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Building Operational Excellence

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Introduction

Since 2005, KLM Technology Group along with our senior group of consultants has been hosting regional conferences, teaching at universities as adjunct professors, and conducting in-house training on Building Operational Excellence. In September 2005 KLM Technology Group hosted a regional conference titled "Building Operational Excellence in the Hydrocarbon Industry" (title courtesy of Mr. Jeff Gray). Since 2015, Operational Excellence has become the next phase of Lean Manufacturing – you can thank or curse us later.

A generally recognized definition of Operational Excellence is:

Operational Excellence is the execution of the business strategy more consistently and reliably than the competition. Operational Excellence is evidenced by results. Given two companies with the same strategy, the Operationally Excellent company will have lower operational risk, lower operating costs, and increased revenues relative to its competitors, creating value for customers and shareholders. (1)

Some interpretations of this management philosophy are based on earlier continuous improvement methodologies, such as Lean Thinking, Six Sigma, OKAPI and Scientific Management. However, the focus of Operational Excellence goes beyond the traditional event-based model of improvement toward a long-term change in organizational culture. Companies in pursuit of Operational Excellence do two things significantly differently than other companies: they manage their business and operational processes systematically and invest in developing the right culture.

Operational Excellence manifests itself through integrated performance across revenue, cost, and risk. It focuses on meeting customer expectation through the continuous improvement of the operational processes and the culture of the organization. (2)

Big Elephant

Wow, that is a big elephant to swallow and many companies make the mistake of trying to swallow the elephant at once. They spend millions of dollars rolling out the latest buzz word with limited improvements.

There is the old joke, how do you eat an elephant? The answer is one bite at a time. We believe that operational excellence can be broken into reasonable bites that can be managed to bring real results.

Operational Excellence

Operational Excellence has at least five reasonable bites to the many aspects of operational optimization. Five of the most reasonable bites include:

1. Health, Safety and Environment
2. Reliability – Continuity of Operations
3. Quality
4. Cost
5. People Development

1. Health, Safety and Environment (HSE)

HSE is the number one concern. No project or operation can be classified as optimized or excellent unless it is done safely (HSE). There are many benchmark studies that show a strong culture of HSE awareness has economic benefits as well as the social and human benefits. Improving your safety comes with an economic cost, but a direct cost benefit of improving your safety is lower insurance rates and improved corporate branding. Many companies with poor HSE records are no longer in business.

A. The health of your employees and neighbors is especially important. Limiting the exposure of hazardous materials is the key to increasing the health of your team.

B. Safety has at least four parts.

1. Construction Safety: This improves over time with the greatest benefit being a reduction in construction deaths.
2. Industrial Safety: Mostly thought of as PPE, ladder safety, etc.
3. Process Safety Management (PSM) as required by OSHA: Great progress has been made in PSM, but many companies still do not meet the minimum requirements published by OSHA.
4. Risk Management: Due to the large number of annual major incidents across the industry limiting risk is critical. KLM Technology Group's senior consultants have a special focus in risk management to assist in this critical area.

C. We live, work, and play on this earth. Moving forward we need to do a better job of preserving the earth. We are improving and polluting less, but we still have work that we can do to reduce our footprint.

2. Reliability - Continuity of Operations

A stable, reliable plant is the largest revenue source. A reliable high-cost plant will generate more revenue than a low-cost plant with multiple outages. The on-stream factor is a benchmark of reliability. Industry average is 97%, but the top quartile approaches 100%. This three percent increased production is a significant difference in revenue.

Operations Group is the first part of reliability.

A. Best in-class operation procedures need to be developed. Of the operational procedures reviewed by KLM Technology Group most would rate as poor and do not meet OSHA minimum standards or OEI / KLM Technolgy Group best practices. Most operation procedures are not as very comprehensive and many operation procedures reviewed are only a few pages in length.

B. The risk of not developing best in-class operation procedures is poor operator training based on existing substandard procedures. OEI / KLM Technology Group can provide senior consultants to assist with building best in-class operating procedures and then assist with operations training.

C. Verifying operation procedures were followed is key. If you have traveled in an airplane you have most likely heard the term "Arm Doors and Cross Check." The "cross-check" part of this operational procedure is particularly important because it verifies that the doors were armed. There are several ways operating procedures can be verified as followed such as a check list or an independent set of eyes to verify the procedure was followed, like in the airplane.

D. Incorporating any near misses or actual incidents into the operating procedures allows companies to correct errors that were made going forward. Hiding near misses or team management flaws does not fix the issue and prevent future incidents.

Maintenance Group is the second part of operational reliability.

If you survey any group of maintenance managers, they will acknowledge a large percentage of maintenance cost is caused by mis-operation. A way to reduce your maintenance cost is to improve your operations group. Reliability of the Operations Group has a cost, but this cost can be offset by lower maintenance and lower insurance rates.

A. Best in-class maintenance procedures need to be developed. Most of the companies that we have reviewed do not have codified maintenance procedures. Instead they rely on equipment data books as their maintenance procedures. I would rate this as poor and they do not meet OSHA minimum standards or best in class practices.

B. The challenge of not codifying good maintenance procedures it that the maintenance training is based on your existing procedures. Without good procedures one cannot have good technician training.

C. Verifying operation procedures were followed is key. If you have traveled in an airplane you have most likely heard the term "Arm Doors and Cross Check." The "cross-check" part of this operational procedure is particularly important because it verifies that the doors were armed. There are several ways operating procedures can be verified as followed such as a check list or an independent set of eyes to verify the procedure was followed, like in the airplane.

D. Incorporating any near misses or actual incidents into the operating procedures allows companies to correct errors that were made going forward. Hiding near misses or team management flaws does not fix the issue and prevent future incidents.

3. Quality

A company's quality is reflective of external and internal aspects.

A. External aspects: By developing and maintaining the company reputation of producing quality products will allow you to charge a premium during economic up turns and maintain your key customers in a downturn.

B. Internal aspects: There is an added cost of non-quality production. Sometime the product can be reprocessed, with an added energy debit. If the product cannot be reprocessed it will need to be sold with a cost debit.

Most companies have quality audits for the sole purpose of receiving a quality certification. This is certainly a good reason, but a better reason would be to utilize the audit to improve the product quality. Most audit finding are above 90% compliance – amazing – and not true. They are going through the motions of an audit.

If one audited diligently, what would be a reasonable compliance percentage? 75% would probably be a high number, but companies consistently audit above 90%. Many audits are time consuming and unproductive, when in reality they could be made very productive by a rigorous independent audit team.

One of our senior consultants was on a safety audit team. The previous audit team found four non-compliance items. Our team found 40+ non-compliance items that should have been previously identified. The senior consultant thought this was a great audit that made the plant considerably safer. A safety colleague asked how we were going to deal with the political implications from the stricter audit. The senior consultant replied there should be no political implications, everyone should understand that we made the plant considerably safer.

The political implications were that within three months the senior consultant was no longer on the audit team for that company. If your audit teams are not finding compliance items, they are not really looking therefore you are not allowing your plant to have higher quality or safety.

At one safety audit the senior management team instructed our audit team to do a rigorous audit, which is great and will lead to higher safety, and lower incidents. The audit team found many non-compliant issues

At the end of the audit the senior management team then ask the plant that was being audited what they thought of the audit team. Of course, we were rated poorly by the plant where we just found many non-compliance issues. You can instruct a team to audit rigorously, and when they do it is not required to ask the audited plant what they think of the team – you already know the answer to this question.

If your company is experiencing high rates of incidents, your audit team is potentially laboring under the politically correct method resulting in incidents, higher injuries and insurance cost.

4. Cost

Cost control is a particularly important aspect of operational optimization. The two largest costs are feedstock and energy. An exceedingly small feedstock reduction can lead to a very large profit improvement. A feedstock reduction team should be developed to review feedstock utilization.

In 2002 at Titan Petrochemicals in Malaysia, a feedstock reduction team was able to reduce feedstock cost over USD 10.0 million dollars while increasing production. In 2008 at PT Chandra Asri in Indonesia, a feedstock reduction team a was able to reduce feedstock cost USD 10.0 Million and in 2009 feedstock and energy optimization increased plant margin greater than USD 20.0 million, while increasing production rates.

The industry averages three percent energy improvement per year. The top quartile will improve more than three percent. If you are maintaining your energy usage year after year, you are falling behind. OEI / KLM Technology Group can provide senior consultants to review your feedstock and energy utilization. Sometime just the increased focus in feedstock and energy can bring a very large Return on Investment (ROI) from a Process Study.

There is also the timeliness of production. To overproduce and store finished or intermediate products many are not the best use of capital. A supply chain plan can provide cost savings.

5. People Development

Most people might rate this higher than fifth. It is an especially important aspect of operational excellent, but talent can be acquired for a price. The best plan is to hire talented people, train them well, pay them well, and retain them, but few companies seem to be capable of accomplishing this task. People development will ensure that items one through four are optimized.

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How to... Float Valve Tray | Part 3

How to design and optimize Float Valve Trays

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 3rd part of a series on different kinds of trays and internals.

Float Valve Trays (also called moveable valve trays or ballast trays) are the most flexible tray type among the standard trays. They have been used for about 80 years in technical applications, they are well studied and they are still the mainly used tray type in towers.

On a distillation tray vapor enters liquid and forms a two-phase regime (bubbling, froth, spray). The tray types differ mainly in the way the vapor enters the liquid.

For float valve trays, at low gas loads the valves are all closed – the gas can only enter the liquid layer by the annular gap of the valve plate's initial lift (i.e. dimples that prevent the valve plate from sticking to the tray panel). Therefore float valve trays do not tend to weep.

At increasing gas loads the float valves start to open and the gas enters the liquid layer pre- dominantly in horizontal direction (resulting in less entrainment compared to sieve holes). The starting point for the movement of the valves is called "Closed Balance Point (CBP)". At the "Open Balance Point (OBP)" all valves are at their maximum opening (i.e. the valve plate has reached its maximum lift). The region between the CBP and OBP is called "working area", where the pressure drop is quite constant.

By increasing the gas load beyond the OBP the pressure drop characteristic behaves like a static tray: the pressure drop is proportional to the gas velocity square.

The tray spacing of float valve trays can be small (300 mm), but is normally – due to inspection and maintenance reasons – about 450 - 500 mm.

The advantage of the flexibility of this tray type is achieved in exchange for the disadvantage of moveable parts within the tower. The movement includes the risk of wear of the valves in operation, getting lost in high pressure cleaning, higher effort in maintenance and higher costs in fabrication of the trays. (The acquisition costs of float valve trays are about twice of those ofsieve trays.)

There are various float valve types:

The most common type is the so called **V1-valve** (see Fig. 1). It is a round valve plate (Ø48 mm) with three legs fitting in a Ø39 mm panel hole.

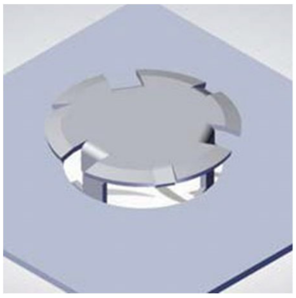


Fig. 1: Standard Round Float Valve

The standard pitch of these valves is 76mm (in flow direction) x 127 mm. The resulting relative free area is about 13%.

