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# Troubleshooting Practice in the Refinery

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## INTRODUCTION

Even in the best run plants operations degrade over time or some processes do not work as well as expected. Plant operating and equipment changes introduce new problems and reveal, older, unappreciated problems. Troubleshooting, the systematic investigation or problems and their solution, is key to maximizing plant profits. A unit with malfunctioning equipment or an incorrect process configuration cannot be optimized by the best control system available. Processes and equipment must work correctly for maximum profits.

Key aspects of successful troubleshooting include: a thorough understanding of both the process and the equipment: application of chemical engineering basics to problem solving: and good field technique and data interpretation skills. The process is not independent of the equipment. The equipment works as part of a process. Both must be understood for effective troubleshooting. Chemical engineering basics are required for understanding both the equipment and the process. Finally, field technique and data interpretation skills are needed to gather required information. Often, troubleshooting fails due to faulty, incomplete and misleading numbers.

This paper demonstrates application of these keys in a case format. A wide variety of cases are presented with an emphasis on distillation troubleshooting. Key details are shown on each case to illustrate the type of problems encountered and basic elements of troubleshooting analysis. For brevity, many secondary details have been eliminated from the discussion.

The ten case studies are (in order):

- Capacity loss due to subcooled reflux in a petrochemical tower.
- Viscosity variation in an atmospheric stripping tower product due to a home-made distributor.
- Solvent loss in a vapor draw in an aromatics unit due to weeping into the draw header.
- Operational instability in a petrochemical tower due to faulty conversion from a once-through to a circulating reboiler.
- Inability to meet feed rates or product specifications in a benzene-toluene extractive distillation unit due to improper equipment installation and lack of re-distributors and hold-downs.
- Foam induced flooding of an amine stripper in a gas plant due to faulty level indication.
- Inability to meet product purity and recovery specifications in a petrochemical tower due to fouling and an out-of-level distributor.
- Control instability on reboiler heat input on a stab-in reboiler on a petrochemical plant deisobutanizer due to improper design.

- Inability to meet product purity and recovery specifications in a petrochemical tower due to a malfunctioning level controller.
- Liquid-liquid wash column performance limited by a fouled and incorrectly installed feed distributor.

## CAPACITY LOSS DUE TO SUBCOOLED REFLUX USE

Sometimes feeds going to a distillation column are subcooled. One reason a feed may be subcooled is because during energy optimization of the plant the feed stream going to the column was identified as a source of heat that could be cross-exchanged with colder stream. Another reason is it may be advantageous to subcool the overhead product of a column at the condenser instead of using an additional product cooler. The biggest problem is that these factors are often not taken into account when sizing column internals. This may lead to a premature flooding, loss of column efficiency, and reduced capacity.

Excessive sub-cooling of a reflux feed to a distillation column can lead to a variety of operational problems. Excessive subcooling of a reflux feed condenses some of the internal vapor traffic. This, in turn, increases the liquid traffic in the affected area of the column (Figure 1). Often, the operations department will try to counter this by cutting back on the amount of reflux being introduced into the column or by adjusting the condenser duty. Reductions in reflux being introduced will have an effect on the effectiveness of the column [1].

Sub-cooling also reduces overall column efficiency. The subcooled reflux or feed shifts some of the internal equipment from a mass-transfer service to a heat-transfer service.

### *Subcooled Liquid Feed Mechanism*

Sub-cooled liquid feed is at a temperature below its column-pressure bubble point. The effect of a subcooled feed or reflux can be estimated by [2]:

$$L_F = F\left(1 + \frac{h^* - h}{H_{eq} - h^*}\right)$$

| | |
|---|---|
| $L_F$ | change in liquid flow at the feed stage |
| F | total moles of feed (reflux) |
| H | molar enthalpy of liquid feed at conditions to the column |
| h* | molar enthalpy of liquid feed at the column pressure boiling point |
| $H_{eq}$ | molar enthalpy of vapor which would exist in equilibrium with the feed if the liquid feed were at the column pressure boiling point. |

Referring to Figure 1, we see that $L_F$ equals $L_2$-$L_1$.

When a sub-cooled liquid feed is used, the increase in liquid molar flow at the feed stage is greater than the liquid molar feed rate alone. Vapor rising to the feed stage is condensed in order to raise the feed conditions to the bubble point temperature. The condensing vapor increases the liquid flow leaving the feed stage, flooding the column (Figure 2).

***Operational Example***

After a revamp of a commercial petrochemical column the column was started up and lined out. Operations brought the column up to the new design-operating rate. Before the column reached the new design-operating rate the column started to experience a loss of efficiency. The capacity of the column fell five to ten percent short of the design capacity. The column was gamma scanned and the scan revealed that the top five to six trays had an extremely high liquid level on the tray active area. The down comer also had an extremely high clear liquid back up. All of these conditions were consistent with a premature flooding condition.

