ENGINEERING PRACTICE

VOLUME 4 NUMBER 15

OCTOBER 2018

Distinguished Practicing Engineer Awards

IACPE

INTERNATIONAL ASSOCIATION OF CERTIFIED PRACTICING ENGINEERS

WWW.IACPE.COM

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KNOWLEDGE. CERTIFICATION. NETWORKING.

INTERNATIONAL ASSOCIATION OF CERTIFIED PRACTICING ENGINEERS

ABOUT

International Association of Certified Practicing Engineers provides a standard of professional competence and ethics. Identifies and recognizes those individuals that have meet the standard. And requires our members to participate in continuing education programs for personal and professional development.

In additional to insuring a professional level of competency and ethics the IACPE focuses on three major areas of development for our members: Personal, Professional, and Networking.

HISTORY

The International Association of Certified Practicing Engineers concept was formulated by the many young professionals and students we meet during our careers working in the field, running training courses, and lecturing at universities.

During question and answer sessions we found the single most common question was: What else can I do to further my career?

We found, depending on the persons avail able time and finances, and very often dependent on the country in which the person was from, the options to further ones career were not equal.

Many times we found the options available to our students in developing countries were too costly and or provided too little of value in an expanding global business environment.

The reality is that most of our founders come from countries that require rigorous academic standards at four year universities in order to achieve an engineering degree. Then, after obtaining this degree, they complete even stricter government and state examinations to obtain their professional licenses in order to join professional organizations. They have been afforded the opportunity to continue their personal and professional development with many affordable schools, programs, and professional organizations. The IACPE did not see those same opportunities for everyone in every country.

So we set out to design and build an association dedicated to supporting those engineers in developing in emerging economies.

The IACPE took input from industry leaders, academic professors, and students from Indonesia, Malaysia, and the Philippines. The goal was to build an organization that would validate a candidates engineering fundamentals, prove their individuals skills, and enhance their networking ability. We wanted to do this in a way that was cost effective, time conscience, and utilized the latest technologies.

MISSION

Based on engineering first principles and practical real world applications our curriculum has been vetted by academic and industry professionals. Through rigorous study and examination, candidates are able to prove their knowledge and experience. This body of certified professionals engineers will become a network of industry professionals leading continuous improvement and education with improved ethics.

VISION

To become a globally recognized association for certification of professional engineers.

LETTER FROM THE PRESIDENT

KARL KOLMETZ

Distinguished Practicing Engineer Awards

Dear Friends,

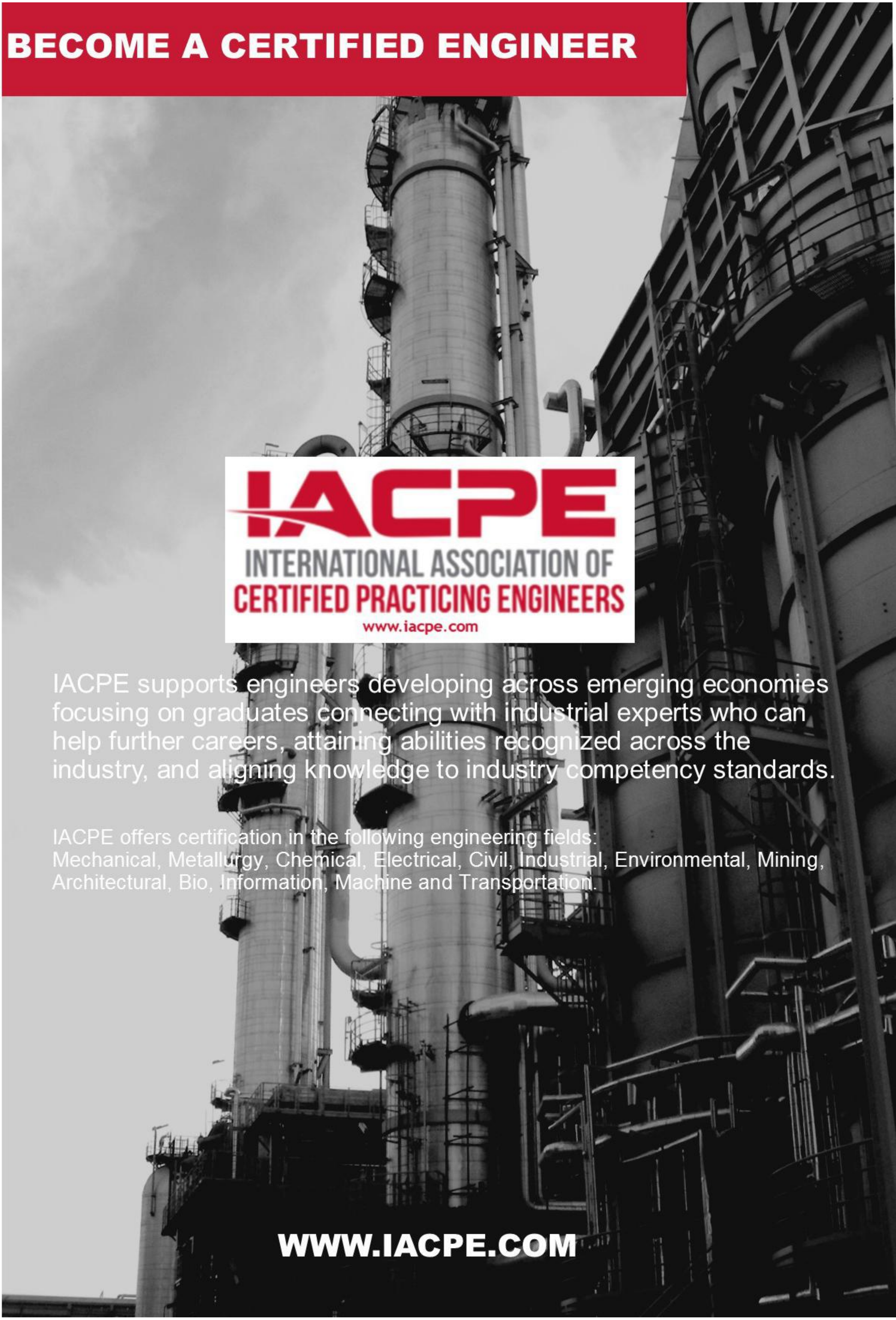
I hope you are doing great. This month we are pleased to nominate the International Association of Certified Practicing Engineers 2018 Distinguished Practicing Engineers. We have a great group of people that has assisted and mentored their friends and colleagues.

The IACPE will annually recognize the outstanding accomplishments of engineering education and engineering technology through the "Distinguished Practicing Engineer" awards program. By their commitment to their profession, desire to further the Association's Mission, and participation in civic and community affairs, IACPE award winners exemplify the best in engineering education and engineering technology.

This award will salute leaders in engineering for their dedication to their field and their commitment to advancing the human condition through great engineering achievement and/or through innovation in engineering education and technology. We will have an Academic Division, Technology Division, and Young Engineer Divisions. In the July Engineering Practice Magazine, we will nominate for each division and in the October Engineering Practice Magazine we will recognize the 2018 group of awardees.

We have begun build some IACPE Training Videos. There will some that will be training for the CPE Levels and some will be for continuing education certification. You can review the videos on our website video page and on U-Tube.

All the best in your career and life,
Karl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Distinguished Practicing Engineer Awards

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This award salutes leaders in engineering for their dedication to their field and their commitment to advancing the human condition through great engineering achievement and/or through innovation in engineering education and technology. There are three divisions: Academic Division, Technology Division, Young Engineer and Student Divisions.

ACADEMIC DIVISION

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Dr.Achmad Daengs GS.,SE.,MM.,CPPM.,CPE

Industrial Engineering
Dr. Dwi Yuli Rakhmawati, ST.,Ph.D.,CPE

Mechanical Engineering
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Alex Michinel

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Process Safety
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Information Technology
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Project Management
Dr. Marcio Wagner da Silva, MBA

Materials Movement
Vijay Sarathy

Process Simulation
Andrew Nathan

Commissioning
Jeff N Gray

DESIGN GUIDELINES FOR PROPYLENE SPLITTERS EFFICIENCIES

Karl Kolmetz CPE—KLM Technology Group
Contributing Authors: Timothy Zygula, Andrew Sloley, Randy Miller, Brian Clancy-Jundt, Daniel Summers

Introduction

Actual field tray efficiencies are affected by many factors. These include;

a) tower pressure
b) geometry and design of contacting equipment,
c) flow rates and flow paths of the liquid and vapor streams,
d) composition and properties of the vapor and liquid streams.

All these items can affect tray efficiencies and there are field examples were some have greatly impacted tray efficiencies. This paper will review some case studies and develop some design best practices.

We would like to thank our contributing authors who had added knowledge for this paper and support in our career. We would also like to thank Robert Miller and Simon Xu for their help and support.

Propylene Splitter Distillation Fundamentals

Propylene is a colorless, gaseous hydrocarbon. It is a petrochemical feedstock used primarily in the manufacture of plastics via polypropylene or cumene. It is also used to produce propylene oxide, acrylic acid, oxo alcohols and isopropanol. There are four grades of propylene that are sold; research grade (99.99% minimum purity); polymer grade (99.5% minimum purity); chemical grade (93-94% minimum purity); and refinery grade (60-70% purity).

There are over 250 Propylene Splitters all over the world. Most Ethylene Plants and large refineries have Propylene Splitters. They are the largest and tallest twin distillation columns in an Ethylene Plant.

Distillation is the separation of key components by the difference in their relative volatility, or boiling points. It can also be called fractional distillation or fractionation. Distillation is favored over other separation techniques such as crystallization, membranes or fixed bed systems when;

1. The relative volatility is greater that 1.2,
2. Products are thermally stable,
3. Large rates are desired,
4. No extreme corrosion, precipitation or sedimentation issues are present,
5. No explosion issues are present,
6. Low scale up cost factors - capacity can be doubled for about 1.5 additional cost,
7. Suitable for heat integration.