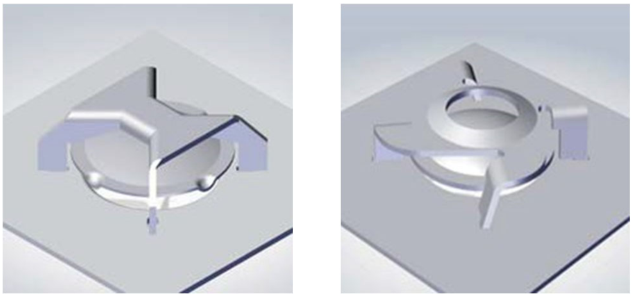
As the valve elements interact to create a proper two-phase layer, the standard pitch should only be varied to a small extent.

The standard V1 valve has a material thickness of 1.5mm (weight 24g/valve) and dimples for initial lift (to prevent it from sticking to the tray panel).

In corrosive applications and for increasing the operation range of a tray (see later) the material thickness can be 2.0mm (weight 30g/valve) as well.

To achieve even higher valve weights one can add ballast plates to the valve plate.

The legs of the V1 valve guide the element in the panel opening and limit the lift of the valve ("legged valve"). Another float valve type is called **Caged valve** (see Fig. 2). These valves consist of two parts: The (moving) valve plate and the (static) cage. This type of float valve is used in fouling services (higher turbulence caused by the cage) and to minimize pressure drop (valve plate lighter than legged type). The panel hole diameter is Ø39mm.



Type T **Type A3**

Fig. 2: Caged Float Valves

Beside these standard types there are several other types:

Rectangular float valves (see Fig. 3) have two legs in flow direction. The values for panel and curtain area of the type "BDH" is very similar to the V1 valve data. The "BDP" type has twice the length of the BDH.

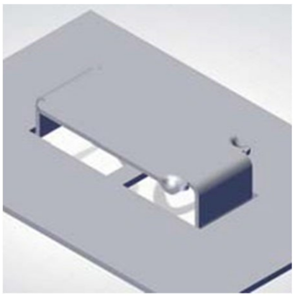


Fig. 3: Rectangular Float Valve

Venturi shaped panel openings (see Fig. 4) are fabricated to achieve minimum pressure drop. The length of the legs has to compensate this additional "thickness" of the tray panel. The venturi height is normally about 6.7mm. The legged valves are often called "V4", the caged units "A4" or "T0" (depending on the cage type).

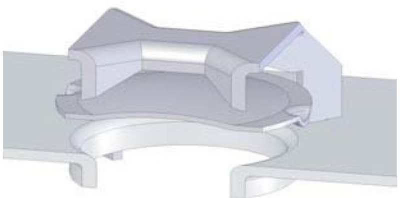


Fig. 4: Venturi Deck Opening

Mini valves (see Fig. 5) have a panel hole diameter of about Ø24mm and normally two legs orientated towards inlet and outlet (like the rectangular valves). Their size helps to achieve a small pressure drop (same at sieve trays: The pressure drop of the identical open area of small holes is better than that of large holes) and a good coverage of the active area (easier to achieve by small elements than by larger ones). These advantages are in competition with higher production costs.

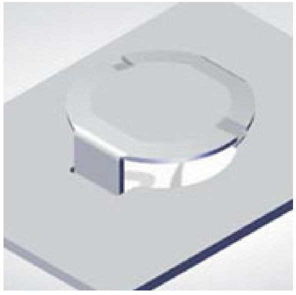


Fig. 5: Mini Valve

Moving elements cause abrasion and wear. To prevent spinning of the valves, the panel holes are equipped with **anti-spin noses**. This helps to reduce the risk of loosing valves by abrasion of the legs or of enlarging the panel holes by rotation. Nevertheless, you will find sticking valves, where the legs have worked their way into the panel deck.

Double disk valves (see Fig. 6) have two valve plates in their cage. The upper plate (called "ballast plate") has three small legs for static lift. The lower plate is called "orifice cover" and closes the deck hole.

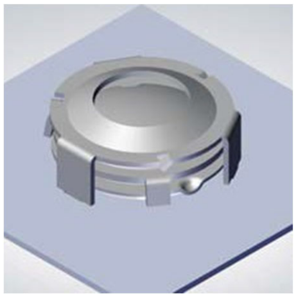


Fig. 6: Double Disk Float Valve

All valves are normally installed in the workshop of the supplier. The elements are inserted into the panel openings, the panel is turned and the valves are locked from the opposite side. In case of maintenance within a tower, you need one worker above and one below the panel to put in new valve element. Because this is quite ex- pensive (sometimes you have to add thousands of valves), repair valves have been developed (see Fig. 7). They can be inserted from the top side of the panel by one worker.

Fig. 7: Repair Valve

(Caution: There are so called "SnapIn-Valves". These are not suitable for substitution of V1 valves as they are made for openings of Ø40mm.)

A special type of caged valve is the **Varioflex valve** (see Fig. 8). Its valve plate has (as standard) a round hole (Ø20mm).

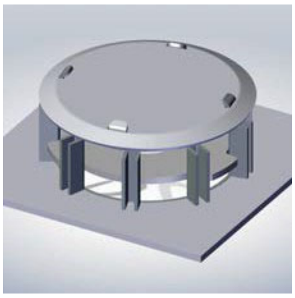


Fig. 8: Varioflex Valve

Apart from the presented float valve types, there are still several others types (e.g. cup- valves, double disk legged valves or legged valves with one shorthend leg for acting as push valve at high gas loads).

Pressure Drop Characteristic

Below the CBP the gas enters the tray by gaps or by opening single valves. At this load the tray is not safe in operation, because the liquid may use lanes to cross the tray – without getting in contact with the gas. To enlarge the operation range to low gas flow rates, you can equip the tray with light and heavy valves in alternating row blocks (parallel to weir). This design is called "multi-weight". In Fig. 9 the pressure drop characteristic of a single and a multi-weight design is shown.

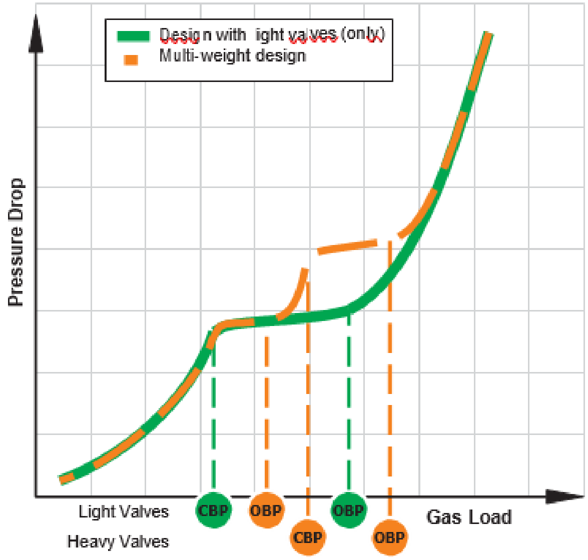


Fig. 9: Pressure Drop Characteristic

The *Operating Area* of a float valve tray is de- fined by different limits. In Fig. 10, a qualitative operation diagram is shown. Please note, that the position and shape of all curves depend on the physical data, the tray and cap geometry and the gas/ liquid load. Each curve can be limiting!

The GLITSCH bulletin can be considered as the standard calculation procedure for float valve trays. As it deals only with some types of valves, there have been developed new models for calculation.

The *Operation Point* (Op in Fig. 10) of the design case (as well as the minimum and maximum load) has to stay inside all limiting curves. The design load case should additionally be above the OBP (not only the CBP): The movement of valves should not take place in the design load case as the abrasion is too high!

For stable operation and good efficiency there is a useful operation area with narrower limits (e.g. 80%-FFCF and 85%-FFJF curves).

The first step in analyzing a design is – of course

– calculating all relevant parameters. For a float valve tray design there are 10 main parameters shown as curves in Fig. 10. These parameters are discussed in this article. There are some additional effects you will have to look at: entrainment, head loss at downcomer exit (clearance), flow regime, downcomer residence time, efficiency, sealing, construction issues, statics, …

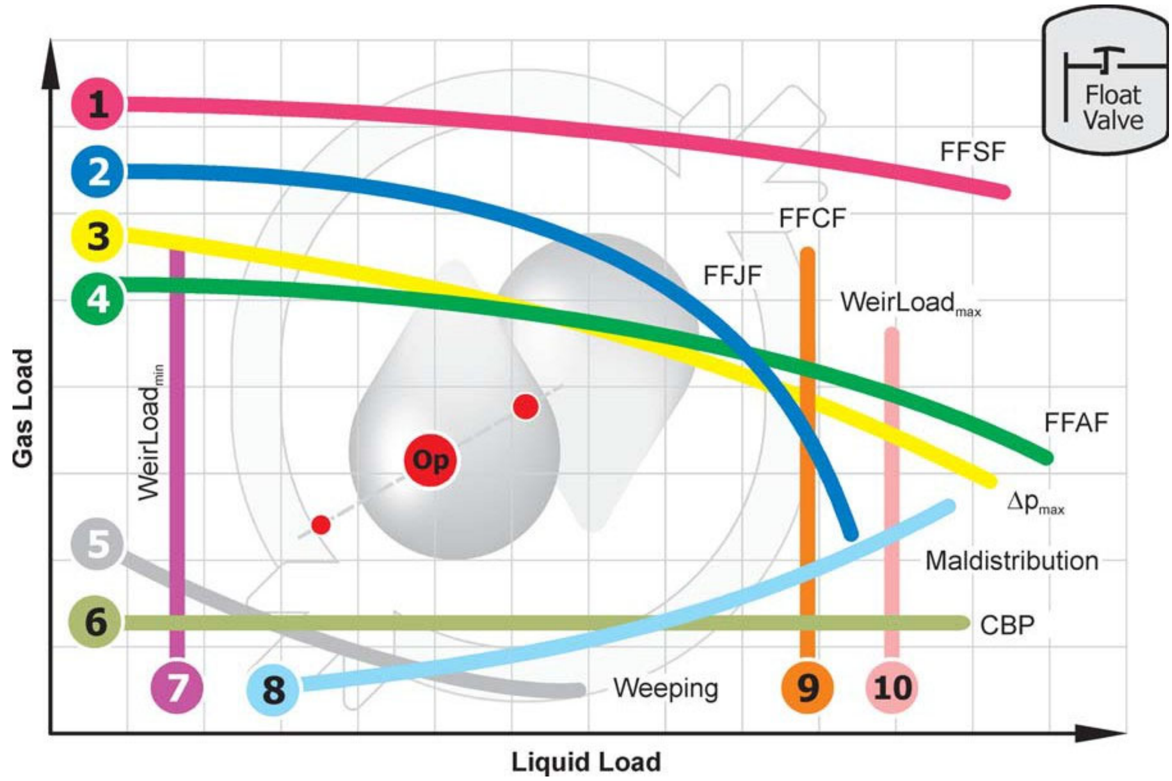


Fig. 10: Qualitative Operation Diagram for Float Valve Trays

Please note, that all free suppliers' software only show a limited number of these parame- ters and therefore are not save to use for de- sign, rating and troubleshooting of trays. For safe design you should be able to calculate all parameters! (e.g. software TRAYHEART OF WELCHEM)

In the following sections, all 10 main parameter curves of Fig. 10 are described. Each suggested ac- tion for preventing a certain effect may result in fertilizing another. The main task for designing trays is to balance these different and contra- dicting effects.