A test run was performed to evaluate the column's performance and to verify the design. The data collected from the test run was used to evaluate the model.

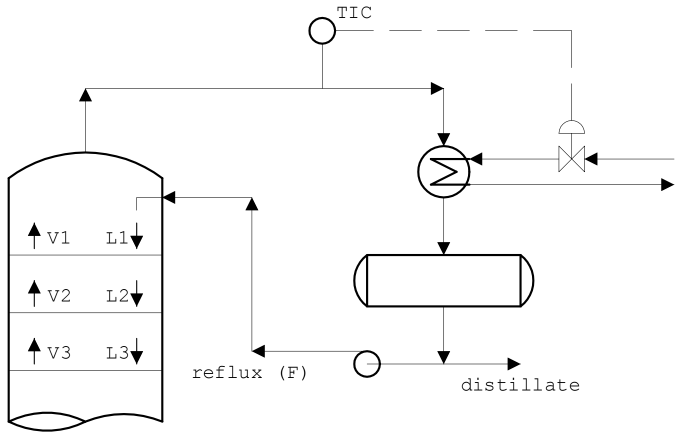
While reevaluating the model, it was discovered that 100°F (55ºC) subcooled reflux was being introduced into the column. The use of sub cooled reflux was missed during the design phase of the column revamp project. The simulation was rerun using the subcooled reflux conditions.

The results of the simulation indicated that the liquid traffic in the rectification section of the column was dramatically higher than previously used in the design. The internals were re-rated with the loadings from the simulation. It was determined that the column would get about five percent less capacity than planned. This was consistent with the results seen from the simulation and the gamma scan.

Sub-cooling the reflux was a normal part of the operation of the plant. The reflux was subcooled to make the plant more energy efficient. This practice could be abandoned if necessary. The operations group agreed to increase the temperature of the reflux in order to determine if this was the problem with the column. Once the reflux was introduced to the column at bubble point temperature, the capacity and efficiency of the column increased. The column was able to handle the new design-operating rate and the efficiency of the column was within design specifications.

***Conclusions***

The introduction of subcooled feed into a column may cause operational problems and lead to premature flooding. If a subcooled reflux feed is going to be used, the effects of this stream must be accounted for in the design of the column.




$$L_2 - L_1 = L_F = F(1 + \frac{h^* - h}{H_{eq} - h^*})$$

Figure 1
Subcooled reflux to tower

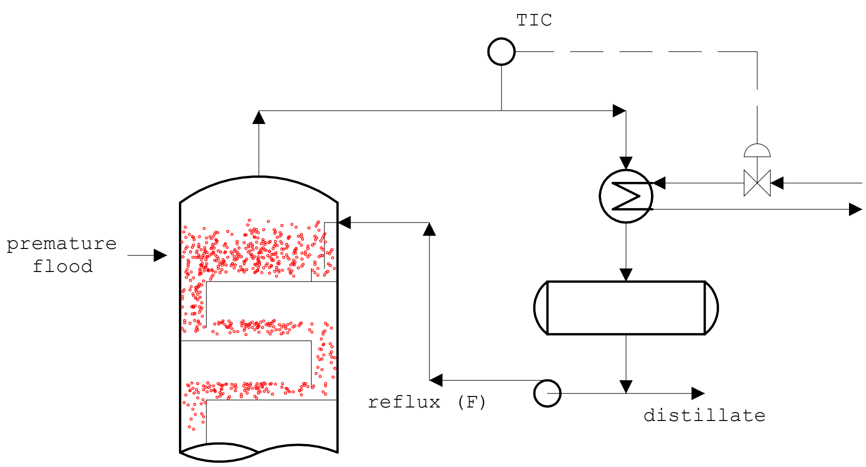



Figure 2
Subcooled reflux and premature flood

***REVIEW SOLVES OPERATING PROBLEM***

Operating problems with a client's product quality from an atmospheric pressure stripping tower had become unacceptable. The processing unit was a two-foot diameter atmospheric stripping unit with a gas feed as the stripping medium to meet a viscosity specification on the liquid product. In operation, the viscosity cycled in a range between approximately five centipoise below the specification value to seven centipoise above the specification value. The product specification had an acceptable range of plus-or-minus one centipoise.

The viscosity variation downgraded the product to a lower value market disposition. Incurred costs were approximately five hundred dollars per day in lost profit potential. Continuing efforts by plant maintenance during unit shut-downs had not been able to locate the source of the problem.
One of the authors was called in to troubleshoot and solve this problem. A staged approach was used to approach the operation. The major steps included:

1. Review of the unit operation
2. Verification of the equipment and piping configuration
3. Problem identification
4. Solution preparation
5. Solution installation

Figure 3 shows a perspective view of a liquid distributor that the author found installed in the column. Six years earlier, the plant had converted the stripper from a trayed column to a packed column. During the conversion, three homemade "supports" as shown in Figure 3 were installed to support the packing beds. Figure 4 shows the support locations in the side stripper. No after-modification test was run to check the performance of the modified unit.

