle where the steam velocity increases from 120 ft/sec to 12,000 ft/sec. Where does the energy come from to accelerate the steam?
4. The purpose of a valve tray in a fractionator, is to fractionate between the lighter and heavier components. True or false?
5. A centrifugal pump mechanical seal flush pressure should be:
    - A. Higher or close to the discharge pressure.
    - B. Just a little lower than the discharge pressure.
    - C. Just a little higher than the suction pressure.
    - D. Just a little lower than the suction pressure.
6. The energy of an old-style reciprocating steam driven engine, using 150# pressure steam, that exhausts to the atmosphere, comes from what property of the steam?
7. How does lowering a distillation tower pressure affect the relative volatility between propane and butane?
    - A. Bigger
    - B. Smaller
    - C. No effect
8. Which type of tray has a better potential to develop efficient tray fractionation efficiency.
    - A. Sieve
    - B. Valve
    - C. Bubble Cap
    - D. Grid
9. When we change from a 4” pipe, to a 2” pipe, at constant volumetric flow, the pressure drop would increase from 2 psi to:
    - A. 4 psi
    - B. 8 psi
    - C. 16 psi
    - D. 32 psi
    - E. 64 psi
10. Why does water passing through a cooling tower cool, to less than the ambient temperature?
11. What happens to the density of water as it cools? Increases or decreases?
12. To increase the Net Positive Suction Head available to a pump, pumping liquid from a vapor-liquid separator, at <u>equilibrium</u>, should we increase or decrease the vessel’s pressure?
13. Operators sometimes spray water on a hot pump suction line, to suppress cavitation. Does this actually help the pump’s performance?
14. Steam turbines running with 400 psig steam, exhausting to the atmosphere, are using what property of the steam to drive the turbine:
    - A. Temperature
    - B. Pressure
    - C. Latent Heat
15. When unloading a reciprocating compressor, the most efficient way to reduce load is:
    - A. Open spill-back.
    - B. Open head-end unloader bottle.
    - C. Throttle the discharge.
16. When we increase the thickness of an orifice plate, at constant flowing velocity, what happens to the pressure drop?
    - A. Higher
    - B. Lower
    - C. Unchanged

17. Can the pressure inside of a fired heater radiant section, exceed atmospheric pressure?
18. How does wind affect draft in a fired heater:
    A. Bigger
    B. Smaller
    C. No effect
19. Why do the non-condensables, vented from a steam heater, typically burn when ignited?
20. What is the best way to control reflux and reboiler duty in a light-ends distillation tower or an aromatic fractionator?

19. ANSWERS

1. 2,000 (remember Rankine or Kelvin)
2. Accelerate the flow – not to increase the pressure.
3. Mostly from the temperature of the steam. Almost none comes from the steam's pressure.
4. False. The purpose of the tray is to mix the vapor flowing up from the reboiler, with the liquid flowing down from the condenser.
5. C – Just a little bit higher than the pump suction pressure.
6. It's the latent heat – not its pressure.
7. Bigger
8. Bubble cap
9. 32 psi
10. Evaporation of the water.
11. Increases to 40°F (4°C) and then gets less dense. A unique property of water.
12. At equilibrium, it doesn't matter.
13. Yes
14. Temperature + Latent Heat (i.e., enthalpy).
15. Open the head-end (or crank-end) unloader bottle.
16. Lower – by a lot.
17. Yes
18. Bigger
19. It's hydrogen from CO_2 corrosion. The hydrogen ions cause hydrogen assisted Stress Corrosion Cracking.
20. Closed-loop Spectroscopic analyzer control. Gas Chrome lag time is excessive. Check with Tec5USA.

Thanks for taking our test. We hope that it will help in your work. If you would like to have an explanation of any of our answers, please email: norm@lieberman-eng.com

Or, you can consult our website and select one of the eleven books we have authored pertaining to Process Engineering (www.lieberman-eng.com).