Close boiling mixtures may require many stages to separate the key components. For vapor and liquid equilibrium a K- value is defined for each species i by,

$$K_i = Y_i / X_i$$

where Y is the mole fraction in the vapor phase and X is the mole fraction in the liquid phase. (1)

For vapor liquid separation operations, an index of the relative ease of separation for two chemical species i and j is given by the relative volatility alpha defined as the ratio of their K values

$$\text{alpha}_{ij} = K_i / K_j = P_i / P_j$$

Pi and Pj are the vapor pressures of components i and j at a given temperature.

The number of theoretical stages required to separate two species to a desired degree is strongly dependent on the value of this index. The greater the departure of the relative volatility from a value of one, the fewer the equilibrium stages required for a desired degree of separation.

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Our services are provided to most industrial process markets including chemical / petro-chemicals, petroleum refining, alternative fuels, power generation, metals/mineral processing industries, foods, pharmaceuticals, and more.

PROCESS' personnel are experienced, highly-motivated Chemical Engineers with track records of significant achievement in the process engineering field.

PROCESS has experience in providing exceptional process design services for the development of conceptual and detail process design, process modeling and simulation, and operations support. Typical services include process option screening studies/evaluations, preliminary process designs (FEL & FEED), comprehensive Schedule A process design packages, detail engineering process support, process construction oversight, startup assistance, and more.

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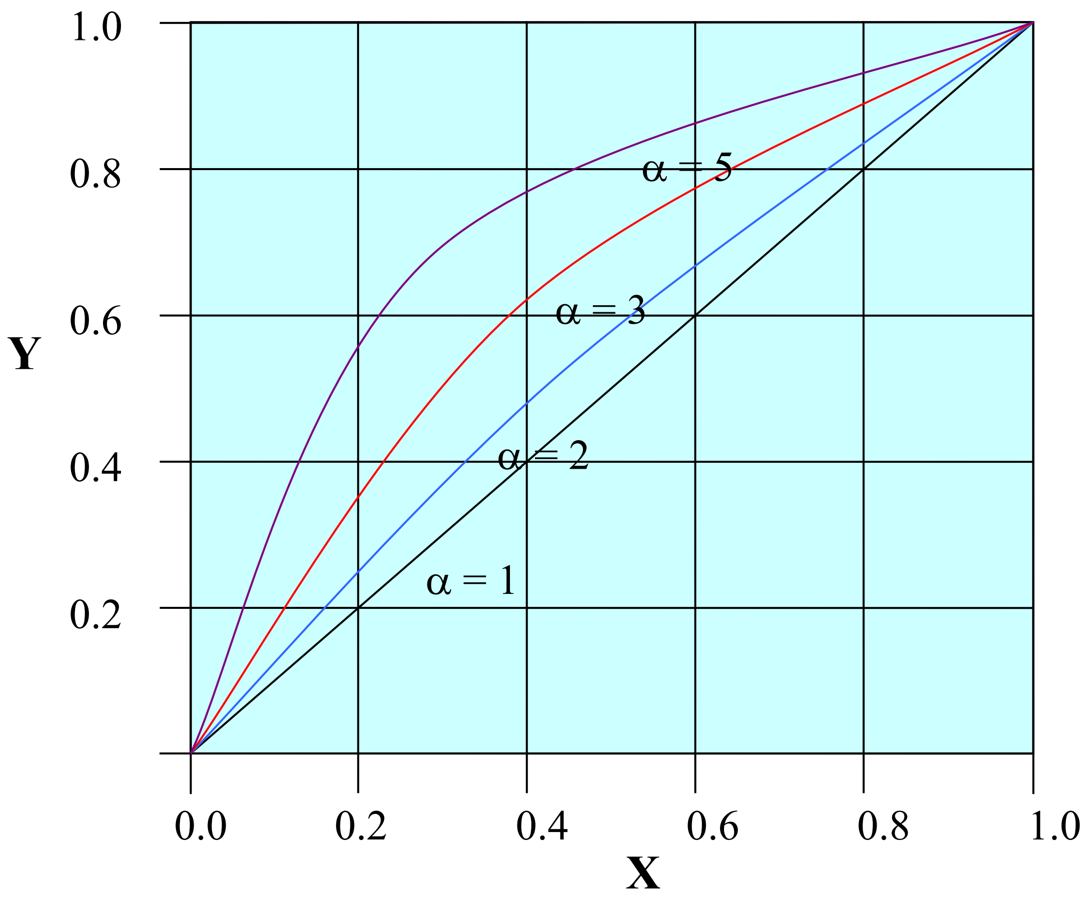

Knowing the relative volatility for a system is also useful in determining the amount of separation possible. A relative volatility of 1.0 indicates that both components are equally volatile and no separation takes place via normal distillation. When the relative volatility is low, less than 1.05, separation becomes difficult because a large number of stages are required. The higher the relative volatility, the more separable are the two components; this connotes fewer stages in a distillation column in order to effect the same separation between the overhead and bottoms products. Lower pressures increases relative volatilities in most systems.

The choice of the best application should be based on the life cycle cost. The life cycle cost is the initial capital cost of the plant along with the first ten years operating and maintenance cost. The life cycle cost should include a reliability factor, which is very important in designing any process plant equipment, reactors or separation equipment. Improved reliability has a very large impact on return on investment (ROI).

Many life cycle cost only review energy, but not solvent, adsorbent, or catalyst cost because of accounting rules and this can lead to skewed economic decisions. Accounting rules which list some items as capital cost and other items as operating expense need to be totaled or a skewed life cycle cost can be generated. A partial list would include;

1. Capital
2. Catalyst
3. Solvents
4. Energy
5. Maintenance
6.Industry average on stream factor (95% - 20 days per year)

For distillation the largest life cycle cost would be energy and maintenance concerns. Distillation is typically the single largest consumer of utilities in a chemical plant or refinery, and also the largest producer of finished product in most facilities. For energy cost a review of tray and packing efficiencies is warranted. For maintenance cost a review of reliability and simplicity is warranted. Distillation may be the most economical and is the most utilized globally to obtain improved purity products.

PYCOS

Modern Steam Cracking Furnace Kinetics Model

Overall Furnace Heat and Material Balance

PYCOSSW calculates the product yields for each feedstock, based on its conversion and selectivity. The selectivity is affected by operating conditions, coil geometry, steam-to-oil ratio and the coil outlet pressure. PYCOSSW calculates: pressure drop and average residence time inside the radiant coil; process heat required in the radiant zone; radiant coil heat transfer coefficient; coking rate; maximum tube wall temperature. PYCOSSW simulates the process along the entire radiant coil and calculates the fuel consumption by modelling the Firebox side. The coupled simulation of process and combustion sides allow calculating the process and metal temperatures along the entire length of the radiant coil and flue gas temperature at Bridge. PYCOSSW uses a proprietary routine to calculate the flame temperature. The program calculates the flue gas flow and enthalpy through the radiant and convection zones.

Dr. Maurizio Spoto Is Committed To Providing You With The Highest Quality Support For PYCOS.

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Some general estimates of tray efficiency might be

Demethanizer 65%	Air Separation 90%
Deethanizer 70%	C2 & C3 Splitter 85%
Depropanizer 75%	Stabilizer 80%
Debutanizer 80%	Hydrocarbon/Water 15%
Depentanizer 80%	EB/Styrene 90%
Low alpha Aromatics 80%	Alcohol - Water 75%
High alpha Aromatics 70%	Amine Contactor 33%

This data is for crossflowing trays and SRK Property Package. Best to compare on an equal basis. There is a general trend in this data. If the boiling points are close together like in a C2 and C3 Splitter (low alpha k), the separation will require many stages, but each stage will have a relatively high efficiency. If the boiling points are far apart like hydrocarbon and water (high alpha k) the separation will require few stages, but each stage will have a relativity low efficiency. A benzene water stripper might only require 5 to 7 stages in a simulation, but 30 trays in the field because of the low tray efficiency.

General Tray Efficiency

There are several general tray efficiency models. O'Connell Type Correlations can be used to predict the overall column efficiencies, many of which were developed in the 1940s and 1950s.

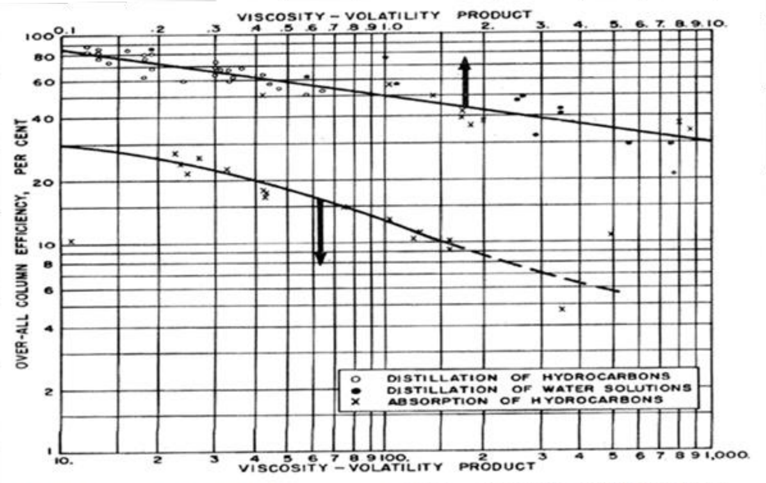


Figure 2.7: Lockhart and Leggett's augmented O'Connell correlation (Adapted from: Lockhart, F. J., and Leggett C. W. . (1958). *Advances in Petroleum Chemistry and Refining, 1*, 323-326)

One of the very first overall tray efficiency was the O'Connell Equation from 1946.

$$E_o = 49.2\ (\alpha\ \mu)\ {-0.245}$$

μ is viscosity of feed
α is relative volatility
both at average tower temperature.

when viscosity and/or relative volatility are increased tray efficiency is decreased.