1 System Flood FFSF

There is a system limit set by the superficial vapor velocity in the tower. When the vapor velocity ex- ceeds the settling velocity of liquid droplets („Stokes Law Criterion"), vapor lifts and takes much of the liquid with it. A well known model was published by STUPIN AND KISTER 2003.
This flooding effect cannot be reduced by use of other tray types or by increasing tray spacing.
The only way is to enlarge the vapor cross section area (e.g. enlarging tower diameter or re- duce downcomer area).

Jet Flood 2

There are several definitions in literature for the so-called Jet Flood. Similar definitions are Entrain- ment Flood, Massive Entrainment, Two- Phase Flood or Priming. For practical under- standing, Jet Flood describes any liquid carried to the tray above by the gas stream. This leads to a shortcut recycling of the liquid with loss of tray efficiency, additional pressure drop and additional down- comer load. For good tray performance, the Jet Flood value should be less than 75-80%.

You can reduce Jet Flood by

a. lowering the gas velocity (higher open area,
b. i.e. more valve elements)
c. enlarging the tray spacing
d. lowering the froth height on the tray deck (by reducing weir height or weir crest height)
e. enlarging the active area (i.e. the gas flow ar- ea) by sloping the downcomers

3 Pressure Drop

In most cases there is specified a maximum allow- able pressure drop of the tower. You have to en- sure that the pressure drop per tray does not exceed a certain value. This leads to a limiting curve within the operation diagram.

To reduce the pressure drop of a design, you can

a. lower the gas velocity by enlarging the number of valve elements. As you shouldn't vary the pitch, you have to optimize the panel dimensions to achieve the maximum number of units

b. use venturi openings in the tray deck panels

c. lower the froth height on the tray deck (by reducing weir height or weir crest height)

d. enlarge the active area (with place for more valve units) by reducing the downcomer area or sloping the downcomers

4 Aerated Downcomer Backup FFAF

This limiting effect is also known as Downcomer Backup Flood. It describes the (aerated) backup of the downcomer due to pressure drop effects. It is important to not mix this up with the

Choke-Flood-effects (ref. to 9).

The level of the liquid in the downcomer is the result of (i) head loss at the clearance, (ii) the liquid height on the outlet deck, (iii) an inlet weir (if present) and (iv) the pressure drop of the tray itself. All these effects can be expressed by "hot liquid height". This resulting level in the downcomer has to compensate these effects! Taking into account the aeration of the liquid in the downcomer, the level has to be less than tray spacing plus weir height.

To reduce a high Aerated Downcomer Backup value you have to

a. reduce the pressure drop of the tray (ref. to 3)

b. reduce the head loss of the clearance (use higher clearance height or radius lips or recessed seal pans in case of insufficient sealing)

c. avoid inlet weirs

Please note, that it is no option to enlarge the downcomer area to reduce this flooding effect!

Weeping 5

Weeping is a minor subject of float valve trays. If you have a very low MIN load, you have to ensure, that weeping is minimized. Therefore you can

a. reduce the number of valve elements

b. use valve plates without initial lift (only in clean and non-corrosive services). As weeping occurs normally below CBP the heavy units of multi-weight designs are often build without initial lift dimple.

Closed Balance Point (CBP) 6

Below the Closed Balance Point the operation of the tray is not safe. The liquid is not getting in contact with the gas.

To lower the CBP you can reduce the number of valve elements.

Note: If you are running the tray below the OBP (not all valves completely open), you should consider using a multi-weight valve design to ensure that there are "bubbled areas" in the liquid flow path.

Minimum Weir Load 7

The uniform thickness of the two-phase layer is essential for the successful operation of a tray. To achieve this uniform flow, the tray panels have to be in level and the outlet weir has to be installed accurately.

To compensate small tolerances, the weir crest should be higher than 3mm and the weir load more than 9 $m^3/m/h$. In case of low weir loads you will normally have to consider gasketing the tray to avoid any leakage and loss of liquid.

To ensure these minimum values, you can use

a. notched weirs

b. blocked weirs

Gas Maldistribution 8

In all types of trays the liquid must have a driving force to flow from the inlet to the outlet. As long as there is no gas driven flow, the hydraulic gradient is the main reason for liquid flow.

Because the valve units are obstacles in the liquid flow pass, the hydraulic gradient has to be considered for valve trays, too.

Why might the hydraulic gradient be a problem? At a high hydraulic gradient, the tray will not work properly (see Fig. 11): At the tray inlet the liquid "closes" the valves. The gas will use less liquid affected valves for passage. This leads to a gas maldistribution and a bad efficiency of the tray. Furthermore, if the liquid head of rows with high gradient gets too high, weeping occurs!

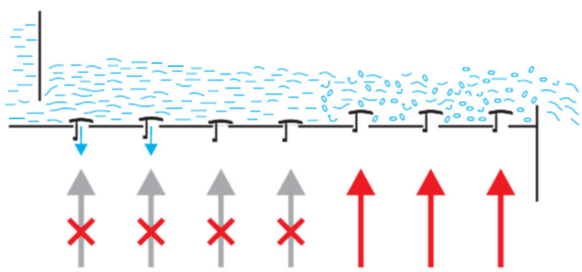


Fig. 11: Gas Maldistribution

To reduce gas maldistribution you have to

a. reduce the number of valve rows (e.g. by switching to a design with more flow passes)

b. cascade the active area

At a high downcomer liquid outlet velocity there is the risk of "undermining" the first valve rows: The valve plates are acting as a baffle guiding the liquid directly to the next tray. To avoid this short cut one can place so-called interrupter bars (see Fig. 12; height about 13mm). Do not confuse these bars with inlet weirs!

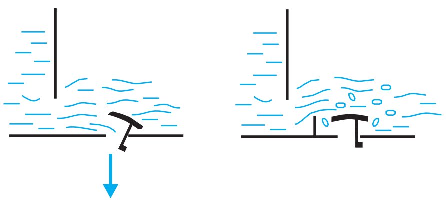


Fig. 12: Interrupter Bar

Choke Flood FFCF 9

The maximum liquid throughput of a down- comer is limited by the liquid velocity and the effect of overload (so-called Choke Flood). The maximum allowable liquid velocity in the down- comer depends on the density ratio of gas to liquid, the tray spacing and the system factor. (The system factor describes the difficulty of phase separation. For common applications it is 1.0.) The most popular downcomer choke flooding calculation was published by GLITSCH 1993.

Another effect of Choke Flood at center and off-center downcomers is initiated by the mutual interference of the two liquid flows into the downcomer.

To prevent downcomer Choke Flood you have to

a. enlarge the downcomer area

b. implement more flow passes (with in sum an overall higher downcomer area)

c. enlarge the tray spacing (if limiting)

d. install anti-jump baffles for center / off-center downcomers

Maximum Weir Load 10

The maximum liquid flow handled by a down- comer can also be limited by the weir.

If the weir crest exceeds 37mm or the weir load 120 $m^3/m/h$, the liquid will not enter the downcomer properly.

To prevent overload of the weir, you have to extend the weir length by

a. larger downcomers with longer weirs (or multichordal downcomers)

b. more flow passes

c. swept back weirs at the side downcomers

Conclusion

There are multiple limiting effects that have to be considered at the design and operation of float valve trays. The float valves are still the working horses of the contact elements. Float valve trays are very flexible and their efficiency is constant over a broad load range. Due to the moving parts they are higher in costs (fabrication as well as maintenance).

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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Heat Transfer in Batch Reactors

What is the best way to transfer heat in batch reactor? Is it a simple jacket, coil, half pipe or dimple

Abdullah Al Bin Saad, Process Engineer

Background

Heating and cooling systems are used in the chemical industry to, among other things, control reaction temperatures, making these systems critical for reactor product quality and safety. From that standpoint, this article will highlight some design aspects of heat transfer methods in batch tank reactors, with the same principles potentially applicable to continuous tank reactors.

A reactor tends to be the heart of a chemical process. An industrial reactor is a complex device where heat transfer and mass transfer may happen along with the chemical reactions. Reactors can be classified into two main types based on their mode of operation:

1. Continuous Reactor

Raw material and product flow continuously through the duration of the process.

2. Batch Reactor

Raw material is loaded into the reactor at the start of process and product is removed sometime later.

Alternately, some batch reactors may operate semi-continuously with one or more of the raw materials metered into the batch during the reaction time.

This article focuses on batch reactors, although the principles can be applied to continuous and semi-continuous reaction systems that use a similar configuration. Batch reactors are found in various industry areas such as chemical, pharmaceutical, as well as food. One of the benefits of the batch reactor is that it can be used to carry out sequence operations that would result in producing different products or require product traceability or frequent cleaning.

A typical batch reactor may consist of a vessel that is equipped with an agitator and an integral or external heating and cooling system. Batch reactors may well differ in size and materials. For instance, a glass-lined reactor can be used when highly corrosive materials such as strong acids are present.

Heat Transfer Methods

Chemical reactions contained by batch reactors typically either liberate heat (exothermic) or absorb heat (endothermic). Hence, heat transfer plays a major role in reactor performance.

Heating is needed for endothermic reactions, for vaporizing liquids during evaporation steps, and for bringing a vessel up to the desired operating temperature. Cooling is required for exothermic reactions, condensing vapors, and to bring a vessel down to the desired operating temperature. Heat is normally transferred to the reactor through a jacket, internal coil or external heat exchanger. Both coil and jacket are limited by the physical dimension of the reactor. Hence, where coil or jacket cannot provide the required surface area, such as in case of a highly exothermic reaction, a recirculation loop with external heat exchanger can be used; the external heat exchanger is noted but not discussed further since this article focuses on coil or jacket configurations. However, it is noted that the external loop adds to the batch reactor volume, which must be considered in the design.

Jacket Types

Jackets typically come in three shapes - simple (conventional), dimpled or half pipe jackets. Other jacket types may be used, such as clamp-on types, but are not discussed in this article.

The simple (conventional) jacket

It is an open jacket with an annular space holding the heat transfer media on the external surface of the reactor. It has several features such as simplicity in construction and full coverage for the reactor. (See Figure 1)

Figure 1- Simple Jacket (Provided by Pfaudler Group)

Conventional Jackets could be classified into two main classes, either baffled or non- baffled.

A simple jacket that has no internal baffles is usually inefficient for heat transfer for fluids without a phase change, such as hot oil and cooling water, due to the low velocity and associated low external heat transfer film coefficient. However, the heat transfer coefficient for fluids with a phase change such as steam essentially do not depend on velocity.

Additionally, sometimes the overall heat transfer resistance is already limited due to the vessel wall thickness or a low internal heat transfer film coefficient for the reaction materials, in which case the external heat transfer coefficient may not have much influence on the overall heat transfer coefficient.

As the conventional jackets naturally suffer from low heat transfer rates in case of non-isothermal fluid, the remedy will be adding a spiral baffle in the jacket. Spiral baffles induce turbulence and increase the heat transfer coefficient and eventually heat transfer rate.

The dimple Jacket

Dimple jacket is a thin exterior shell which is attached to the reactor shell with spot welds located in a regular pattern. (See figure 2)

Figure2- Dimple Jacket

The dimple jacket design provides a large heating or cooling transfer area up to 300 psig (20 barg). The cost of the dimple jacket is comparatively inexpensive. As a result of turbulence in dimple jacket, the pressure drop is around 10 to 12 times than the simple jacket meaning the velocity should be limited to 0.6 m/s (2 ft/s).