ABOUT THE AUTHOR

Norm Lieberman has a degree in Chemical Engineering (Cooper Union, 1964). His principal activity is field troubleshooting refinery problems. Over 21,000 operators and engineers have attended his 950 seminars since 1983, that explain Troubleshooting Techniques. Lieberman has authored 11 books on this subject. He resides in New Orleans, where he was formerly the Plant Manager of the Good Hope Refinery in Norco.

Guidelines for Process Plants Alarm Grouping & Alarm Prioritization

Praveen Nagenderan C

1.0 IMPORTANCE OF ALARM SYSTEM

Alarm systems are increasingly important in the safe management of plant and machinery. Alarm systems forms an essential part of the operator interface which provides vital support to the operator by warning them of situation that need their attention. Alarm systems thus have an important role in preventing, controlling and mitigating the effects of abnormal situations. The effects can be very serious if these alarm systems does not work well. Alarm systems are a very important way of automatically monitoring the plant condition and attracting the attention of the process plant operator to significant changes that require assessment or action. They help the operator to maintain the plant within a safe operating envelope. A good alarm system helps the operator to correct potentially dangerous situations before the Emergency Shutdown (ESD) system is forced to intervene. This improves plant availability, helps to recognize and act to avoid hazardous situations, helps to identify deviations from desired operating conditions that could lead to financial loss and helps to understand complex process conditions. Alarms should be an important diagnostic tool and are one of several sources that an operator uses during an upset. Alarms are signals which are annunciated to the operator, typically by an audible sound, some form of visual indication, usually flashing, and by the presentation of a message or some other identifier. An alarm will indicate a problem requiring operator attention, and is generally initiated by a process measurement passing a defined alarm setting as it approaches an undesirable or potentially unsafe value. Alarm Management helps to identify process problems like valves / equipment & instruments malfunction and controller tuning problems, reduces unplanned down time of plant, reduces production losses, prevents incidents, improved productivity – both equipment's and personnel's. Alarm systems should be designed to meet user needs and operate within the operator's capabilities. This means that the information alarm systems present should be relevant to the operator's role at the time, indicate clearly what response is required, be presented at a rate that the operator can deal with, be easy to understand.

2.0 ALARM GROUPING

Alarm grouping is one of the important theories in the plant alarm management system where the alarms configured are to be grouped into different categories. All alarms which are configured in the system should be specified with a group name. The alarm grouping activity later supports alarm rationalization activity as well significantly for analysis.

2.1 GENERAL GUIDELINES

- Instrument tags related to process which are configured with alarms are to be provided with the group name with respect to the respective defined process areas. These alarms are commonly called as "PROCESS ALARM".
- Safety systems related alarms shall be provided with the group name of "SAFETY".
- Equipment start/stop related alarms shall be provided with a group name of "EQUIPMENT".
- Shut down valves, Blow down valves and Control valves related alarms shall be provided with a group name of "VALVES".
- DCS or PLC system related and Communications related alarms shall be provided with group name of "SYSTEM".
- Electrical switch gear related alarms shall be provided with group name of "ELECTRICAL".

3.0 ALARM PRIORITY

Alarm priority is used to aid the operator determining the order in which to respond to alarms. Effective prioritization typically results in higher priorities chosen less frequently than lower priorities. Most of the alarms should be assigned to the lowest alarm priority (least important) and the fewest to the highest alarm priority (most important) with the consistent transition between the two. The resulting priorities should have alignment with the consequence and allowable response time such that the lowest priority alarms have the least severe consequence and longest allowable response time and the highest priority alarms have the most severe consequence (Example: Fire and Gas system alarms) and the shortest allowable response time.

3.1 GENERAL GUIDELINES

All alarms configured in the system will be prioritized based on the class names.