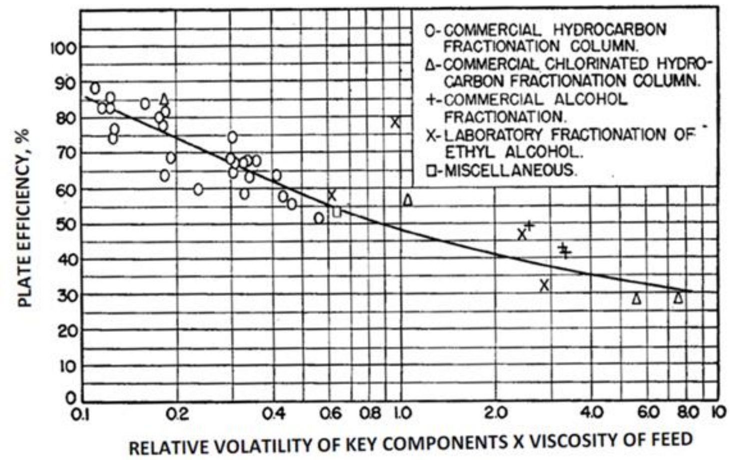
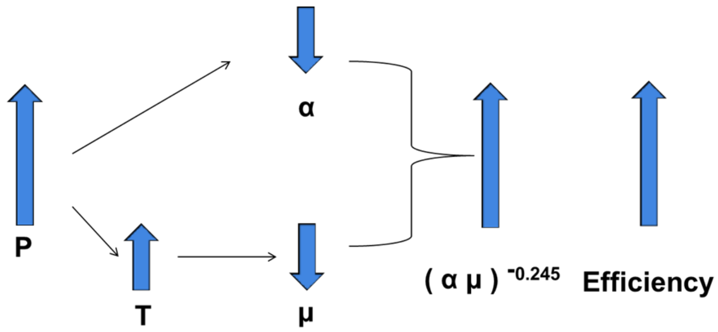


Figure 2.2: O'Connell correlation (Adapted from: O'Connell, H. E. (1946). Plate efficiency of fractionating columns and absorbers. *Transactions of the American Institute of Chemical Engineers, 42*(4), 741-755)

The Effect of Tower Pressure on Efficiency

For a fixed system (e.g. a C3 splitter), efficiency might go up with increased operation pressure as shown in the O'Connell Equation. This is true for many systems.



This pressure effect can be seen in C3 Splitters from the O'Connell Equation and Field Data.

PSIG	O'Connell	Field Tray Efficiency	
250	88	75-85+%	Numerous Papers
150	84	70-80%	Observed data
100	81	65-75%	Observed data
57		66%	AIChE 2011
50	75	60-70%	Observed data

The is data is for cross flowing trays and SRK Property Package. At very close alpha Ks there can be a large difference in property packages, as much as 5 to 10%. PR might be a more accurate property package, but make sure you are comparing apples to apples.

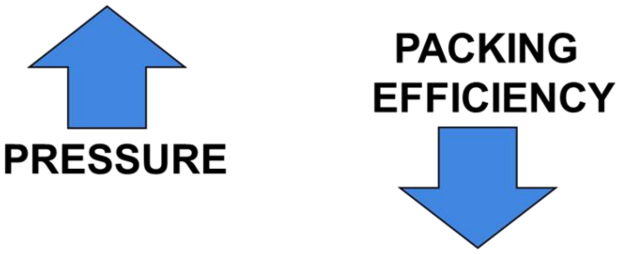
The end result is what counts – does the tower meet capacity and product purities. Each good

vendor utilizes a tuned model that give the proper end result.

There are over 200 C3 Splitter's in operation. O'Connell correlation works well for predicting the effect of pressure on efficiency.

Folklore and Myths can proclaim that lower pressure gives higher efficiencies – you need to review your system data – may not be true. This myth may have been developed from packing data, where HETP is much higher at lower pressures.

The field data for trays confirms that as pressure increases the efficiency increases. This is not the case for packing. There are two ideas of why this might be happening.



The first idea - some studies have showed that at higher pressures there appeared more liquid hold up on the packing, creating a larger boundary layer – leading to lower efficiency.

The second idea - there is a relationship between the vapor density / the liquid density and packing efficiency. At higher pressure the densities become closer together, leading to backing mixing effect of the liquid by the vapor - leading to lower efficiency.

Geometry and Design of Contacting Equipment

Once the preliminary tower diameter has been set the internals can be chosen. The task of choosing the type of tower internal to use is very important. The type of column internals used dictates a column's efficiency and capacity. All of the modeling and careful design work will mean nothing if the wrong type of column internals is chosen

The types of internals that have been used in propylene splitter columns are:

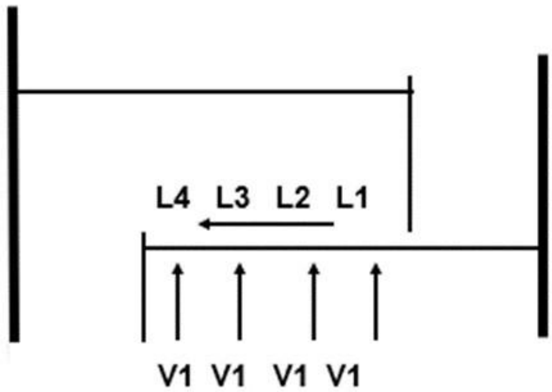
- Conventional Cross Flowing Trays
- Dual Flow Ripple Trays
- Packing
- High Capacity Trays
- Multiple Downcomer Trays

Conventional Multi-Pass Trays

Conventional multipass trays are typically used when a column is initially designed. Four pass or six pass trays are usually used because of their ability to handle high liquid loads like seen in propylene fractionation. The downside to using multipass trays is the reduction in separation efficiency that is experienced due to the reduction in active area and path flow length.

Great care must be taken when sizing downcomers in high-pressure distillation applications. The difference between vapor and liquid densities becomes smaller and separation of vapor from liquid in a downcomer becomes more difficult. This can result in increased aeration back-up and possible premature downcomer flooding. (2)

General Tray Efficiency may be determined by several formulas. There are two types of tray efficiency. There is the point efficiency and path flow efficiency. The point efficiency is where V1 meets L1. This is what is seen in dual flow trays.



V1 meets L1 – about 60 % Efficiency. Then L1 becomes L2. Then V1 meets L3 – about 65% Efficiency

$$E_{oq} = \frac{(Y_n - Y_{n-1})}{(Y^o_n - Y_{n-1})}\ \text{point}$$

Point Efficiency

$$E_{MV} = \frac{y_j - y_{j+1}}{y_j^{*} - y_{j+1}}$$

Overall Tray Efficiency

Where

$$y_j^* = K\,x_j\,,$$

x_j ***is liquid composition at DC outlet.***

Dual Flow Ripple Trays

Dual Flow Ripple Trays were installed in a few Propylene Splitters in the 1960s and in the 1990s. The challenge of dual flow trays are the hydraulic instability. The top of a distillation column will move as much at two feet in a wind storm. This movement at the top will cause the liquid to start down one side of the column and the vapor traveling up the opposite side of the column with limited mass transfer.

Dual Flow Ripple Trays Case Study

In a Malaysian ethylene plant, a two-column in series C3 Splitter was constructed to produce polymer grade (99.50 wt %) propylene. The towers were equipped with 258 dual flow trays. The trays are corrugated into a sinusoidal wave, with alternate trays installed with the waves at right angle.

Typical Dual Tray Loading Schematic

Notice:

1. **Froth Height**
2. **Rain Space**
2. **Corrugated Tray Deck**

The propylene service was commissioned in late 1999. It achieved both the nameplate capacity and propylene product quality. Unfortunately, the propylene loss in the propane recycle stream was observed to be significantly higher than the original design heat and material balance. This has resulted in an overall loss in propylene yield, higher purchased energy in the pyrolysis furnace and to a smaller extent, reduced the on-stream factor of the recycle propane gas pyrolysis furnace zone.

During a high load test carried out in July 2000, data was collected to pinpoint the high propylene loss was attributed to lower tray efficiency. By means of simulation to match the plant operating analyses, the efficiency was determined to be in the range of 45%. This is a significantly difference from the 65 - 70% tray efficiency assumed in the design. The tower effectively has less equivalent stages of fractionation and unable to achieve the desired separation. The average propylene in the propane recycle was averaging 45%, much higher that the designed 8%.

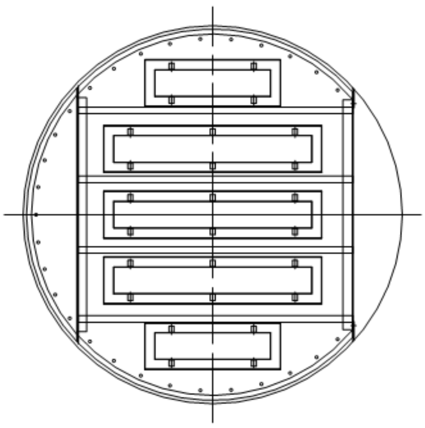
A gamma scan on the tower was carried out prior to a shutdown in early 2001 to eliminate potential tray damage. The scan showed all the trays were still intact. However, the liquid density profile showed mal-distribution occurring after the first 30 trays of each column. The decision was made to inspect the column on the results of the gamma scan.

The tower was opened for inspection during the February 2000 turnaround. The trays are were intact and level but large 6" I-beams and U-Channels were found laid perpendicular across the centerline of each tray. The I-beams and U-Channels effectively divided each tray into four quadrants.

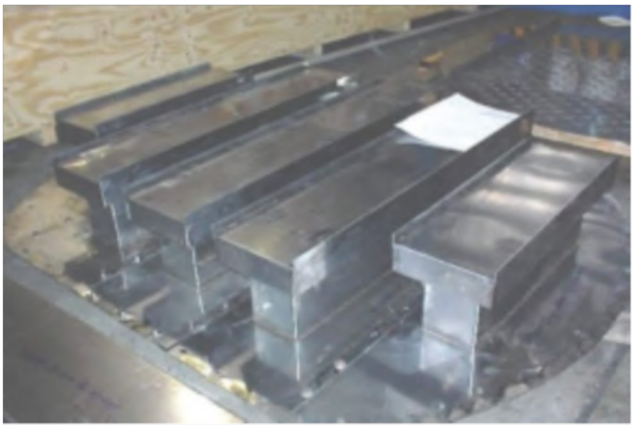
If the liquid flow was inconstant across the four quadrants, the gas flow will follow the path of least resistance further reducing the fractionation efficiency. A top of a column will move in a typical meteorological disturbance. This movement will cause the hydrologic load to migrate among the four quadrants. If any hydrologic flow instability were developed it would remain down the column. This hypothesis is consistent with the results from the gamma scan.