As a general engineering practice, the dimple jacket will be more economical than other choices when the internal pressure of the reactor is less than 1. 7 times the jacket pressure. Due to design constraints, dimple jackets are not applied to small reactors (less than 10 gallon).

The dimple jackets are used primarily if the heat-transfer medium is a liquid. To ensure efficient use of dimple jackets the heat-transfer medium should be clean and not contain solids. As any solid carried in the liquid can choke up the small recesses of the jacket.

Although the dimple jackets are more efficient than simple jackets in terms of heat transfer rate and heat distribution, their use is not recommended for thermal cycling as its tin wall can crack easily.

The half pipe jacket

A half pipe jacket comprises of a welded half pipe that furls around the external of the reactor, creating a circular path for the heat transfer fluid to pass through. (See Figure 3)

Figure3- Half Pipe Jacket

One of the advantages of using a half-pipe (or dimple jacket) is that the thickness of the Reactor vessel can be lowered, and this is very helpful in higher pressure applications especially when using higher cost Alloys for the Reactor Body.

The disadvantage of the half-pipe is that the heating surface is reduced by about 15 to 20% compared with the simple jacket. A second possible issue is the stresses caused on the weld joints if there is excessive thermal cycling of the Reactor. This can lead to stress damage and leakage of the half-pipe.

The cost is generally higher than the dimple and a simple jacket at lower jacket design pressures. Conversely, it may be cheaper for high pressure on the service side. However, a limited surface area can be applied to the reactor as the larger number of welds can lead to mechanical concerns.

Jacket Type selection

The jacket type can be based on some general guidelines as following, although selection can also be influenced by other parameters such as metallurgy requirements or heat transfer media type:

A) For Reactor volume ≤ 500 gal (Use simple jacket)

B) For Reactor volume ≥ 500 gal (Use the dimple or half pipe jacket based on pressure)

C) Jacket Pressure ≤ 300 psi (use the dimple jacket)

D) Jacket Pressure >300 psi (use the half pipe jacket)

E) If reactor pressure is greater than twice the jacket pressure, use the simple jacket.

These guidelines for selection of jacket type are taken from (reference 1).

Internal coil

Internal coils for agitated vessels are generally full helical coils to provide additional surface area, especially for a highly exothermic reaction. However, coil cleaning can be an issue in industries that need extreme sanitation and/or food safety protocols. (See Figure 4)

Internal coils do not provide the degree of safety from contamination by the heat transfer medium that jackets can offer.

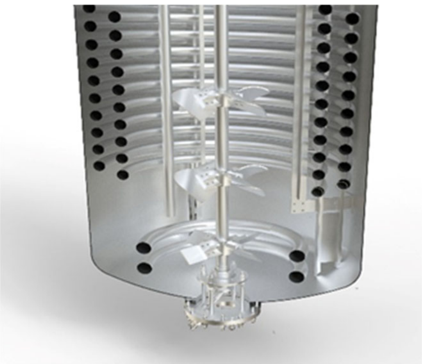

Figure4- Internal coil (Provided by Pfaudler Group)

The choice between jacket and coil is based on several factors such as:

The jacket is preferred when the process fluid is highly reactive or corrosive. Since it is not in direct contact with the process fluid, and the jacket provides the degree of safety from contamination by the heat transfer medium. For instance, the jacket is used in a glass-lined reactor where highly corrosive materials are used in the reaction such as strong acids.

The coil has a benefit that larger surface areas can be provided such as in case of the highly exothermic reaction.

Jackets generally provide a more even heat transfer distribution.

Simple jackets (conventional) have a lower utility velocity that can increase the fouling rate.

Coils provide quicker and more aggressive heating and cooling.

Cost Comparison

The typical cost comparison presented below is provided by Pfaudler group which is the world-leading process solutions company.

In general, the cost difference, based on the design of a medium pressure Reactor, will be

- Half-pipe Jacketed Reactor – 100
- Dimple (Pillow plates) jacket – 108 to 112
- Conventional jacket – 115 to 125

As pressure increases, the conventional jacket becomes even more expensive, while for low pressure the conventional jacket will be more cost effective.

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Acknowledgment

Iam grateful to Pfaudler Group (world-leading process solutions company) for providing some pictures and sharing some practical information about batch reactors .

About the Author

Abdullah Hussain holds a bachelor's degree in chemical engineering from King Fahd University of Petroleum & Minerals with honor. His professional experience of over seven years covers various design stages including conceptual studies, feasibility, front-end engineering and detail design.

Load Sharing for Parallel Operation of Gas Compressors

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

The art of load sharing between centrifugal compressors consists of maintaining equal throughput through multiple parallel compressors. These compressors consist of a common suction and discharge header. Programmable logic controllers (PLCs) can be incorporated with load sharing functions or can be incorporated as standalone controllers also. Control signals from shared process parameters such as suction header pressure or discharge header pressure can be then fed to individual controllers such as compressor speed controllers (SC) or anti-surge controllers (UIC) to ensure the overall load is distributed efficiently between the compressors.

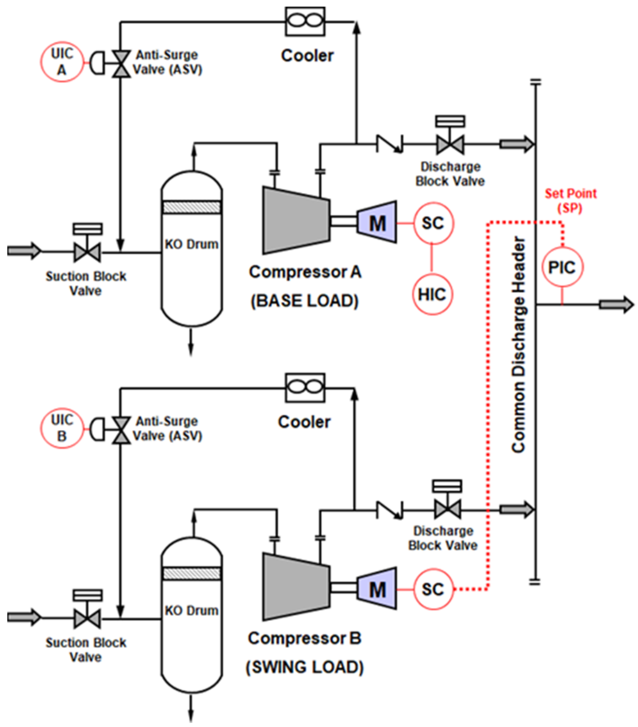

Figure 1. Base Load Operation Method

The following article covers load sharing schemes for parallel centrifugal compressor operation.

Load Sharing Options

The load sharing options covered are as,

1. Base Load Method
2. Suction Header - Speed Control Method
3. Equal Flow Balance Method
4. Equidistant to Surge Line Method

Base Load Method

In Base Load method of operation, one compressor is allowed to run on manual mode while the other is controlled through speed manipulation based on the discharge header pressure. The pressure controller on the discharge header is termed as the Master Pressure Controller (MPC) that alters the second compressor's speed a.k.a "Swings" the compressor speed to cater to varying throughputs. In Fig 1, the speed of compressor A is manually set (HIC) for a maximum throughput, i.e. Base Load.

The speed of compressor B is altered based on the master pressure controller (PIC) set point (SP) to attend to the swing in flow throughputs.

During periods of low process demand, Compressor B (swing machine) can be recycling & sometimes even close enough to the Surge Control Line (SCL) causing the swing machine to trip. Additionally, due to differences in piping layouts &

pressure loss, the compressor operation would not be symmetrical, causing operators to frequently intervene. With these limitations, the base load method is least preferred.

Suction Header - Speed Control Method

In the suction header - speed control method, no base load exists. Instead the master pressure controller (PIC) is shifted to the suction header. The advantage offered is, both compressors operate independently despite a common set point provided by PIC to the speed controllers (SC) of both compressors.

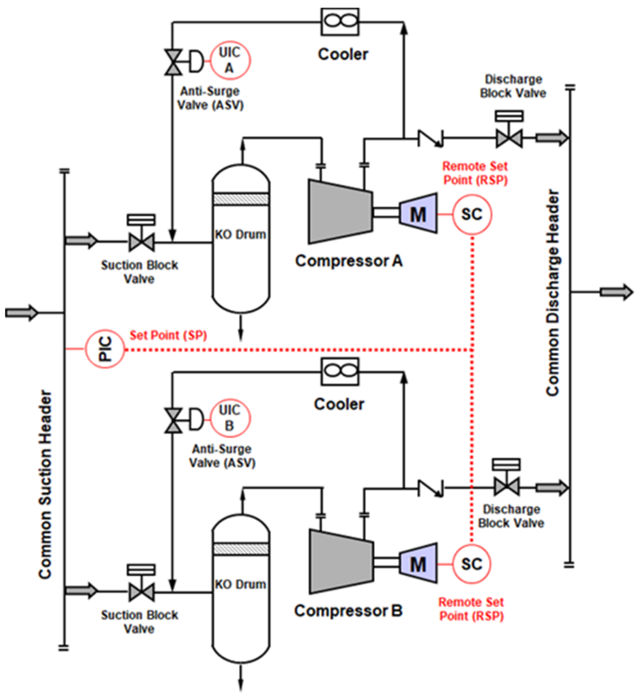

Figure 2. Suction Header Speed Control Method

It may be noted that both compressors would not necessarily be running at the same speed or flow due to differences in the piping layout as well as during a compressor recycle since both anti-surge controllers (UIC A/B) also act independently of each other.

To ensure no production losses, the configuration consists of standby machine along with working compressors. During the failure of one of the compressor, say machine A, the PIC issues a signal to increase the speed of compressor B, until the standby compressor can be brought online to maintain throughput. In case of layouts that have no standby compressors, a 2 ´ 50% configuration, with no recycle during regular operation must be chosen. This enables the remaining working compressor to cater to 100% of the throughput/load at higher speeds during failure of the one of the compressors.

Equal Flow Balance Method

In the equal flow balance method, the Master Pressure Controller (PIC) on the common discharge header determines the total load demand and alters the speeds of Compressors A & B via SC. The individual flow control signal to each speed controller is achieved by scaling the total load demand (BIAS A & BIAS B) to the individual flow controller (FC) on each compressor. Both Compressor operations are independent of the Anti-surge valve (ASV) operation.

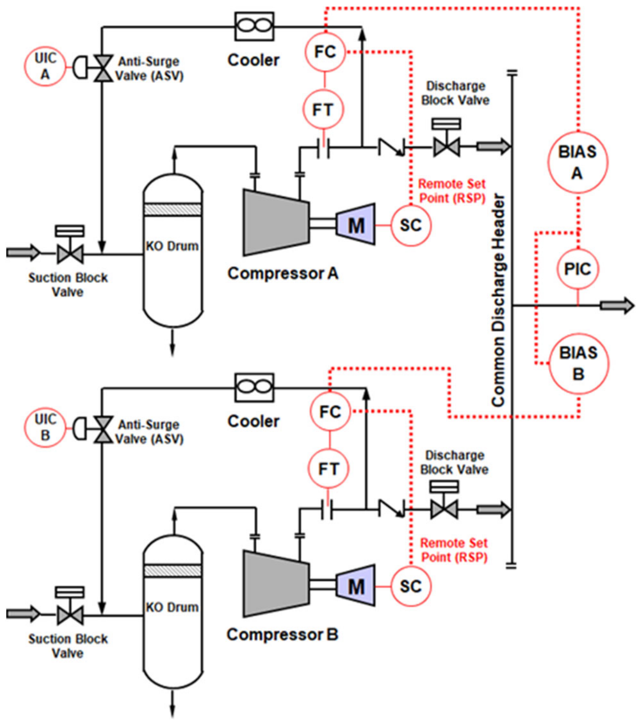

Figure 3. Suction Header Speed Control Method

However certain limitations exist with the flow balancing method. Due to additional control elements, CAPEX cost increases. Furthermore since the flow element & transmitter (FT) is installed on the compressor discharge, additional pressure drop occurs which represents energy losses.