- Facility shall generally use four classes for process related alarms and one class for system related alarms.
- Alarms which are grouped under SYSTEM should be classified as "SYSTEM".
- Other alarms which are grouped under different names as per the alarm groups guidelines as specified above are to be classified as "CRITICAL", "HIGH", "MEDIUM" and "LOW".
- Priority distribution of the process related alarms class shall be: CRITICAL: 5%, HIGH: 15%, MEDIUM: 30% and LOW: 50%. (Determination of priority percentage is a case dependent activity with respect to the respective process industry/plant. General percentage which works and practically possible to implement were suggested here).
- Alarms which are critical to process safety of the protection of human life or personnel safety protection by default are to be considered for "CRITICAL" priority class.
- Alarms for commercial loss or product quality shall be considered for "CRITICAL" priority class.
- Instrument tags which are configured with alarms having direct relation to the total plant trip or a section trip shall be considered for "CRITICAL" priority class.
- Instrument tags which are configured with alarms which affects the section of the plant but provides relative response time to the operator or which doesn't trigger total plant shut down immediately shall be considered for "HIGH" priority class.
- Equipment stop related alarms shall be considered for "HIGH" class based on the criticality of the equipment with respect to the process conditions whereas equipment start related alarms shall be considered for "LOW" class based on the criticality of the equipment with respect to the process conditions.
- Alarm priority class for Shutdown valves or solenoid valve status alarms to be decided based on the criticality of the process conditions. Shutdown vales or solenoid valve status feedback CLOSE/OPEN alarms which are having direct relation to the total plant trip or respective valve closure / open status which gives very less response time for operator to take action can be considered for "CRITICAL".
- Instrument tags which are configured with alarms which performs remote operation i.e. operation performed by operator to be considered for "LOW" class.
- Characteristically all bad PV (Process Value) alarm shall be configured with low priority except for the points/tags with high priority alarm where bad PV shall also be assigned high priority.
- As per HAZOP study report, Alarms listed as safeguards measures or recommendation shall be assigned with priority as defined by the severity consequence matrix.

4.0 REFERENCES

ANSI/ISA-18.2.2016 Management of Alarm Systems for the process industries

EEMUA PUBLICATIOn No.191 Edition 2 Alarm Systems A Guide to Design, Management and Procurement

ABOUT THE AUTHOR

Praveen Nagenderan C is a Chemical Engineer with experience in the field of Oil & Gas production & processing facilities and Refinery process units. Professional experience covers in Production operations, Facility surveillance, Technical safety, Technical Services - Process, and Projects. Praveen has worked with major Oil & Gas companies in India namely Nayara Energy formerly known as Essar Oil Limited and Cairn Oil & Gas.

Adding Value to Natural Gas | Gas to Liquids and Petrochemicals Technologies

Dr. Marcio Wagner da Silva

INTRODUCTION AND CONTEXT

Despite the current scenario where is observed a surplus of crude oil followed by low prices, the downstream industry lives constantly with uncertainties related to guarantee of access and supply of crude oil in the quantity and quality required, problems related to the geopolitics or even about the exhaustion of existent recoverable reserves are driving forces to the development of alternatives technologies the crude oil. This topic is especially attractive to countries that present a lack of a significant amount of crude oil reserves and great dependence by crude oil derivatives as Japan and China.

In this case, the look for alternatives to crude oil aiming to sustain the energy demand necessary to sustain economic development become a strategic issue, as mentioned earlier, mainly considering the volatility of the crude oil prices as well as the geopolitical scenario. Despite the higher cost in comparison with the traditional crude oil refining, the necessity of the production of high-quality fuels can make the nonconventional alternatives attractive. An example was the coal liquids hydrogenation during the Second World War to produce liquid fuels on a large scale due to the scarcity of crude oil by some countries. Crude oil production players with great reserves of natural gas can find in the Gas to Liquid technologies an attractive way to ensure higher added value to his natural resources.

GAS TO LIQUIDS ALTERNATIVE

One of the most promising and well-developed technologies currently is the conversion of syngas (CO + H2) in longer-chain hydrocarbons such as gasoline and other liquid fuel products, known as Gas to Liquids Technologies (GTL). The liquid hydrocarbons production can be carried out by direct syngas conversion, in Fischer-Tropsch synthesis reactions or through methanol production as intermediate product (Methanol to Olefins technologies).