A decision was made to install six vapor and liquid re-distributors every thirty trays to correct any maldistribution that had occurred in the column. Additionally the U-Channel was constructed in three parts and the middle part of the U-channel was removed on each of the 14th and 15th trays between the re-distributors.

Schematic of Typical Re-Distributor Tray



Picture of Typical Re-Distributor Tray



Results

With the addition of these vapor and liquid re-distributors the tray efficiency of the column was increased 10% resulting in improved fractionation, even with the total reduction in the number of fractionation trays. The propylene in the propane recycle was reduced from 45% to below 10%. The tower maximum capacity before was 112%, and has presently run as high as 115% without reaching a limit.

Structured / Random Packing

There were two towers where structured packing was installed in Propylene Splitter Service and they were quickly replaced with the original trays. The structured packing was unable to meet the required product purity.

Structured packing was successful in very high pressure services like air separations, and low pressure applications like vacuum towers. It would be a logical assumption to expect that a medium pressure application (200 psig), like a propylene splitter would be an ideal application for structured packing. Both towers under preformed.

In the refinery where I worked in Houston, an alky debutanizer was converted from trays to random packing. Again a medium pressure application where random packing would be a logical assumption. The tower under preformed and the original trays were reinstalled.

What was discovered in these failures was that the pressure was not as important as the density difference between the vapor and liquid phases. If the vapor and liquid phase were very close in density, the vapor would back mix the vapor leading to low efficiency. In high pressure air separation there is a large difference in vapor and liquid density. In low pressure application there is a large difference in vapor and liquid density.

In medium pressure applications, above 150 psig (10 bar) the vapor and liquid density become close together and efficiently of the packing (HEPT) is reduced. It is seen in both structured and random packing applications. Low liquid-density/vapor-density ratios tend to create backflow in packed beds. Capacity and purities are often much lower than expected. Above a vapor density of 1.5 lb/ft3, packing may not be recommended. General guidelines are for the vapor to liquid ratio to be above 10.

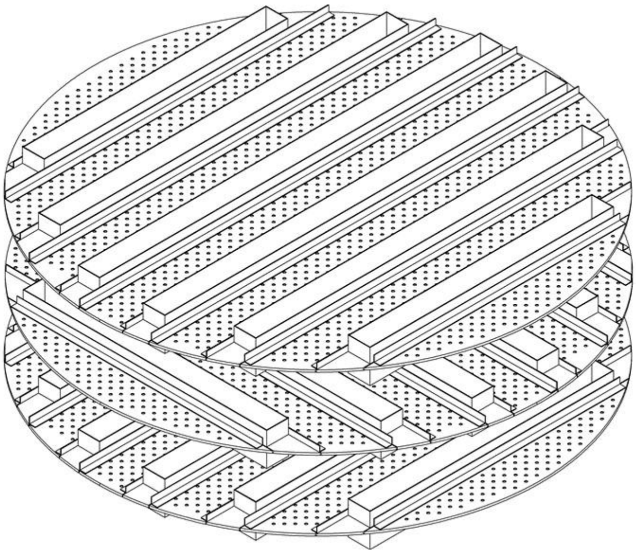
There is some documentation that has packing utilized in low pressure C3 Splitters.

Multi-Downcomer Trays

Multi-Downer trays are used for large liquid loads, particularly when the volumetric ratio between vapor and liquid rates is low. These situations occur in medium to high-pressure distillation, in absorption and stripping, and in direct contact heat transfer applications.

Multi-Downcomer trays can be used at close tray spacing. This will allow a reduction in both height and diameter of a new column compared to a column fitted with conventional multi-pass trays. Vessel shell costs can be significantly reduced with the use of Multi-Downcomer trays. When retrofitting an existing column with Multi-Downcomer trays, a significantly greater number can be installed, providing increased product purities and recoveries, as well as reduced reflux ratio for reduced energy consumption and/or increased column capacity.

The use of Multi-Downcomer trays has often reduced the number of columns needed in difficult separations, such as the fractionation of propylene-propane.



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Multiple Down Comer tray has less path flow length and lower efficiency. One rule is to keep the Path Flow Length above 450 mm (18 inches) to maintain good as possible efficiency. Efficiency is greatly reduced below 450 mm.

The question everyone ask - How much less efficient? You can hear numbers from 2% to 20% according to who you ask. One time a MD Salesman told me the efficiency loss was only 2%. Another time a cross flowing tray Salesman told me it was 20%.

The good news is that we have real data. There is some published efficiency data on Multiple Down Comer Trays

Date	Aug 92	June 95
Tower	C2 Splitter	C3 Splitter
Pressure	290 PSI	250 PSI
Reflux	237,834 kg	1,400,000 lb/hr
Trays	155	325
Capacity Gain	25%	35%
Efficiency	74%	74%

One might expect that a crossflowing tray at this pressure to be about 85% efficient.

Multiple Down Comer Trays – Best Practices

Understand there is a loss in tray efficiency – but because than may be installed on 18" trays spacing or less, you can install more trays and possibility increase overall tower efficiency – based on the reflux to stages curves. For the same tower shell diameter, capacity increase can be greater than 35%. If designed properly there can be an efficiency and capacity increase.

Design of a Propylene Splitter

High Pressure Verses Low Pressure Splitters

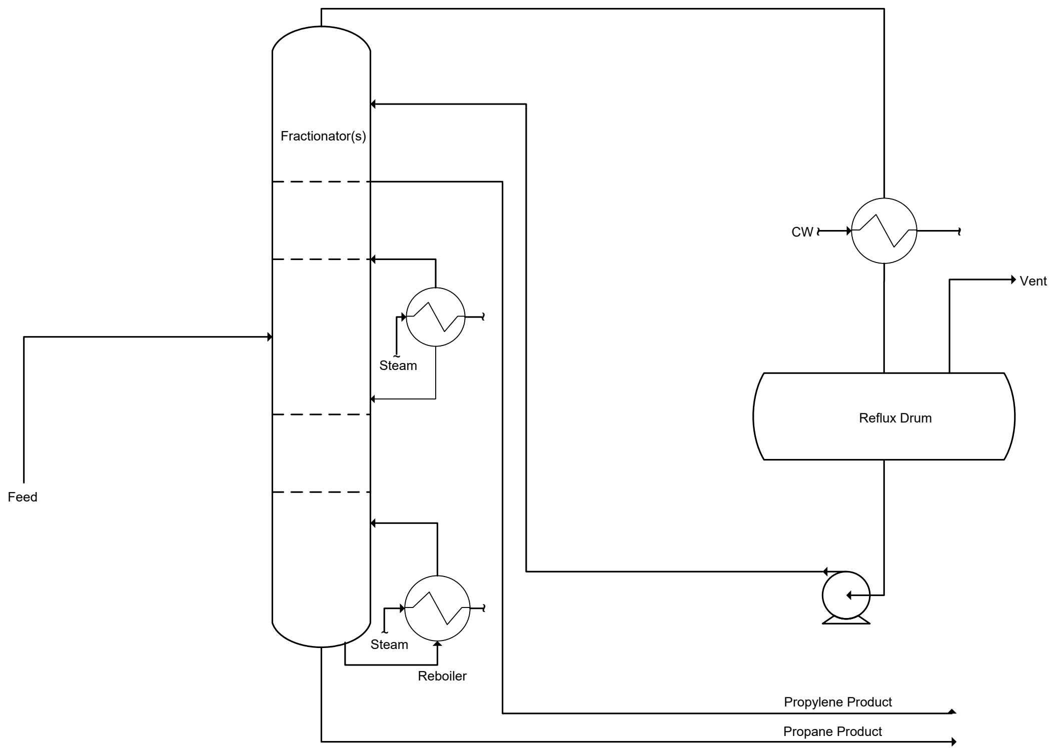
Determining the design of a Propylene Splitter requires an understanding of the simulation model used to generate the internal loads and physical properties, vapor and liquid equilibrium data utilized, tray hydraulics, and how the selection of the internals will affect the actual efficiency of the installed equipment in the field.

The typical design of a propylene splitter is not complex and there are two general variations in design. The first is a called high-pressure system, and the second is called a low pressure heat pumped system. A high-pressure system is designed to utilize cooling water as the source to cool the overhead vapor, and a high pressure is needed to condense the propylene vapor at ambient temperatures of about 40 degrees C.