For the cascaded control used, PIC Þ FC Þ SC, the inner loop (FC) must respond faster than the PIC outer loop. This causes the master pressure control, PIC to be sluggish. A faster FC loop also means, the compressor speed would increase rapidly than required often reaching maximum speed. Hence this does not offer the best control strategy.

Equidistant to Surge Line Method

In the equidistant method, the aim is to ensure, the deviation/distance between the operating point and the surge control line (SCL) in both trains is equidistant.

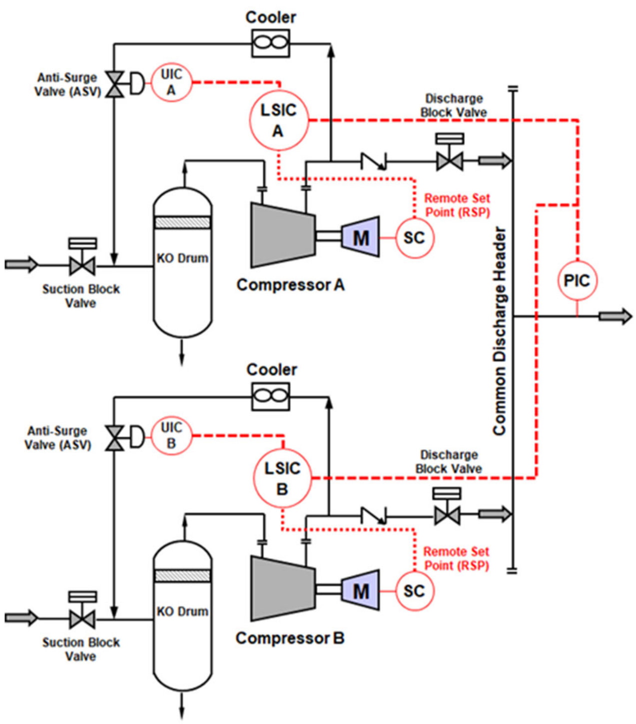

Figure 4. Equidistant to Surge Line Method

In this configuration, neither the throughputs through each compressor or the operating compressor speed is the same, but only the deviation between the operating point & SCL. It may also be noted that the load sharing function (LSIC A/B) that alters the compressor speed, is not fed with the signal from the suction flow transmitter (FT), but instead the anti-surge controller (UIC A/B) and the master pressure controller (PIC) installed on the common discharge header. This would mean, both UIC A/B and LSIC A/B have to coordinate in real time.

A significant advantage of the equidistant to surge line method is the configuration's ability to cater to asymmetrical performance curves, i.e., dissimilar compressors. In brownfield modifications, any addition of new compressors can offer synchronicity issues including variation in throughputs & pressures due to differences in performance curves & piping layouts. Therefore the equidistant method becomes an effective configuration for varying loads ensuring both compressors independently adjust their respective operations and avoid surge.

Some Design Considerations

1. The Master pressure controller which provides shared information across all compressors can often be subjected to harsh field conditions. To circumvent these issues, redundancy with multiple transmitters can be provided. This ensures not only maximum availability but also hardwiring the transmitters prevents any loss of signals to the Load sharing system.

2. Depending on the reliability of the control systems, controllers need to be replaced

sometimes with third party OEM vendors, each with their own proprietary control systems. Hence load sharing systems must be able to integrate different vendors.

3. Real Time optimization (RTO) techniques based on regression models of steady state data have gained sufficient footing in recent years. Short Time RTO of the order of a few minutes & Long term RTO of the order of a few days can be employed to determine the best load distributions between compressors.

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About the Author

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 13 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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The Future of the Downstream Industry Refining and Petrochemical Processes Integration

Marcio Wagner

Introduction and Context

The current scenario present great challenges to the crude oil refining industry, prices volatility of raw material, pressure from society to reduce environmental impacts and refining margins increasingly lower. The newest threat to refiners is the reduction of the consumer market, in the last years became common, news about countries that intend to reduce or ban the production of vehicles powered by fossil fuels in the middle term, mainly in the European market.

Facing these challenges, search for alternatives that ensure survival and sustainability of the refining industry became Constant by refiners and technology developers. Due to his similarities, better integration between refining and petrochemical production processes appears as an attractive alternative.

Available Synergies between Refining and Petrochemical Sectors

The petrochemical industry has been growing at considerably higher rates when compared with the transportation fuels market in the last years, additionally, represent a most nobler destiny and less environmental aggressive to crude oil derivatives. The technological bases of the refining and petrochemical industries are similar which lead to possibilities of synergies capable to reduce operational costs and add value to derivatives produced in the refineries.

Figure 1 presents a block diagram that shows some integration possibilities between refining processes and the petrochemical industry.

Process streams considered with low added value to refiners like fuel gas (C2) are attractive raw

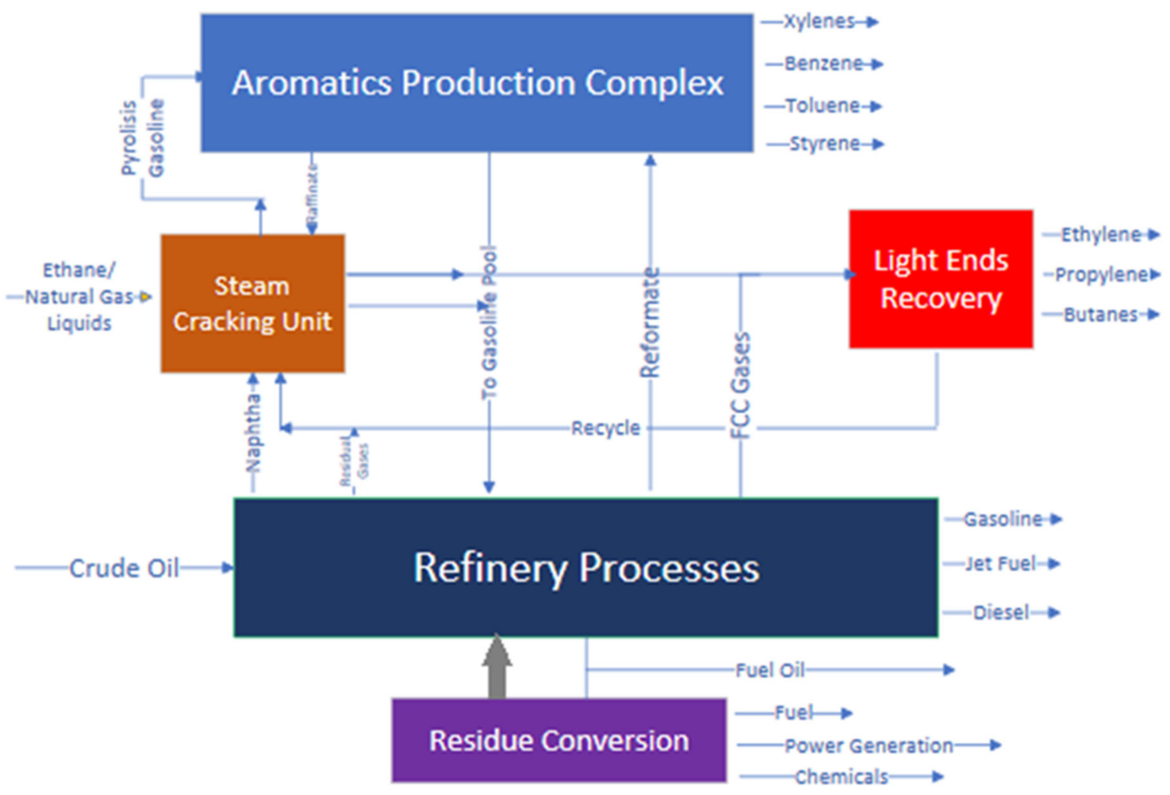

Figure 1 – Synergies Possible between Refining and Petrochemical Processes

materials to the petrochemical industry, as well as streams considered residual to petrochemical industries (butanes, pyrolisis gasoline, and heavy aromatics) can be applied to refiners to produce high quality transportation fuels, this can help the refining industry meet the environmental and quality regulations to derivatives.

The integration potential and the synergy among the processes rely on the refining scheme adopted by the refinery and the consumer market, process units as Fluid Catalytic Cracking (FCC) and Catalytic Reforming can be optimized to produce petrochemical intermediates to the detriment of streams that will be incorporated to fuels pool. In the case of FCC, installation of units dedicated to produce petrochemical intermediates, called petrochemical FCC, aims to reduce to the minimum the generation of streams to produce transportation fuels, however, the capital investment is high once the severity of the process requires the use of material with noblest metallurgical characteristics.

How to Improve the Yield of Petrochemicals in the Refining Hardware?

An example of FCC technology developed to maximize the production of petrochemical intermediates is the RxPRO™ process by UOP Company, this process combines a petrochemical FCC and separation processes optimized to produce raw materials to the petrochemical process plants, as presented in Figure 2. Other available technologies are the HS-FCC™ process commercialized by Axens Company, and INDMAX™ process licensed by McDermott Company.

Figure 1 indicates the integration between process streams, however, operational costs can be considerably reduced through the integration of utilities as steam, water, hydrogen, etc. Normally, refineries have low availability of hydrogen while petrochemical plants can export this utility, on the other hand, petrochemical processes have high demand by electric and steam Power that can be supplied by refineries in an integrated process.

Bottom barrel process units as Delayed Coking can be quite versatile in this scenario, petroleum coke gasification can produce syngas which in turn can be used as raw material to produce high-demanded chemicals, like ammonia, methanol, sulfuric acid, dimethyl ether, etc. The use of IGCC power generation plants can ensure energy supply to the refining and petrochemical processes, an example of this technology is the FLEXICOKING™ process, developed by ExxonMobil Company.

Some process technologies were developed aim to produce petrochemical intermediates from streams considered secondary to refiners, a good example is the AROMATIZATION™ process developed by GTC Company to produce aromatics from olefins (C6-C8) which are produced in FCC units. The LCO-X™ process developed by UOP Company is capable to convert Light Cycle Oil (LCO) produced in FCC units in aromatics with high added value.

Light paraffin dehydrogenation processes also allow a better integration among refineries and petrochemical plants once convert residual streams (fuel gas) into petrochemical intermediates with high added value.

One of the available dehydrogenation technologies is the CATOFIN™ process, commercialized by McDermott Company, as presented in Figure 3.