Fischer-Tropsch is a chemical process through is possible the liquid hydrocarbon production according to the following chemical reactions:

Paraffin Production:

$$n\,CO + (2n+1)H_2 = C_nH_{2n+2} + nH_2O$$

Olefin Production:

$$n\,CO + 2nH_2 = C_nH2_n + nH_2O$$

These reactions are strongly exothermic and the CO/H2 ratio in the syngas is a key parameter to define the hydrocarbon chain extension that will be produced.

The reactions occur normally under temperatures that vary from 200 to 350 oC and operating pressures in the range of 15 to 30 bar. The catalyst commonly applied to these reactions is based on cobalt or iron as active metals deposited upon alumina as carrier.

Figure 1 presents a block diagram for a typical process plant dedicated to producing liquid hydrocarbons from Fischer-Tropsch synthesis.

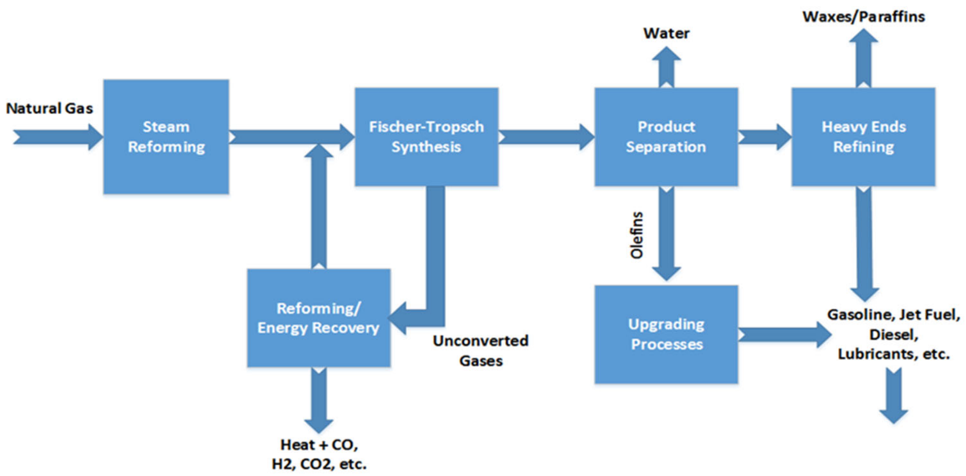


Figure 1 – Block Diagram to a Typical Fischer-Tropsch GTL Process Plant

Showed process in Figure 1 is based in the syngas gas generation from steam reforming of natural gas, this is the most common route, however, there are process variations applying syngas production through coal, biomass or petroleum coke gasification route.

The process starts with syngas generation and, as aforementioned, the produced hydrocarbon chain extension is controlled in the Fischer-Tropsch synthesis step through the CO/H2 ratio in the syngas fed to the FT reactors (beyond temperature and reaction pressure), following the produced hydrocarbons are separated and sent to refining steps as isomerisation, hydrotreating, hydrocracking, catalytic reforming, etc. According to application of the produced derivative (Gasoline, Diesel, Lubricant, etc.).

Some side reactions can occur during the hydrocarbons production process, leading to coke deposition on the catalyst, causing his deactivation according to following chemical reactions:

$$2\,CO = C + CO_2$$
(Boudouard Reaction)

$$CO + H_2 = C + H_2O$$
(CO Reduction)

The type of reactor applied in the FT synthesis step have strong influence on the yield and quality of the obtained products, the campaign time of the process units also depends on the type of reactor. Fixed bed reactors are widely employed to FT synthesis, however, show a reduced campaign time due to the low resistance to catalyst deactivation phenomenon. Modern process units apply fluidized bed or slurry phase reactors that present a higher resistance to coke depositionon the catalyst and better heat distribution, leading to higher campaign periods.

Most recently is observed a reduction trend in the demand by transportation fuels and some refiners are looking for change his production focus from transportation fuels to petrochemicals. The gas to liquids can be applied in synergy with conventional refining processes to improve the yield of petrochemicals in the refining hardware through the production of high quality naphtha that can be applied to FCC or steam cracking units to produce light olefins, ensuring higher added value to the processed crudes and gas as well as participation in a growing market.