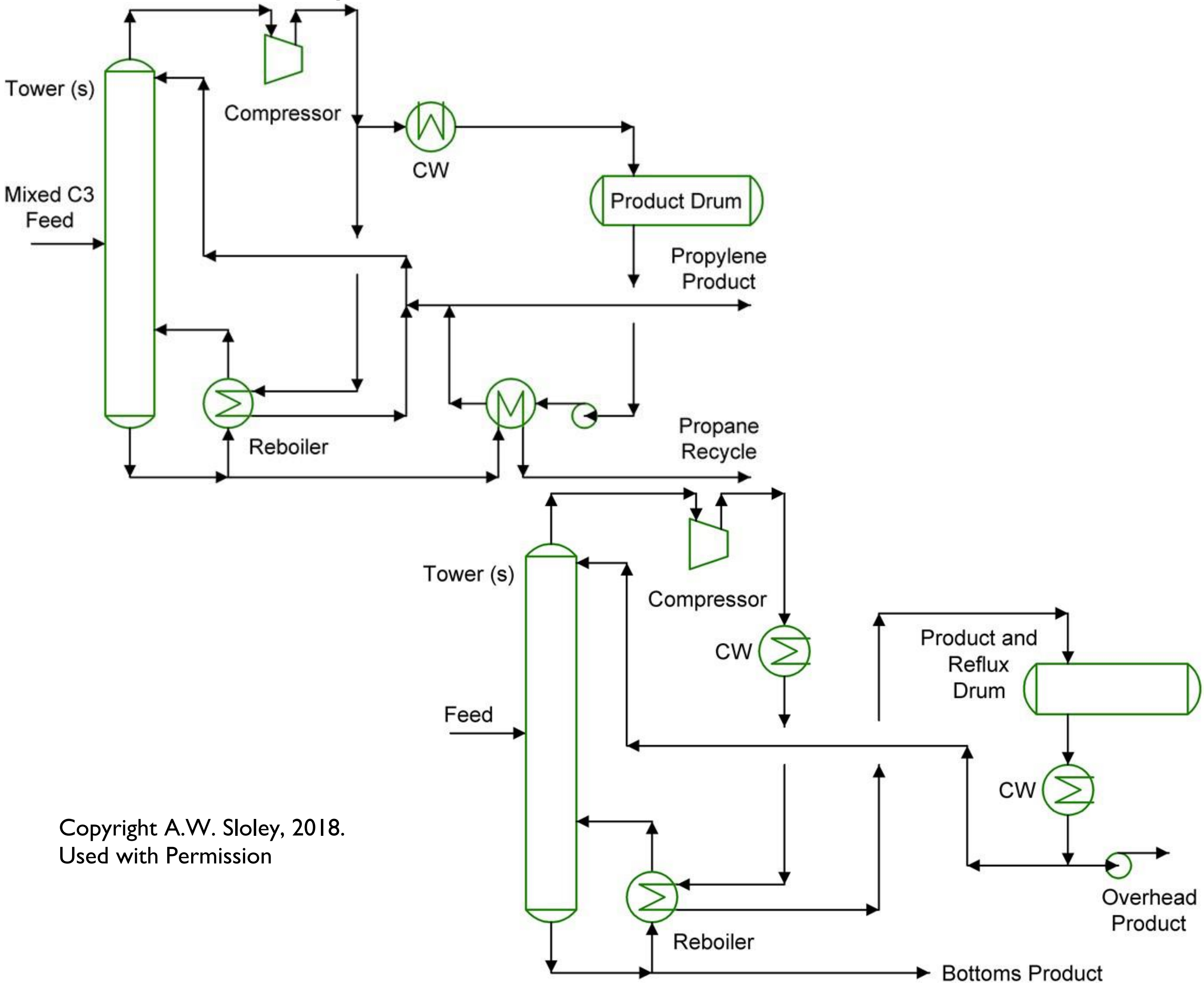
A heat pump system utilizes a compressor to reduce the tower pressure to allow the distillation column to be smaller in height, but larger in diameter. In most distillation application, relative volatilities can be improved by lowering the pressure. This results in lower number of stages required and reflux ratios, but at the cost of higher energy requirements of the compressor.

A good rule of thumb is that if the propylene system is associated with an ethylene plant, in which there is typicality an abundance of quench water that can be used to heat the C3 Splitter reboiler, a non-heat pump system may be the best choice. If no source of sufficient low-grade heat is available, for example in a refinery FCC unit, a revamp, or a propane dehydrogenation unit, then the use of a Heat Pump may be the economical choice. One should perform an economic analysis utilizing the six factors mentioned in the life cycle cost above.

High Pressure Systems



Low Pressure Heat Pumped Systems



There is a wide range of pressure choices for Propylene Splitters from 50 PSIG to 300 PSIG. What might be guidelines to choose the best pressure?

High Pressure Splitters

Advantages
- Ability to utilize cooling water for overhead condenser
- Ability to utilize medium level heat – there is a surplus in an Olefin Plant
- Higher Individual Tray Efficiency

Disadvantages
- Capital Cost – thicker tower shell and foundation

Low Pressure Splitters

Advantages
- Capital Cost – thinner tower shell and foundation
- Energy – if there is not a surplus of medium level heat – the compressor heat can utilized for the energy
- Lower reflux ratio – combination of J-T effect and relative volitivity
- Fewer stages – combination of J-T effect and relative volitivity
- Ability to utilize Packing

Disadvantages
- Capital, Energy and Maintenance Cost of a compressor
- Larger Tower Diameter

You may need a study to determine best pressure for your Splitter - one vendor recommends 90 PSIG and a second vendor recommends 110 PSIG.

- 90 PSIG is lower capital but higher energy cost
- 110 PSIG is high capital but lower energy cost.

Conclusions

Field efficiency of trayed towers, may increase with operational pressure, as shown in the O'Connell correlation and field data. Field efficiency of packed towers, from the data appears to going down with increasing operational pressure. Proper design and selection of trays, packings and internals are critical for success of distillation towers.

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Sulfur Recovery Technologies – Reducing the Environmental Footprint of Crude Oil Refining Industry

Introduction

One of the main criticisms to the crude oil productive chain and his derivates is the environmental impact of the industries activities related with the productive processes, mainly the refining step. Over the last decades, environmental regulations increasingly severe create the need to reduce the contaminant content in the final products, mainly sulfur and nitrogen compounds, aim to reduce the emission of environment harmful compounds as SOx and NOx.

One of the most important process units in the refining complex is the Sulfur Recovery Unit. This unit is responsible for recovery, in the elemental sulfur form, the sulfur removed from process streams treated in sweetening units of light fractions (LPG and Fuel Gas) produced in the deep conversion units as Delayed Coking, Fluid Catalytic Cracking and Hydrotreating, furthermore the gaseous streams produced in the sour water stripping unit are directed to the sulfur recovery unit.

Process Arrangement

The sulfur recovery unit feed stream, called sour gas is composed basically of H_2S (50-80%) and contaminants like CO_2, H_2O, NH_3, and hydrocarbons. The most employed technology to recovery sulfur in the refining industry is the Claus process. Claus process is based on two H_2S conversion steps, a thermal step followed by a catalytic step. In the thermal step the H_2S is partially burned according to chemical reaction below:

$H_2S + 3/2\ O_2 \rightarrow SO_2 + H_2O$ (1)

Then, the remaining H_2S reacts with the SO_2, producing Elemental sulfur according to following chemical reaction:

$2H_2S + SO_2 \leftrightarrow 3S + 2H_2O$ (2)

The global Claus process chemical reaction is represented below:

$3H_2S + 3/2\ O_2 \leftrightarrow 3S + 3H_2O$ (3)

Figure 1 shows a process flow diagram for a typical sulfur recovery unit.

The thermal step is carried out in the burner which operates under temperatures higher than 900 oC, close to 1/3 of H_2S is converted in SO_2, following the reaction 1, that is endothermic. This step is also responsible to destroy the sour gas contaminants as ammonia and hydrocarbons, the thermal step is responsible for 60 – 70% of sulfur recovery.

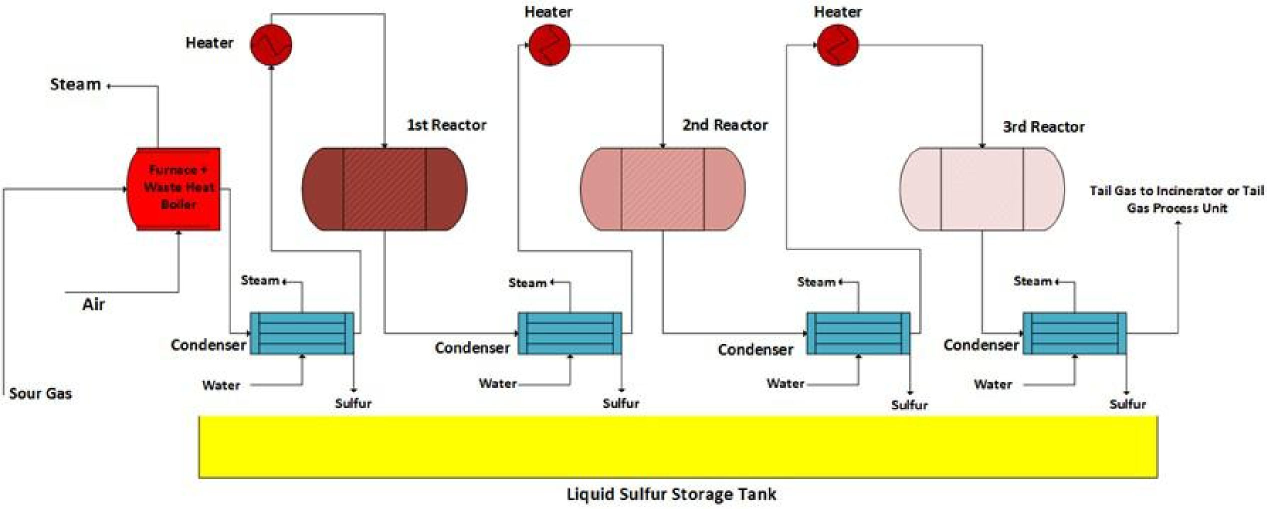

Figure 1 – Process Flow Diagram for a Typical Sulfur Recovery Unit

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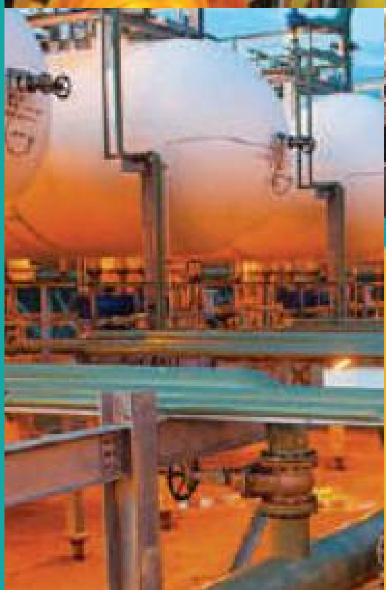

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The catalytic step is realized in fixed bed reactors containing TiO_2 or activated alumina as the catalyst. The catalytic step (reaction 2) is slightly endothermic when compared with the thermal step, so is carried out at lower temperatures (200 – 350oC). One of the main catalyst developers to Claus process is the Axens Company.