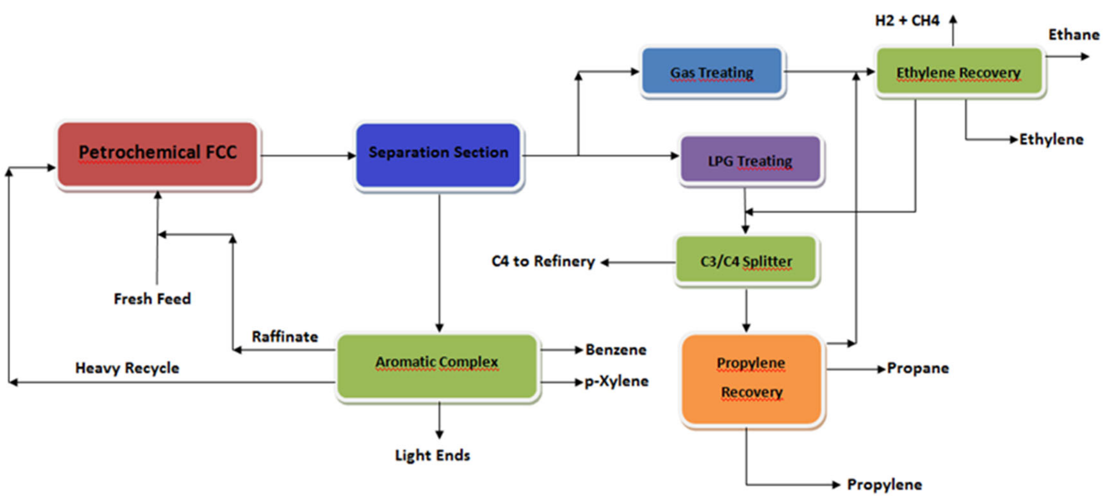

Figure 2 – RxPRO™ Process Technology by UOP Company.

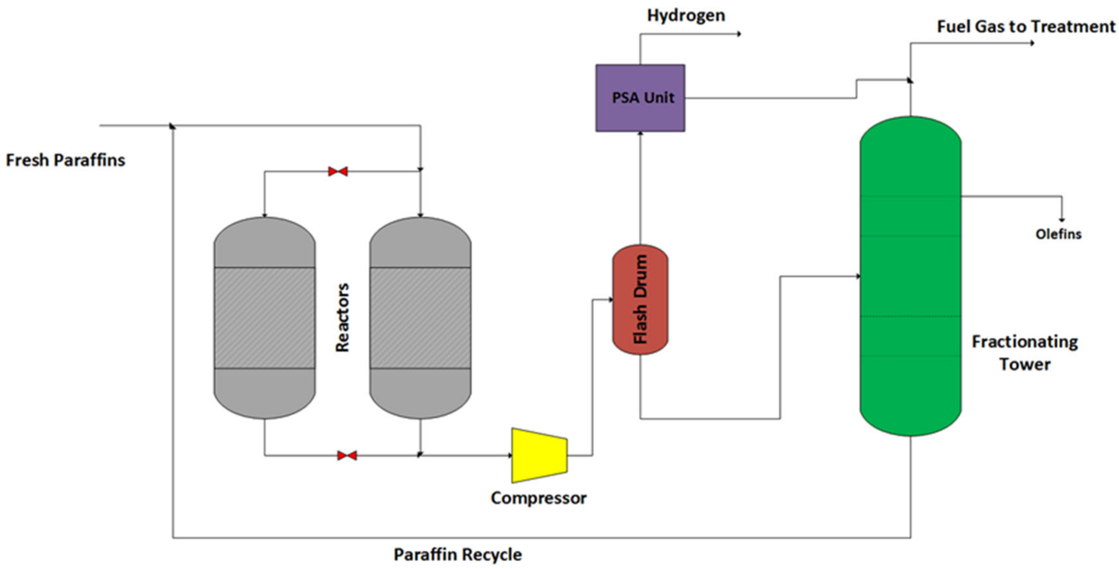

Figure 3 – Simplified Scheme for CATOFIN™ Dehydrogenation Technology by McDermott Company

Conclusion

The synergy between refining and petrochemical processes raises the availability of raw material to petrochemical plants and makes the supply of energy to these processes more reliable at the same time ensures better refining margin to refiners due to the high added value of petrochemical intermediates when compared with transportation fuels. Another advantage is the risks reduction of transportation fuels oversupply, facing the current scenario of demand reduction and restriction of fossil fuels.

It's important to consider that integrated processes lead to a higher operational complexity, however, given current and middle term scenarios to refining industry, a better integration between refining and petrochemical processes is fundamental to the economic sustainability of the downstream industry.

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Reactor Problem Solving | Part 1

Joe M. Bonem

Solving reactor problems can often be both an interesting and challenging assignment. This is true whether the reactor is in a plant, pilot plant or on a bench scale. Most reactor problems can be classified into two broad categories – temperature control and reaction kinetics. Part I of this document deals with temperature control or perhaps lack of temperature control would be a better title. A future edition will cover problems associated with kinetics.

Essentially all reactions have a heat of reaction. In addition, 85 to 90 % are exothermic. Exothermic reactions are those that generate a given amount of heat for each pound of material that reacts. These will be the subject of this discussion. When considering heat removal from an exothermic reaction, the first question deals with what is the heat of reaction. There are multiple ways to determine the heat of reaction for any given process. They are as follows:

Literature Sources – For many reactions, the open literature contains the heat of reaction. There may be conflicting data if more than a single source is considered

Heats of Formation – The heat of reaction can be calculated from the heats of formation given in various data tables.

Experimental Sources – While this may seem like the most reliable method. Facilities should be carefully designed to eliminate heat loss from the experimental equipment.

Of course, the heat to be removed must be adjusted to include factors such as sensible heat of the incoming feeds or for vaporization of incoming reactants in the case of a gas phase or boiling reactor.

Another data need when considering reaction temperature control is how does the reaction rate vary with temperature. This is particularly true when considering the loss of temperature control where an increase in temperature causes an increase in reaction rate which might lead to a higher temperature. This is determined by the Arrhenius constant. Again, literature sources and experimental sources are possible modes of determining the Arrhenius constant. In addition, there is a well-known approximation that the reaction rate of any reaction doubles with each 20°F increase in temperature. The utilization of the Arrhenius constant is described later.

The focus of this article is a continuous reactor with both catalyst and feed being injected continuously. However, the same principles apply to a fixed catalyst bed continuous reactor. The analysis of batch reactors is somewhat different.

Removing the exothermic heat of reaction in these continuous reactors requires a combination of heat transfer and adequate fluid flow. An example of this is an agitated stirred tank reactor with either an internal heat transfer surface or a jacket. The agitator in the stirred tank provides fluid flow across the heat transfer area. The internal tubes or jacket provide the heat transfer area. For steady state operation:

$$Qg = Qr + Qs \qquad (1)$$

Where:

Qg = Heat generated by the reaction.

Qr = Heat removed by cooling.

Qs = Heat removed by incoming feed (sensible or vaporization).

When considering equation (1), it is clear that if the heat being generated (Qg) exceeds the heat removed by cooling (Qr) plus the heat removed by the incoming feed (Qs), that the temperature of the reactor will begin to increase. This may or may not result in an uncontrollable increase in temperature. This will depend on both how fast the reaction increases with temperature and how the heat being removed increases with temperature. The most obvious cases occur when the circulation rate decreases significantly or the reactor heat transfer area fouls. In both of these cases, the heat being removed by cooling approaches zero and the reactor temperature becomes uncontrollable. This is often referred to as a "temperature runaway". Often, facilities are in place to inject a chemical that stops the reaction which will avoid a release from the safety valve.

In a more complicated scenario, assume that the reactor temperature experiences a slight increase. This might be due to an increase in production, an increased activation of the catalyst or inadvertent increase in the catalyst rate. In this case to determine if there will be a temperature runaway equation (1) can be differentiated with respect to the reactor temperature and expressed as an inequality to give equation (2). If this inequality is true, there will be an uncontrollable increase in temperature. If the inequality is not true, then any small increase in reactor temperature will result in the heat removal capability being greater than the heat generated and the temperature of the reactor will return to the control point.

dQg/dT > dQr/dT (2)

In the development of equation (2) and equation (4), there are several assumptions as follows:

The sensible heat being removed is constant. Thus, the differential of this term is zero.

The cooling water temperature is constant. This will occur if the demand on the control system is at a maximum before the slight increase in reactor temperature.

The thermal capacity of the reactor was not considered since for the question of would a temperature runaway occur this was not significant. If one were interested in how fast the reactor temperature would increase, this would have to be considered.

The validity of this inequality can be evaluated by using equation (2) along with equation (3) through equation (7) shown below:

Qg = K *e(-22000/R*TX) (3)

Qr = U*A*lnDT (4)

Where:

K = A constant that depends on monomer and catalyst concentrations, reactor volume, and heat of polymerization. A typical value for this specific process and operating conditions is 3.9*(1014). In addition, the reactor temperature is 130°F and the cooling water temperature is 100°F.

TX= Reactor absolute temperature, °Rankine.

R = Gas constant 1.987 BTU/(lb-mol-T).

U = The exchanger heat transfer coefficient BTU/Ft2-°F-hr.

A = The exchanger surface area Ft2.

LnDT=The log temperature difference between the reaction side and the cooling water side.

The value of 22000 represents the Arrhenius constant.

Equation (4) can be simplified by realizing that when considering the initial situation when the temperature of the reactor becomes uncontrollable, the cooling water temperature will likely be at a minimum and will not be changing. In addition, the reactor temperature will be constant. In this case equation (4) becomes

Qr = U*A*(TX - TW) (5)

Where:

T X= The reactor temperature, °R.

TW = The average cooling water temperature, °R.

Since the temperatures are a difference, they can be either Rankine or Fahrenheit.

These two equations can be differentiated with respect to the absolute reactor temperature (TX) with the following result.

dQg/dTX = (K*11000/TX2) *e(-11000/TX) (6)

dQr/dTX = U*A (7)

Equation (7) is valid since the water temperature is constant.

For the reactor temperature to remain under control, the following inequality must be satisfied:

dQr/dTX > dQg/dTX (8)

EXAMPLE CASE STUDY

Now let's look at the case of a polymerization reactor where the heat is removed by internal tubes. The heat transfer area fouls gradually and when in the judgement of operations, it is "dirty" it is removed from service for cleaning.

The question is – "What is the minimum heat transfer coefficient for operating if it is desirable to avoid a temperature runaway?"

A typical reactor might have the following constants:

Reactor Temperature = 130°F = 590°R

Cooling Water Temperature = 100°F

K value	=3.9(1014)
A	=2500 square feet

Using equation (6), the dQg/dTX term equals 87250.

To avoid the temperature runaway

87250 < dQr/dTX

Since dQr/dTX =U*A and A =2500, U>35 BTU/hr-ft2 -°F

To be conservative, a value of 38 to 40 should be used as the minimum acceptable heat transfer coefficient. If the heat transfer coefficient drops below this level due to fouling, the reactor should be removed from service for cleaning. This will avoid potential temperature runaways with the related safety risks and extended downtime for cleaning.

While this example problem is of a stirred tank reactor with internal cooling coils, the concept and basic equations are valid for gas phase reactors, loop reactors or boiling reactors with an external condenser.

A more difficult consideration is a batch reactor where the exothermic reaction depends on addition rate of a reactant. An exothermic batch reactor will be covered in a future document.