Another attractive alternative and synergy opportunities to refiners is the production of ammonia that are the base of any fertilizer. Despite the flat demand over the last years, is expected a growing market in the next years due to the increasingly demand by food at global level. As presented in Figure 2, is also expected a growing demand of Methanol in the next years, this intermediate can be used to produce high demand products like formaldehyde that is applied to produce plastics and coatings, allowing great added value to the crude oil and natural producing chain.

AVAILABLE TECHNOLOGIES

The several geopolitics crises over the history motivated the development of new technologies and the improvement of the Fischer-Tropsch original process. There is a wide array of process technologies developed aiming to liquid hydrocarbons production from syngas, among the principal available technologies we can quote the processes SYNTHOL™ and SPD™ developed by Sasol Company, the GASEL™ process by Axens

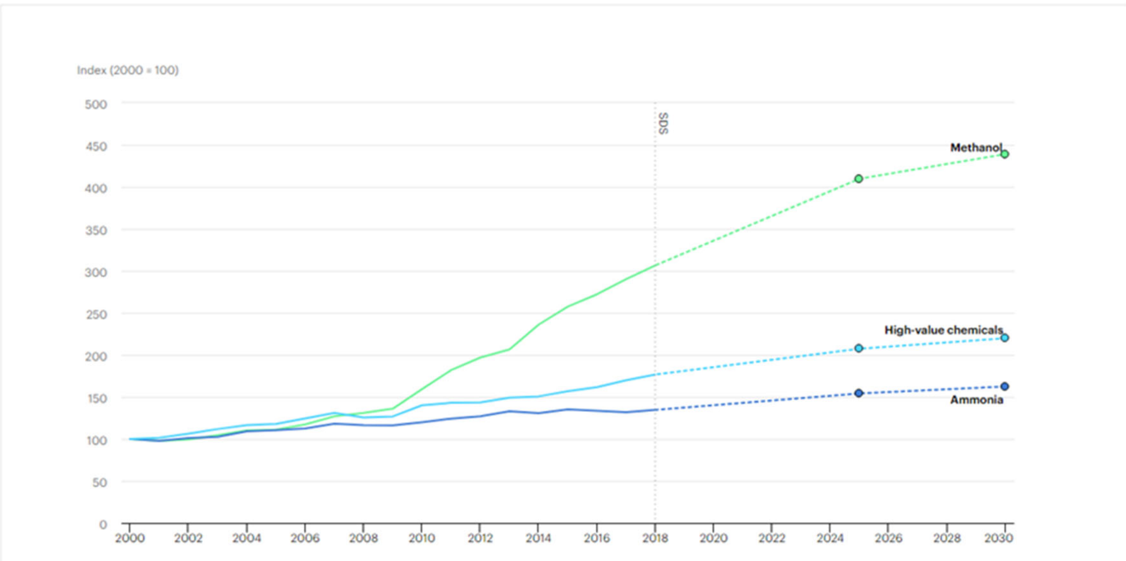


Figure 2 – Primary Chemicals Production Forecast (IEA, 2020)

Company, the E-Gas™ technology by ConocoPhillips company, the SCGP™ by Shell, the TIGAS™ developed by Haldor Topsoe Company, among others.

Figure 3 presents a basic process flow diagram for the GASEL™ technology developed by Axens Company which apply a slurry phase reactor.

As cited earlier, one of the main advantages of GTL technologies is the possibility of application several raw materials to produce syngas, which ensures great flexibility. In regions with large coal availability, the gasification technologies has strategic character given that the great restriction of this fuel use in the energetic matrix due the high environmental impact, in these cases the coal conversion into syngas and posteriorly in liquid hydrocarbons is very economically attractive, another alternative is to apply renewable raw material (biomass) to produce syngas.

On the other hand, in regions with great availability and easy access to large natural gas reserves, the syngas production through natural gas reforming steam still is shown as the most economical route to produce this raw material in industrial scale.