The cooling of process stream between the thermal and catalytic steps is realized in a waste heat boiler producing steam which is sent to consumers in other processes in the refinery. The process applies multiple reaction stages with the removal of produced sulfur among the stages aim to shift the chemical equilibrium to the products, in units containing three catalytic stages is possible recover 98% of the sulfur contained in the sour gas that is fed to the unit.

In order to comply with currently SOx emissions regulations, the sulfur recovery units normally needs a sulfur recovery efficiency between 99 to 99,5%. Aim to raise the sulfur recovery efficiency, the modern sulfur recovery units rely on tail gas treating units, as presented in Figure 2.

The tail gas treating unit receives the off-gas from the sulfur recovery unit and converts the remaining SO_2 and others sulfur compounds in H_2S that is sent back to the sulfur recovery unit, raising the sulfur recovery efficiency.

This process consists of a heating step of the residual gas that raises the temperature over the sulfur condensation temperature, avoiding that this phenomenon occurs in the reactor and supply the energy need to the conversion reactions, in this step still occurs the hydrogen production which act as reduction gas to convert the sulfur compounds to H_2S in the catalytic process step.

Following, the tail gas receives an injection of hydrogen before to enter to the fixed bed reactor containing Cobalt and Molybdenum catalyst (CoMo), in this step the sulfur compounds are converted to H_2S according to following chemical reactions:

$$SO_2 + 3H_2 \rightleftarrows H_2S + 2H_2O$$

$$S + H_2 \rightleftarrows H_2S$$

$$COS + 4H_2 \rightleftarrows CH_4 + H_2S + H_2O$$

$$CS_2 + 4H_2 \rightleftarrows CH_4 + 2H_2S$$

H_2S produced is processed in amines treating columns to purify this compound and then the H_2S is sent back to the sulfur recovery unit, the other gases are sent to an incinerator.

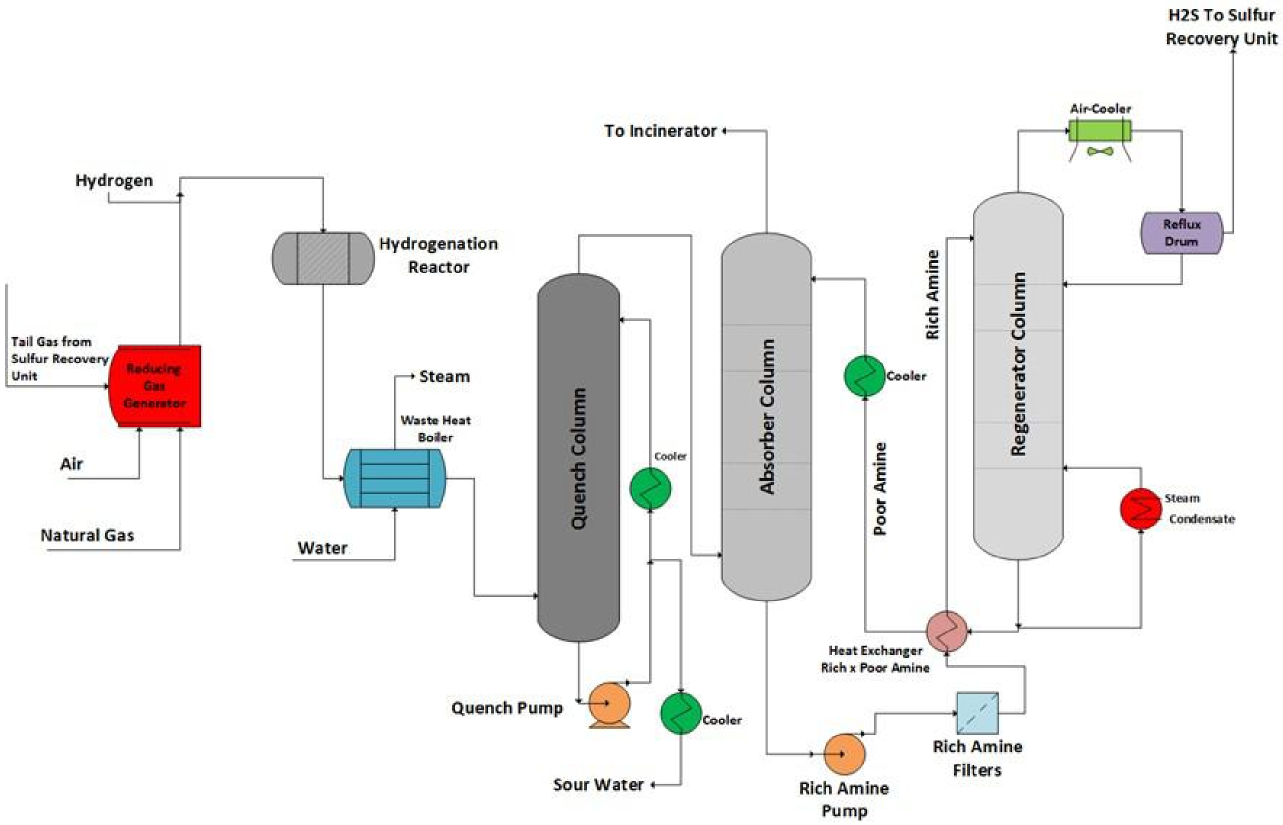

Figure 2 – Process Flow Diagram to Typical Tail Gas Treating Unit

Available Technologies

Among the available tail gas treating technologies its possible quote, the SCOT ™ (Shell Off-Gas Treater) licensed by Shell Company, the RESULF ™ process, developed by CB&I company and the FLEXSORB ™, developed by EXXONMOBIL.

The sulfur recovery complex is extremely important to adequate modern refinery operation, normally operational instabilities and shutdown of sulfur recovery units force a reduction in the flow rate and in some cases the shutdown of refinery process units impacting significantly the refiners profitability.

The main process variables of the sulfur recovery units are the air/sour gas ratio and the temperatures of the combustion chamber, in the reactors in the catalytic step and the condensers temperatures. The air flow rate supplied to the process need to be sufficient to burn completely the hydrocarbons and NH3 present in the feed stream plus the necessary to convert the third part of H2S in SO2. Combustion chamber temperature is normally sufficiently high to promote the Claus process reactions and to destroy the sour gas contaminants (NH3 and hydrocarbons), in refineries that apply sour water stripping units with a single tower, the NH3 content in the sour gas is higher, in this case, the combustion chamber is normally higher (above of 1100 oC).

One of the main operational problems of sulfur recovery units is related to processing sour gases containing high hydrocarbon content that raises the air consumption and can lead soot formation (heavier hydrocarbons) which provokes catalyst deactivation and high pressure drop in the reactors, furthermore requiring higher temperatures in the Claus process thermal step.

Nowadays, aiming to increasingly minimize the environmental impact of the refining processes, some licensers have devoted his efforts to developed new technologies focused on sulfur and nitrogen recovery from waste gases produced in the refining processes. Among these technologies, one of the most economically attractive is the SNOX ™ technology, developed by Haldor Topsoe Company.

In the SNOX ™ process, the sulfur is recovered as a highly concentrated sulfuric acid that can be commercialized directly by the refiner while the nitrogen is eliminated in the nitrogen form (N2), which is not harmful to the environment.

Conclusion

Sulfur recovery units, as aforementioned are fundamental to the operation of modern refining complexes. The produced sulfur is normally commercialized to produce sulfuric acid and fertilizers, that is, beyond minimizing the refining process environmental impact the sulfur recovery units are capable to add profitability to the refiner.

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Biography

Dr. Marcio Wagner da Silva is Process Engineer and Project Manager focusing on Crude Oil Refining Industry based in São José dos Campos, Brazil. Bachelor in Chemical Engineering from University of Maringa (UEM), Brazil and PhD. in Chemical Engineering from University of Campinas (UNICAMP), Brazil. Has extensive experience in research, design and construction to oil and gas industry including developing and coordinating projects to operational improvements and debottlenecking to bottom barrel units, moreover Dr. Marcio Wagner have MBA in Project Management from Federal University of Rio de Janeiro (UFRJ) and is certified in Business from Getulio Vargas Foundation (FGV).

Tuning PID Controllers using Internal Model Control

Jayanthi Vijay Sarathy, M.E, CEng, MIChemE, Chartered Chemical Engineer, IChemE, UK

Abstract:
There is no engineering system today where controllers are not used. Controllers are everywhere. To successfully keep a controller in doing what it is supposed to do, controllers need to be tuned properly and hence the name 'Controller Tuning'. There are enough text books out there to explain the theory and complex equations behind controller tuning but from a practical stand point in the industry, it becomes increasingly difficult to apply them in its raw form with a paper, pencil and calculator while on the job.

The following article is based on using a well known software tool called Aspentech HYSYS# that cuts down the mathematical headache to get your process facility to run the way you'd expect it to. A set of hand calculations is also included to understand how the controller is tuned.

Design Procedure for Controller Tuning

A good controller is one that offers a suitable trade-off between performance and robustness and the standard type of controllers used even to this day are the proportional (P), proportional plus integral (PI), and the proportional plus integral and derivative (PID) controllers. To tune these controllers, 'Gain' (Kc), Integral Time' (Ti) and 'Derivative Time' (Td) are the three basic parameters needed and a procedure is required to estimate it. In this article, the procedure employed is called Internal Model Control (IMC).

What is Internal Model Control

IMC refers to a systematic procedure for control system design based on the Q-parameterization concept that is the basis for many modern control techniques. What makes IMC particularly appealing is that it presents a methodology for designing Q-parameterized controllers that has both fundamental and practical appeal. As a consequence, IMC has been a popular design procedure in the process industries, particularly as a means for tuning single loop, PID-type controllers. The Author uses a case study to demonstrate this.