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Optimizing Papermill Operations

John Fowler, P.E.

Papermill production engineers and production professionals today are constantly "in the hunt" for ways to save money with little or no capital investment. This has been especially true since the 1970's when there was a big push for energy conservation and emphasis on bottom line costs of production throughout the paper industry in the U.S. This "push" expanded internationally through the end of the last century and is critical today. This article discusses and spotlights areas of concern and action throughout a typical papermill.

This article is not an exhaustive analysis of anything in a papermill, but rather a discussion of good places to look for the beginner or intermediate professional who wishes to control and reduce costs and improve mill operations. We will not be able to make an exhaustive study of cost and material losses in a papermill here . Rather this article should give you a good start toward moving forward and optimizing the operations in your papermill. More experienced professionals might find some new ideas to put to use as well.

This article will focus on process based cost improvements that can be accomplished with little or no capital outlay. These types of improvements are always advantageous to a papermill.

There are some limitations to processes covered. Paper machine forming and drying are not included due to different process considerations required for this area. Paper machine forming and drying is covered by others and may be the subject of a future article on papermill operations and optimization. Some coverage of paper machine wet end operations will be included.

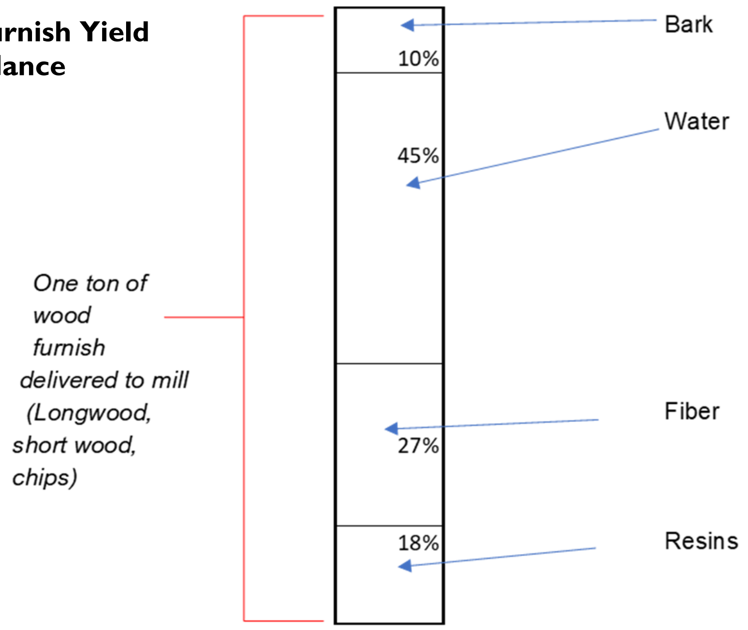
Objectives of this article include:

- Giving the papermill professional places to look for cost savings
- A short form wood material balance for a typical papermill
- Some basic understanding of major product and dollar loss in papermills
- A presentation of starting points for more investigations for the papermill production professional
- A working form of potential cost savings for various operations and areas in the papermill
- A basic understanding of how and why fiber and chemical losses equal dollar losses to the mill.
- Practical methods for pursuing these losses and how to adjust process parameters and get "buy in" from stakeholders such as management, supervision, accounting, operators, etc. for help making process improvements.

Cost performance opportunities exist for today's papermill engineer and production professional in these general areas:

- Fiber loss
- Pulping chemical loss
- Treatment chemical loss
- pH adjustment and control
- Operations management methods to utilize information available.

Figure 1: Wood Furnish Yield Based Material Balance



Material balance in a typical southern craft paper mill and/or many hardwood or groundwood mills and general cost review

The diagram in Figure I is simplified but gives us an idea of what a partial material balance for a papermill looks like. Though a Kraft paper mil is depicted here, this balance can be adjusted and used for groundwood, sulfite, and other papermill processes. Specifics discussed here will translate fairly easily to those other types of mills. For another discussion of Kraft papermill yields, see Reference (1).

Right from the start it is good to realize that in a Southern Kraft papermill depicted here the wood supply is pine, and about 10 % of the weight of the wood furnish is bark. Of what's left after debarking operations, half of the weight is water. What's left on a dry basis is resins and other non-fiber components that make up about 40% of the total dry weight of the debarked wood.

As an example, per each one ton of longwood or bolt wood or wood chips received (with bark included) by the woodyard, one can subtract 10% of the weight for bark. So 45% of what's left is water. In other words 100 tons of wood furnish would equal approximately 90 tons of debarked wood. Of this debarked wood half would be water so that takes us down to about 45 tons of dry debarked wood. If 60% of this dry basis wood is fiber, then we would get 27 tons of fiber from this hypothetical 100 tons of tree furnish. The remaining 18 tons is resin solids which will produce black liquor in the pulp mill, as well as a small amount of byproducts turpentine and Tall Oil soap. Production of these byproducts is not considered here.

Major product and dollar costs and loss in papermills

Table I shows a cost illustration for the typical Kraft mill we have been discussing here for an unbleached kraft paper or linerboard product. We are using a sales cost of $400 per ton of finished product.

Fiber loss cost savings

There are many areas of the papermill process where profits can positively impacted by pursuing fiber loss.

Some of these areas include the pulp mill, the whitewater system, and anywhere water is being sent to the sewer system, whether process or storm sewer.

Table I is simplified but gives us an idea of what where to look for cost savings opportunities. Though a Kraft papermill depicted here, Groundwood and other types of papermill operations could have similar diagrams. The quantities and specifics discussed here will translate fairly easily to those other types of mills, with a few adjustments. The figures shown in Tables I and 2 will, then vary from mill to mill.

For our purposes in this article, we are going to use the production cost of fiber as $300/ton based on finished paper at 7% moisture. So the sales value of dry fiber would be $321/ton.

Let's look at what it costs to sewer 100 gpm of whitewater that has a fiber content of 0.5% by weight. Many mills have a stream like this somewhere going to the sewer.

	% of Production Cost	$/ton
Wood	46	138
Pulping chemicals	24	72
Purchased electricity	8	24
Water	4	12
Labor (all)	12	36
Treatment chemicals	3	9
Insurance and misc. overhead	3	9
Totals:	100 %	$300

Table I: Kraft mill costs of production (based on a $400/ton product FOB sales price and $300/ton production cost)

The amount of fiber wasted to the sewer per day would be =

100 gal whitewater/min x 1440 min/day x 8.3#/ gal x

0.5# fiber/100# whitewater = 5,976 # fiber/ day,

or 3 tons at a cost of 3(321) or $963/day, $28,890 per month, $351,495 per year for a 365-day year.

Identifying and correcting a whitewater and fiber loss like this one could cover the cost of your salary for more than a year!

Whitewater

The term whitewater is used rather generally here. For our purposes here, whitewater is any mill water stream that is not stock or liquor of any kind, and not a treatment chemical additive. In a perfect world, all of the mill whitewater would be recycled and none would be sewered. This is difficult to achieve, especially in older mills built in the last century. Conserving whitewater along with fiber or chemicals it contains in an older mill is challenging.

Some proposed operational guidelines for paper machine whitewater system (2) include:

Maintain totally separate broke handling and fines recovery systems.

Sewer only excess clear (filtered) whitewater.

Minimize load fluctuations to the Saveall.

Provide independent whitewater systems for each section of each paper machine.

Table 2 shows places to look for fiber loss related to whitewater flows in the papermill.

Saveall effluent
Whitewater storage tanks or chests
Whitewater sewered at the wet end of paper machines.
Basements of paper machines (Safety first! Paper machine basements can be dangerous places!)
Bleach plant effluents
Bleach plant pulp washing operations
Overflows or leaks around pulp storage chests or silos
Anywhere one sees water running onto concrete or the ground or floor
Broke system

Table 2: Places to look for sources of fiber loss

It is worth keeping in mind that there are many opportunities for substantial amounts of fiber to get lost in processes throughout the papermill. Many pieces of equipment need to be "fine-tuned" in order to provide the most efficient operation where there is minimal fiber loss. Many processes are easier to operate with slightly larger volumes of water flow through the process. These are sometimes operated without a lot of attention to controlling fiber loss.

It's also worth noting that the typical production parameters and final product quality do not relate directly to fiber loss. So it takes it a fair amount of attention to detail and dedication to run down sources of fiber loss and report these in a concise understandable way to the papermill management team and stakeholders. Stakeholders are interested in reducing fiber loss and profit loss in a papermill. Coming up with a good way to report the fiber loss graphically or some other way to stakeholders in the papermill can be very helpful to all involved.

For each ton of production there is a certain amount of profit attached to this ton. So each ton of fiber loss translates in to lost profit, as well as dollars lost in other raw materials. So a ton of fiber loss would result in a profit loss of roughly $321 if the mill is running in a “sold out” position.

Chip Quality

Another area of potential dollar loss is in the chip quality and its effect on the pulping process. By chip quality I mean the uniformity of chip thickness and size that will lead to a predictable pulping operation and predictable fiber quality from the pulp mill. Others have published detailed articles about chip quality but suffice it to say that the more uniform the chip thickness and chip geometry that is provided to the pulp mill, the more successful the pulping operation. A successful pulping operation produces strong and uniform fiber for a high quality and uniform paper product. The chip quality also affects the production of high-quality liquor which can be recovered in the recovery boilers.

Liquor quality is in part also a product of how possible it is to produce and utilize a uniform liquor with uniform wood chips as discussed above.

Table 2 is basic information but shows areas within the papermill which can result in fiber losses. Pulling samples at appropriate times from waste streams will help estimate fiber loss. It is important to operate any equipment that handles fiber anywhere or any stream handling fiber anywhere in the mill close to optimum conditions to minimize fiber loss.

Samples of the wastewater effluent from the mill to the treatment or holding system will give an idea of total fiber loss from the mill. The fiber content of the stream can be determined by the papermill lab.

Measuring the final wastewater fiber content at the outfall is not a typically a very good way to determine fiber loss. This is because there will have been settlement and possibly other treatment of mill effluent upstream of the outfall.

Table 2 can provide a rough checklist that the papermill production professional can use to form a strategy for moving forward with looking for fiber losses.

Pulping Chemical Loss

The $72/ton used here for pulping chemical includes caustic, sulfur (if used), salt cake (if used), purchased lime, acid, soda ash (if used) and could include another chemical or two. Bleaching chemicals are not included for purposes here. As discussed above for whitewater, sewering whitewater may be unavoidable in some mills. This sewering can waste pulping chemicals and waste money. Leaks and imbalances in systems throughout the mill can also result in chemical losses.

Table 3 includes some good places to look for these chemical losses. Efforts to reduce sewering of process water typically pays off.

Dregs washer

Slakers

Stock thickeners

Pulp washing operations

Savealls and disc filters

Whitewater storage tanks or chests

Green liquor storage areas

White liquor storage areas

Whitewater sewered at the wet end of paper machines.

Bleach plant effluents

Bleach plant washing operations

Overflows or leaks around pulp storage chests

Anywhere one sees water running onto concrete or the ground or floor

Table 3: Places to look for pulping Chemical loss

Treatment Chemical Loss

Various treatment chemicals are used throughout the mill to ensure smooth operations. The chemicals provided by treatment service Vendors can solve a lot of problems. These Vendors will report detailed usage and lab analyses upon request. Request this reporting if it is not already being provided.