An alternative route to produce liquid hydrocarbons from syngas is the non-catalytic conversion of the natural gas to methanol followed by the polymerization to produce alkenes. Methanol is produced from natural gas according to the following chemical reactions:

$$CH_4 + H_2O = CO + 3H_2$$
(Steam Reforming)

$$CO + H_2O = CO_2 + H_2$$
(Shift reaction)

$$2H_2 + CO = CH_3OH$$
(Methanol Synthesis)

In the sequence, the methanol is dehydrated to produce Dimethyl-Ether which is posteriorly dehydrated to produce hydrocarbons, as shown in the sequence:

$$2\ CH_3OH = CH_3OCH_3 + H_2O$$
(Methanol Dehydration)

$$CH_3OCH_3 = C_2H_4 + H_2O$$
(Dimethyl-Ether Dehydration)

The methanol conversion to olefins into hydrocarbons is called Methanol to Olefins (MTO) or Methanol to Gasoline (MTG) technologies. The most known processes dedicated to converting methanol in hydrocarbons are the processes MTG™ developed by ExxonMobil Company and the MTO-Hydro™ process, developed by UOP Company. Figure 4 presents the process flow diagram for the MTG™ process by ExxonMobil Company.

The MTO technologies presents some advantages in relation to Fischer-Tropsch processes, once show higher selectivity in the hydrocarbon production, furthermore, the obtained products require lower additional processing steps to achieve commercial

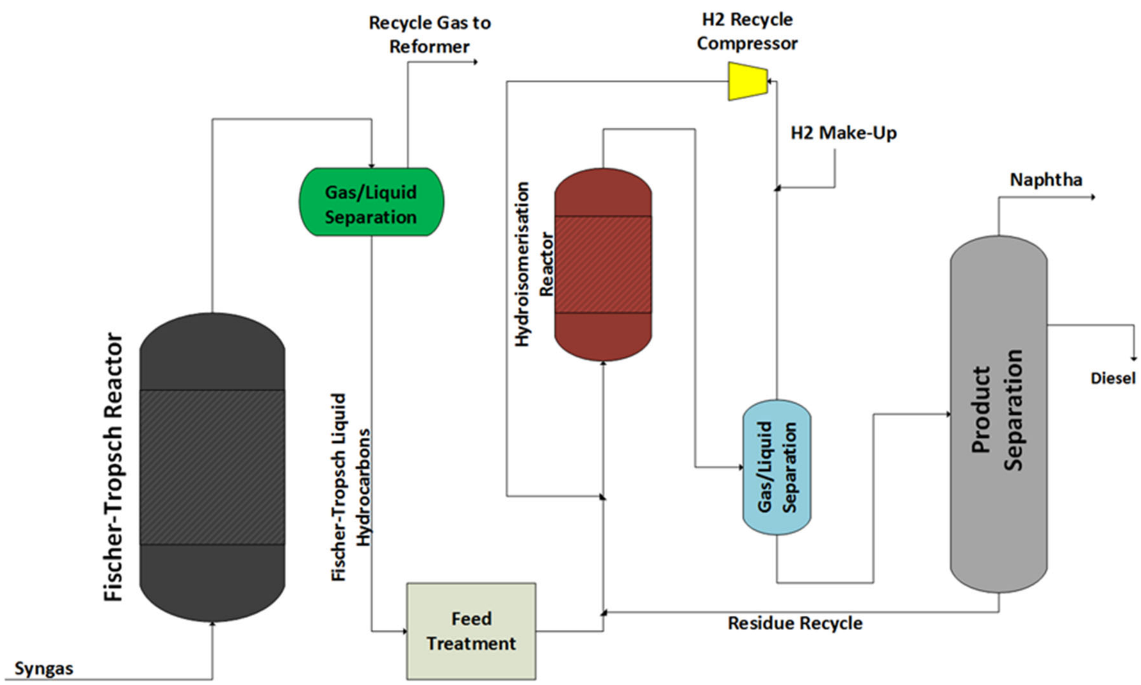


Figure 3 – Process Flow Diagram for GASEL™ Technology by Axens Company

specifications, another important point in that the installation cost is normally lower to MTO process plants when compared with FT units, once Fischer-Tropsch units are economically viable only in large scale. Regarding the olefins production, the maximization of these derivatives can be especially attractive in the current scenario where there is a trend of reduction in transportation fuels demand followed by the growing market of petrochemicals, creating the necessity of closer integration between refining and petrochemical assets aiming to maximize the added value, share risks and costs, as well as ensure market share in a highly competitive scenario of the downstream sector.