Case Study

An inlet stream of 100 kgmol/hr at 5 barg containing 100 mol% methane passes through four valves. The pressure drop experienced at the four valves (VLV-100/VLV-101/VLV-102/VLV-103) is 0.1 bar, 0.2 bar, 0.3 bar and 0.4 bar respectively to exit at a final pressure of 4 barg at the outlet of VLV-103. VLV-100 has a hold-up volume of 2 m3 declared to that accounts for the delay in the output flow variation.

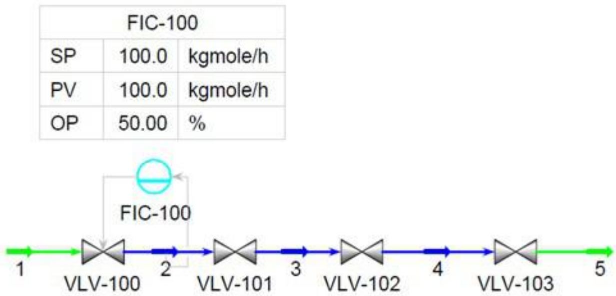

Figure 1. Process Control Scheme

The minimum and maximum flow through valve VLV-100 is taken to be 0 kmol/h and 200 kmol/h respectively. The control valve (VLV-100) opening is modelled to transport 100 kmol/h at 50% opening. A closed loop, reverse acting flow controller (FIC-100) is added to VLV-100 and needs to be tuned as a P-I controller.

The purpose of FIC-100 is to ensure that the flow through the control valve VLV-100 is always maintained at 100 kmol/h by adjusting the valve opening whenever the flow into it changes with time. Therefore the work to be done is to estimate what are the 'Kc' and 'Ti' values of the P-I controller that ensures a flow of 100 kmol/h is maintained by FIC-100 irrespective of the disturbances.

Basics of a PID Controller

Before trying to model and tune a controller, understanding a few basic terms helps. A controller is defined by three parameters - Set Point (SP), Process Variable (PV) and Output (OP).

1. Set Point (SP) - The parameter that you want to maintain. In this example, the flow rate of 100 kmol/h becomes the set point.

2. Process Variable (PV) - The actual flow passing through the valve which can vary. In this example, PV becomes stream '2' through which the pressure driven flow can vary.

3. Output (OP) - What the controller needs to change in order to ensure only 100 kmol/h of flow passes through the valve, VLV-100. In this example, OP is the valve opening that ranges between 0% to 100% opening.

4. Proportional Control (P) - "How Far" the measured process variable (PV) has moved away from the desired set point (SP).

5. Integral Control (I) - "How Long" the measured process variable (PV) has been away from the desired set point (SP).

6. Derivative Control (D) - "How Fast" the error value changes at an instant in time.

7. PV Min - Minimum Process Variable (PV Min) is the lowest amount of flow that can pass through the valve. In this case, the minimum flow is 0 kmol/h.

8. PV Max - Maximum Process Variable (PV Max) is the highest amount of flow that can pass through the valve. In this case, for the sake of the exercise, the maximum flow is taken to be 200 kmol/h. Note: The maximum flow through a valve depends on the size of the valve and is represented by the term 'Cv'. Higher the Cv, larger is the valve and greater is the flow through the valve.

9. Process Time Constant - This describes how fast a measured process variable responds when forced by a change in the controller output. The Process Time Constant is equal to the time it takes for the process to change to 63.2% of the total change in the measured process variable.

In the current case study, FIC-100 in the image is modelled as a P-I controller.

How is a Controller Tuned

A controller is tuned by causing a disturbance to the process and seeing how it reacts. Watching its reaction, the control parameters (Kc, Ti) can be adjusted to ensure the controller behaves as intended.

In the current undertaking, the disturbance is introduced as - the valve (OP) which is at 50% opening

for a flow rate of 100 kmol/h (PV) is momentarily changed manually to 60% opening (OP).

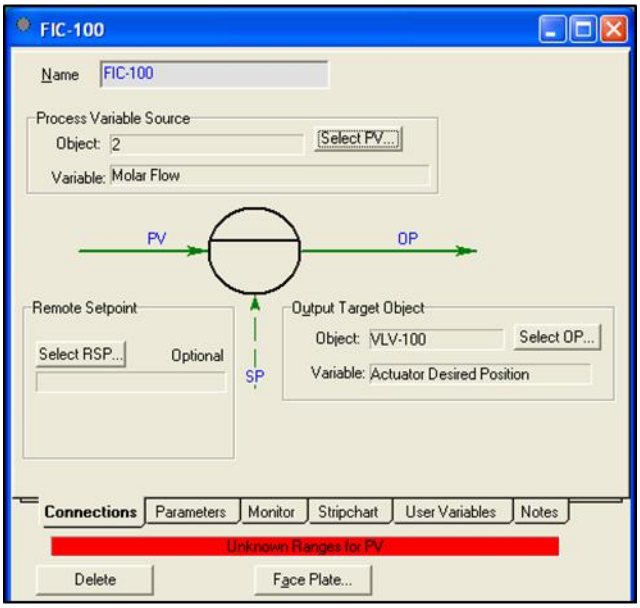

Figure 2. Controller Connections

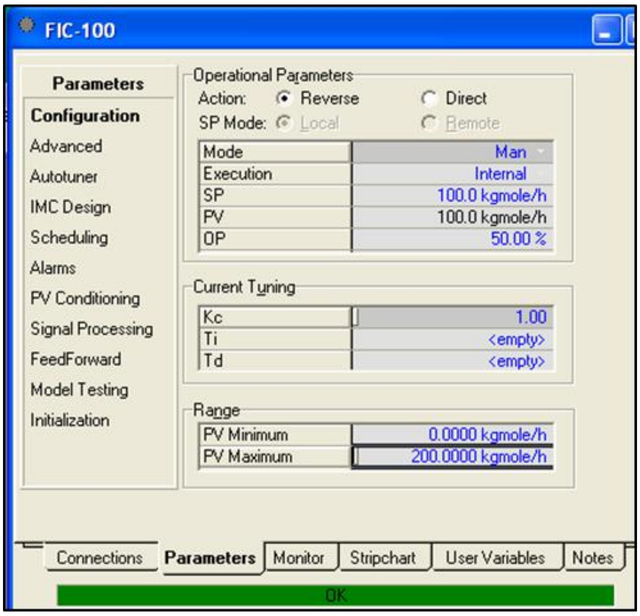

Figure 3. Controller Configuration

Upon doing so, the flow through the control valve, VLV-100(PV) is monitored as shown in the next figure.

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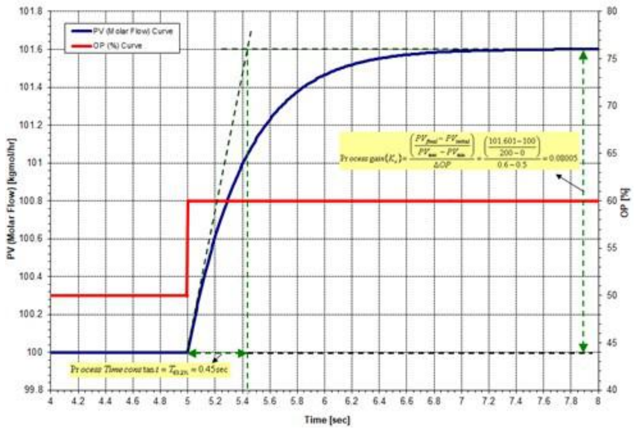

Figure 4. Controller Output with OP variation from 50% to 60% (Manual Mode)

To tune the controller using IMC design procedure, two steps are employed

1. Estimate the value of Process Gain (Kp) and Process Time Constant [Tp] or T63.2%
2. Estimate the value of Overall Gain (Kc) and Integral Time (Ti) i.e., P-I action.

From the above graph, Process gain (Kp) and T63.2% is calculated as,

$$Process\ Gain\ [Kp] = \frac{\left[\frac{PV_{final} - PV_{initial}}{PV_{Max} - PV_{Min}}\right]}{\Delta OP} = \frac{\left[\frac{101.6 - 100}{200 - 0}\right]}{0.6 - 0.5} = 0.08005\ \frac{\%}{\%}$$

$$T_{63.2\%} = 0.45\ sec$$

Using the above values of Kp and T63.2%, the overall gain (Kc) and Integral Time (Ti) as a first order process with negligible dead time can be calculated as,

$$Overall\ Gain\ [K_c] = \frac{T_{63.2\%}}{K_p(\lambda + T_d)} = \frac{0.45}{0.08005 \times (0.45 + 0)} = 12.5\ \frac{\%}{\%}$$

$$T_i = T_{63.2\%} = 0.45\ sec$$

Using the calculated value of Kc=12.5 and Ti of 0.45 sec, the controller is programmed accordingly and executed to run.