Review of this Vendor information is often assigned to the younger engineers or the “lowest person on the totem pole”. However, this information is of great importance to many others.

The person reviewing Vendor treatment data and the Vendor’s treatment philosophy should ask a lot of questions. This Vendor information should be shared with middle management, supervision, and rank and file employees as well. All of these people have very valuable and intimate knowledge of the process, Their knowledge represents years of investment by the mill. This knowledge can be utilized to efficiently use Vendor treatment chemicals and minimize costs in this area.

Based on Table 1 figures, a 1,000 ton/day mill spends $9 per ton on these treatment chemicals, which equates to $3.3 million per year for a 1,000 ton production per day mill. A 10% savings here of $0.90/ton would be welcome in any papermill.

pH Adjustment and Control

Maintaining pH targets are critical to the safe and efficient operation of a papermill. pH measurement can be tricky, but working with the pH instrumentation Vendor can help by providing training and education for all personnel involved in each pH control loop.

Unfortunately, pH control systems leave a lot to be desired in many applications in papermills. Though flow and temperature and pressure control equipment is usually taught to engineers in college, this is not the case for pH control.

Ways to save money by proper pH control can be found in the boiler house, water treatment, and other areas.

One area where a substantial cost saving can be achieved is at the wet end of the paper machine. Sizing chemicals for the paper product are often added and controlled at the wet end of the paper machine. Alum and other chemicals may be needed to properly “set” the size, and alum and these other chemicals are expensive. Some of these chemicals have a buffering action and they can be overfed in an effort to control pH. Controlling pH with acid and caustic is a much more efficient way to control pH. The treatment chemicals can be base-loaded and pH controlled to get the best treatment chemicals performance, thus minimizing costs in this area.

Making the necessary changes to the process here must be done very deliberately and gradually. The first order of business is to discuss the current wet end chemistry operations with all the stakeholders. Then gather data and review. Then propose changes and run a written plan to make gradual changes by all stakeholders. Proceed cautiously and you could save $1 to $5 per ton or more as changes are gradually phased into the operation.

<u>Operations management methods to utilize information</u>

This part of this article could also be called : "How to Get People to See things Your Way", or "How to get people to do the things you think they should do to save money" .

People dislike change, but most of us know that change is inevitable in our lives. This also applies to lives in a papermill. Years of experience has taught many of us that this is the hardest part of being a production professional in a papermill.

Books have been written on this subject, but let it suffice to say that one must proceed cautiously in this area to achieve success. Table 4 shows a list of "ideas" for proceeding in getting buy-in from other parties for changes one wants to make to papermill processes.

Proceed cautiously

Talk to all stakeholders to understand process dynamics and key process parameters

Defer to each stakeholder's experience and ask for buy in. Don't' always expect to get buy-in

Gather and carefully record data

Discuss ways to effect change with stakeholders

Identify key personnel at all levels who will help effect the changes

Communicate pertinent process data through charts, graphs, etc. for all to see

Produce written instructions for all to review and comment on before implementing any change.

Be honest with everyone about the process changes and what success will look like

Congratulate operations people appropriately for their assistance

Recognize stakeholders for milestone successes with a free lunch, small gift, etc. if possible.

Put in place an appropriate management system to keep important changes going over time.

Keep management informed regularly of successes and cost savings

Table 4: Ways to Effect Process Changes and Get Buy-In from Other People

Effective and frequent communications with stakeholders is a key to success!

Terminology and definitions for this article can be found in reference (3).

I hope you have enjoyed this article and have found useful information for your situation. There are so many ways to save money in a papermill, and I wish you success.

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About the Author

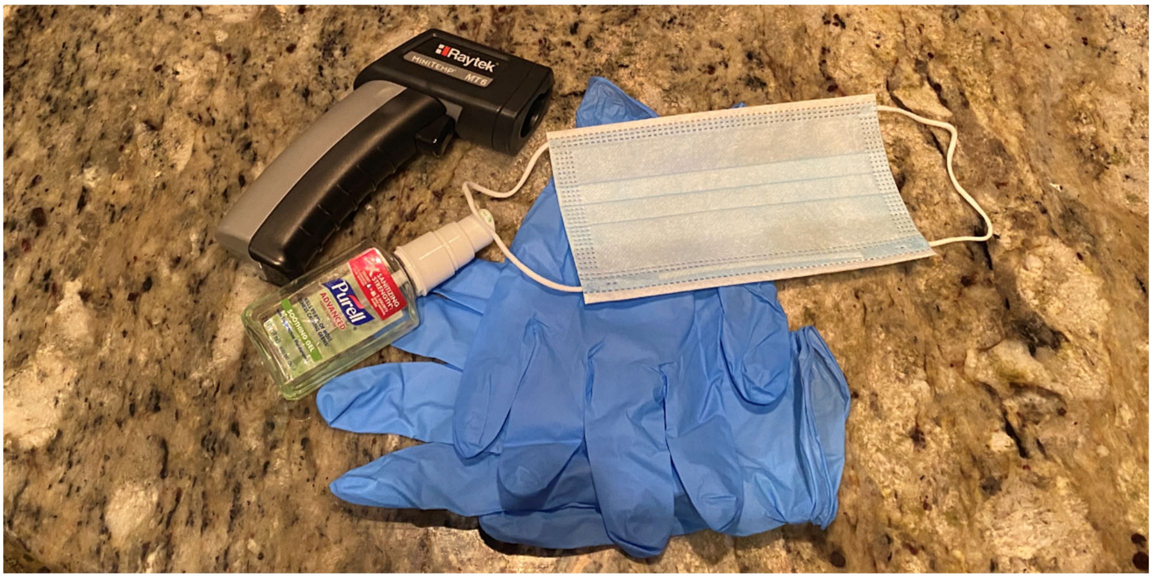
John Fowler, P.E. is a Senior Consulting Engineer at KLM Technology group. He has spent over 40 years performing detailed design for a large number of chemical processes including papermills, petrochemical facilities, refineries, as well as many types of wastewater treatment facilities. Mr. Fowler has been accepted by District, State and Federal courts as an Expert Witness in over 30 court cases.

John has over forty years of progressive experience in the operation, process design and process troubleshooting of a wide of processes as well as years of experience in chemical and papermill production and technical supervision, chemical, mechanical and environmental engineering, cost estimating, and process safety management. He is an active member of TAPPI and AIChE.

Quarterly Safety Connector

For Engineers; Because Safety Is Part Of The Process! By: Chris Palmisano, MESH, IFSAC **July 2020**

Things To Think About Before Reopening Your Operation After A Pandemic Shutdown.



It's happening, slowly but surely, many company operations are phasing into reopening, to what people are calling, "The New Normal". Some employees view the whole pandemic and subsequent shut down as a hoax or nonsense, others are seriously frightened at the thought of returning to work and social activities. It's a lot to think about and requires some planning before you open those doors and gates.

Before your re-start operations, it's a good idea to first, assemble your Safety Committee to brainstorm, plan and review federal, state and local guidelines and always follow the strictest laws, regulations and standards, just as you would with your safety and environmental policies, COVID is no different.

- Is the workplace safe for employees?
  - Do a top to bottom cleaning of the entire building, including sanitizing with approve disinfectant methods.
  - Are HVAC systems and air ducts clean with high quality filters.
  - Daily cleaning policies should be in place.
  - One entrance and one exit works best to control staff and visitors. A sanitation or decon station at the exit will help reduce the chances of employees bringing a virus home to their family or spread potential virus in the community.
  - Should you do screening of visitors and employees at the entrance?

    Well it's not a bad idea, at-the-least for a little while. Take temperatures using No-Touch digital forehead thermometers that can display a body temperature within seconds. Ask questions like have you been to any social at the door with an unyielding warning that states: <u>If you have any of these symptoms, you cannot enter this building</u>.

Documenting temps and answers to screening questions can be very valuable information in the event of claims of negligence or in a worker compensation claim case. An employee for example, may insist that they caught the virus at work, yet this same person may have stated that they attended a sporting event or a wedding with 50 people over the weekend, prior to getting sick.

- o Remember to treat COVID illness claims like any other OSHA "Recordable" injury or illness case, until proven otherwise. Employee <u>Deaths</u> and any <u>Admissions to a Hospital</u> are "Reportable" like any other work related injury or illness case.
- o Consider closing common eating areas or at-the-least reorganize lunchrooms and kitchenettes to provide social distancing. It may be beneficial to provide a time schedule for the kitchen, just like we do with meeting rooms, where people can eat at different times, without overwhelming the room with people, all at the same time.
- o Employees are safest when eating at their desk or at a clean isolated workstation and they should use disposable cups, eating utensils, bowls and plates.
- o Limit business travel to essential travel only and introduce all travel slowly. Start with small trips first and use reputable hotels and encourage your road warriors to be cautious about where they eat. Any road warrior like myself, that travels often, will already be keenly aware of where they should stay and eat. Coach those that are less experienced with business travel on the dos and don'ts.

- Encourage Employees to keep themselves and work areas clean.
  - o Provide what's needed for good sanitation and hygiene
    - ♦ PPE (personal Protective Equipment) Gloves, Eye Protection and Masks
      - ◊ Maintain a PPE use chart and an inventory so you don't run out
    - ♦ Hand Sanitizer
    - ♦ Sanitizer wipes for personal items, such as desk, phones, pens, tools, hand trucks etc.
    - ♦ Paper towels or air dryer unit
    - ♦ Put foot/arm pulls on lavatory doors to discourage touching door handles.
    - ♦ Automatic water flow devices should be on sinks, urinals and toilets.
    - ♦ Liquid soap (for washing hands) signage may be necessary to encourage proper hand washing
    - ♦ No common or community computers, tablets, pens or pencils
    - ♦ Clean phones regularly
    - ♦ Have a temporary isolation area to handle sick people and hold them safely until they can leave or be picked up.
  - o Remind employees with training, signs and/or posters in common areas:
    - ♦ DON'T COME TO WORK IF YOU ARE SICK
    - ♦ Post the symptoms of COVID-19
    - ♦ Coughing/sneezing etiquette (they should cover their mouth)

- ♦ Discourage face touching
- ♦ Hand washing practices
- ♦ Limit visitors/vendors/sales people and others from entering the building
- ♦ Remind workers to social distance
- ♦ Wearing PPE

o Remember one this, about change, complacency is as contagious as the COVID virus. When supervisors and managers are positive about the new policies and safety precautions, staff will be diligent. If management bad-mouths the company policies and they don't take them seriously, neither will the workers.

- Most Important, HAVE FUN! We are back to work! People spend more time with work colleagues throughout a career than they actually spend with their families. Humans are social creatures that crave interaction and a positive atmosphere.

  Do things to entertain staff or help to distract them from this so called new normal of distance, quarantine and isolation. Things like virtual games, mind challenge contests and even a virtual happy hour can help staff unwind. People need a chance to socialize and catch up. Don't forget work anniversaries, birthdays and company announcement. This is an important time to be transparent with your staff and show them you care.

  When it comes to production, lighten up and be malleable as everyone gets back to the grindstone. This has been an ordeal, with information changing every few hours. People are drained, tired and stressed. The uncertainty of what this virus will do in the future is taxing on everyone's mind. Give them a break and let them settle in slowly and don't just tell them, SHOW them how much they are appreciated and how wonderful it is to have them back.

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