The streams produced by GTL technologies show reduced contaminants content (sulfur, nitrogen, etc.) as well as high-quality that makes these products attractive from the point of view of environmental footprint and profitability, these characteristics leads to lower operation cost when compared with the conventional crude oil refining route once it's needed a lower hydroprocessing capacity as well as less severe processes that can be a significant competitive differential. In scenarios with crude oil shortage or overprices, these technologies can be competitive and achieve strategically character to some nations.

Despite these advantages, the streams produced by Fischer-Tropsch process tends to present linear chain (essentially normal paraffin) that lead to poor cold flow properties. In market consumers with cold weather, the use of hydroprocessing units with dewaxing beds can be necessary to meet the quality regulations and ensure adequate performance of the final derivatives. Aiming to minimize this issue, some researchers are studying the upgrade of the traditional Fischer-Tropsch process through the use of a new catalyst. The catalyst Pt/H-ZSM-22 zeolite appears like the most promising aiming to isomerize selectively the n-paraffin produced by the Fischer-Tropsch process, improving then the quality and the added value of the produced streams, in this point is important to consider the high cost of the platinum catalyst being necessary an adequate economic evaluation to support the decision to catalyst change.

LOWER ENVIRONMENTAL FOOTPRINT

Another characteristic that can make the GTL technologies even more attractive in the next years is the increasing restriction to CO2 emissions from fossil fuel combustion, among then, the natural gas. The possibility to fix carbon through GTL process can represent an efficient and profitable alternative.

The environmental footprint is one of the great concern of the society related to the crude oil production chain and the use of gas to liquid technologies can minimize the greenhouse gases, allowing a more sustainable and profitable growing to the society.

CONCLUSION

As exposed above the gas to liquid technologies can represent an attractive alternative to some nations aiming to ensure a source of high-quality liquid hydrocarbons capable to sustain the economic development in a scenario of lack of crude oil resources and supply crisis as well as allow a higher value addition to his natural resources.

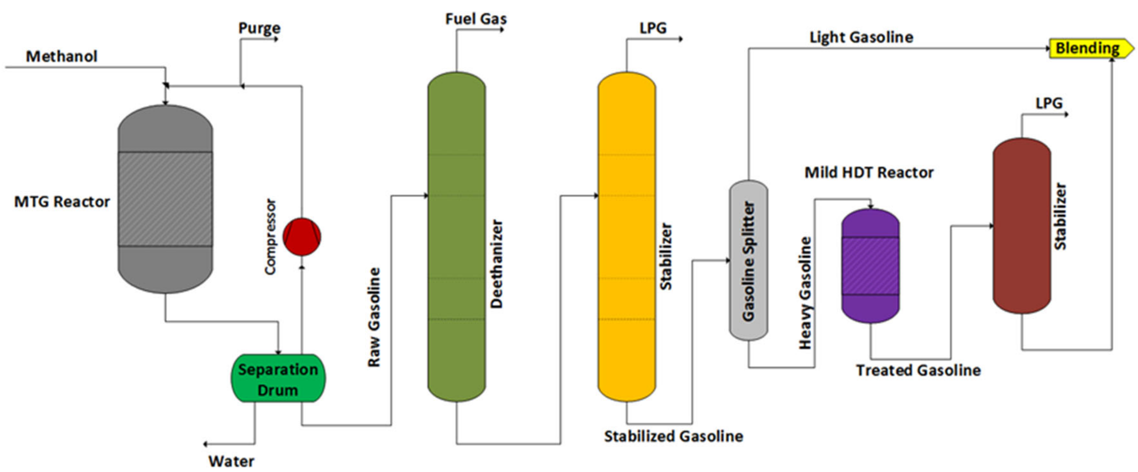


Figure 4 – Process Flow Diagram for MTG™ Technology by ExxonMobil

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