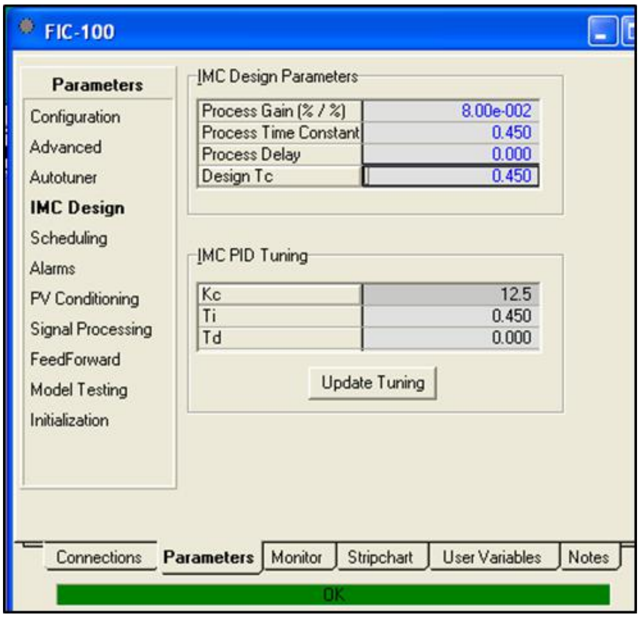

Figure 5. IMC Design Tab

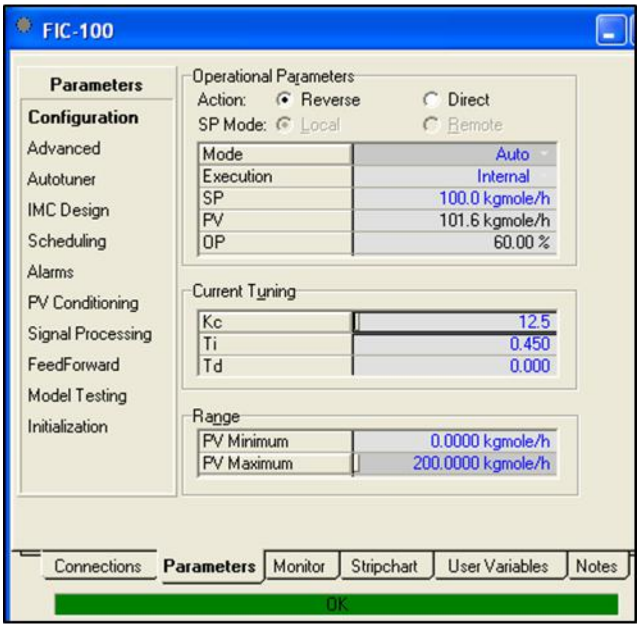

Figure 6. IMC Tuned Controlled Values

The response of the tuned P-I controller (FIC-100) when activated to run in automatic mode, i.e., when the controller is set to run on its own, gives us the result where the FIC-100 automatically adjusts the valve (VLV-100) % opening to ensure 100 kmol/h of flow passes through the system as follows.

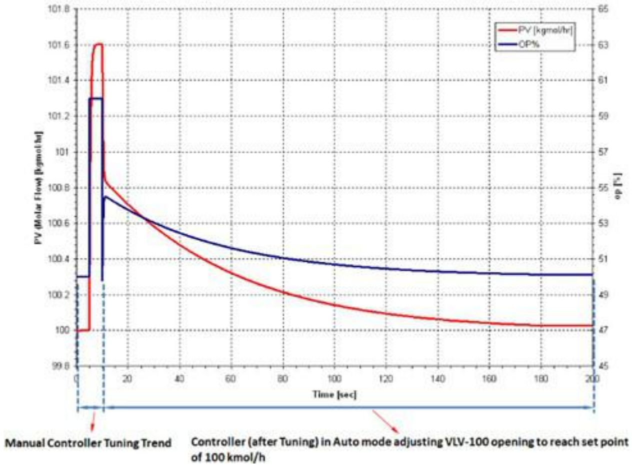

Figure 7. Controller Output with OP Variation with Tuned Controller Values (Auto Mode)

From the above graph, it is seen that the controller FIC-100, automatically runs to bring the valve to 50% opening with 100 kmol/h of flow passing through VLV-100 in about roughly 180 sec (~3 min). Using the tuned controller, with any changes hereafter in the incoming flow would mean, the flow controller (FIC-100) can automatically dictate by how much the control valve (VLV-100) has to open or close to ensure a flow of 100 kmol/h is maintained.

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Biography

Vijay Sarathy holds a Master's Degree in Chemical Engineering from Birla Institute of Technology & Science (BITS), Pilani, India and is a Chartered Engineer from the Institution of Chemical Engineers, UK. His expertise over 10 years of professional experience covers Front End Engineering, Process Dynamic Simulation and Subsea/Onshore pipeline flow assurance in the Oil and Gas industry. Vijay has worked as an Upstream Process Engineer with major conglomerates of General Electric, ENI Saipem and Shell.

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For Engineers; Because Safety Is Part Of The Process!

By: Chris Palmisano, MESH, IFSAC **October 2018**

Understanding X-MOD Is Critical To Any Successful Safety Program.

Like An Iceberg, X-MOD Is A Much Bigger Problem Than You Might Think.

If Safety and Risk Management Programs are going to be successful, it must include knowledge of the many components of Workers Compensation (WC) Insurance and that includes a good understanding of X-MOD, better known as your company's Experience Modifier Rate.

Many Safety Professionals go day-to-day managing their safety programs, formalizing their safety committees, conducting frequent and regular inspections, accident investigations and training employees yet, have little or no involvement in the company's workers compensation insurance process. So what is X-MOD?

Experience Modifier Rate AKA, EMR, E-MOD, X-MOD, MOD – is a rating or numeric representation of your injury claim performance, best explained as a factor in an employer's annual insurance premium cost for their worker's compensation coverage. The rate is based on your **history and/or frequency of employee injury claims filed to your Insurance Carrier.** The bottom line is, the more claims you file, the higher your X-MOD Rate. It is really that simple. Think of your X-MOD as an interest rate on a credit card. The higher the interest rate, the more you pay. Here are some examples of what you can expect from your X-MOD Rate based on its value:

An X-MOD of 0.78 would be considered a Credit MOD or Low MOD to the Insurance Carrier, meaning you may get a discount on your premium, because you likely have very few injuries. In others words, you are a Good Risk.

An X-MOD of 0.91 may mean that you are viewed as average to the Carrier and therefore your premium would be uniform across your industry, because you are viewed as an Average Risk.

An X-MOD of 1.51 is considered Poor. It is an unfortunately High Debit MOD, which means you are likely being quoted exponentially high insurance premiums, compared to your peers, because you are viewed as a High Risk. Additionally, OSHA may use a High X-MOD (which is public information), as reason to come and inspect your operation.

One important thing to know about your X-MOD, is that it takes approx. three long years of good behavior to recover from a High Debit Rate. Therefore, some Insurance Carriers are willing to embrace employers with High Debit X-MODs. Here's why; if your company has generally been a good Risk but just received a newly assigned "High X-MOD Rate" after your many years of good performance, you might be an attractive gamble to an Insurance Carrier. A High X-MOD is typically a wake-up call for most companies to improve worker safety. At the same time, the Insurance Carrier gets to collect approx. 3 years of high premiums, while you struggle to reclaim your old lower X-MOD Rate. So as you can see, while there is risk for the Carrier to have skin in the game, the scenario of covering an employer with a High X-MOD can serve to be profitable for the Insurance Carrier.

Most employers stricken with a High X-MOD, do improve their safety performance and those that don't, are viewed as uninsurable Risks by some Carriers. So, what is the solution to the dilemma of having a High Debit X-MOD Rate? Having a robust safety program helps to significantly minimize the odds of getting yourself into a High Debit X-MOD Rate situation because it's the best way to reduce injury claims. Here are some tips to get you there.

1. You must have a formalized Safety Committee that meets regularly. To be effective the Committee should meet at-the-least monthly. Quarterly or Biannual safety meetings are by no means an effective part of risk management. As an example in the Human Resources Director misses a quarterly safety meeting because he/she is on vacation, six months have gone by without that person engaged in a company safety meeting. A lot of bad things can happen in six months, which is why I promote the monthly safety meeting. Keep meeting minutes. They can prove valuable in OSHA investigations and even in allegations of negligence, when an employee files a disability claim.

2. Develop written safe programs. Programs collecting dust on a shelf do little to protect your employees. The programs need to be reviewed and discussed often and should be part of your ongoing safety training and new hire orientation programs.

3. Provide regular training and include weekly/monthly safety talks for employees. Safety talks should be provided as "leaders" into times of high risk exposure. For example, provide a heat related illness safety talk in the early summer, before the heat strikes your area. It's too late to give a safety talk on heat illness, after a heat related illness injury occurs. Another example might be to offer a safety talk on the hazards of driving on icy roads, given just before cold weather arrives or, offering a talk on proper lifting, ahead of an anticipated large work order or busy season. In other words, anticipate your employee's risks, prior to the exposure and offer what they can do to protect themselves.

4. Conduct frequent and regular inspections and task observations to minimize risk exposures and promote safety in a positive way. Don't make these inspections finger pointing sessions. To be effective, make these checkups helpful reminders of the rules to work safe. Show employees you care about their safety, not so much about assigning blame or discipline.

5. Speaking of discipline, it pays to have a written discipline program that is progressive, giving employees a chance to improve, before throwing them out the door. It needs to be understood, that working safely is a condition of employment. Avoid relying on just safety incentive programs. Employees need to recognize that they are responsible for their actions.

6. Have an accident reporting process. Select a medical provided that you have vetted. Don't be fearful to ask them their philosophy on returning employees to work as soon as medically able, which is important in protecting your X-MOD Rating. Employees remaining on Days Away Status for extended periods of time, serve as a disadvantage for everyone, including the injured employee.

7. Use accident reports to submit claims to the Insurance Carrier quickly. Review these reports with the Safety Committee to address the root cause of accidents and to spark discussion on how you can develop plans to protect employees from similar types of exposures in the future.

8. I can't say enough about having a good RTW (Return-to-Work Program) for employees that are placed on Light Duty Status. Let me say that again, HAVE A WRITTEN RTW (RETURN-TO-WORK PROGRAM) FOR EMPLOYEES THAT ARE PLACED ON LIGHT DUTY STATUS. This bears repeating, because if there is one question just about every Insurance Underwriter will ask before giving you a quote for workers compensation coverage it is, "Do you have a Return-to-Work Program"? You always want to answer this question with, "Yes!" followed by, "Here's a written copy of our program".

9. Have solid HR programs that include well written job descriptions, post offer medical fitness evaluations, background checks and substance abuse testing programs. All are important program to an Insurance Underwriter.

10. Good housekeeping programs promote a safe and healthful workplace as well as improve moral. If you want employees to feel good about their work environment and make professional decisions, make the workplace clean and professional.

I will close with this final thought, best offered in the form of a question and a simple answer for a company faced with a High Debit X-MOD situation. Other than good behavior, what can a company do to reduce insurance premiums during a recovery period from a high X-MOD? The answer is simple, "Leverage". You need leverage to convince the Carrier that you are a good risk. The best leverage you can have with an Insurance Carrier is, at-the-least, implementing the programs listed above. May you all have safe and restful holiday season, free of High Debit X-MODs in the New Year.

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