The homemade support installed has a major defect. It had insufficient open area for vapor flow (9.2%). In a packing support plate, vapor flowing up and liquid flowing down have to share the same openings. For the liquid to flow down, the liquid height above the support plate has to overcome the pressure drop of the rising vapor. A small open area gives a high-pressure drop. The result was a cyclic operation where liquid stacked up above the distributor plate until enough liquid head was available to force the liquid down against the force of the rising vapor. Once liquid flowed, it dropped down the tower in a slug. This drained the liquid level above the support plate and started the entire cycle again. The liquid slugging showed up as a product viscosity variation.

After finding the problem, a proper support plate was installed. Figure 5 shows the correctly designed support plate. A large open area through the support mesh gives low-pressure drop. An external ring around the edge (not shown) gives overall mechanical strength. After installation of the new support plates, the stripper worked correctly with close to zero viscosity variation.

The project cost $40,000 to replace the plates. The existing tower had to be shut-down, the packing dumped, internals removed, new internals installed, and the packing replaced. Payout for the project recovered the entire cost in only 80 days.

However, with proper selection of seemingly minor equipment item (the support plate) a $1,000,000 loss over a six-year period could have been avoided. In cases where product quality is important, locally made solutions should be avoided unless proper engineering experience is obtained to verify that the equipment will work correctly. Smaller, cheaper equipment is often judged to be less important than the larger, more expensive equipment items. This is not always true. Experienced care should be obtained in all steps of mass-transfer equipment installation and repair.

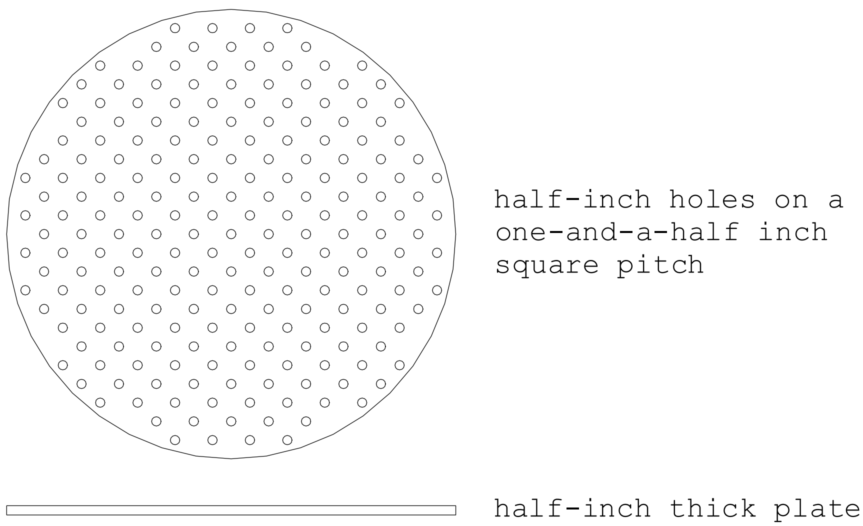



**Figure 3**

Home-made support plate

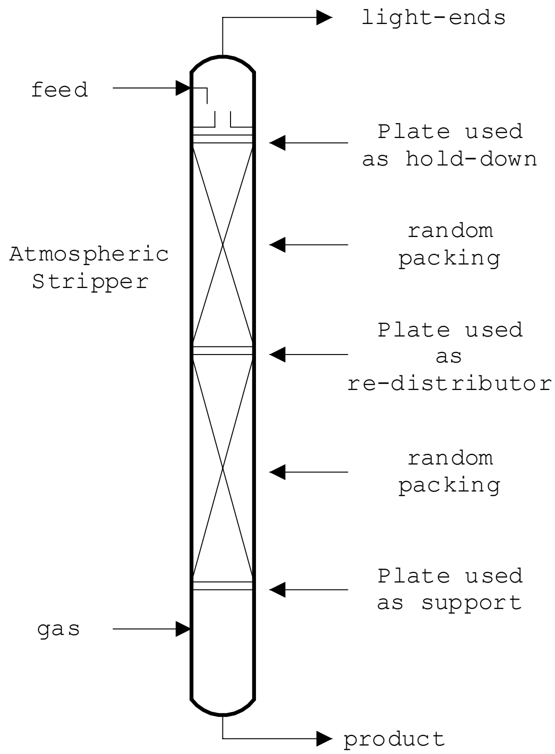



**Figure 4**
Stripper elevation schematic

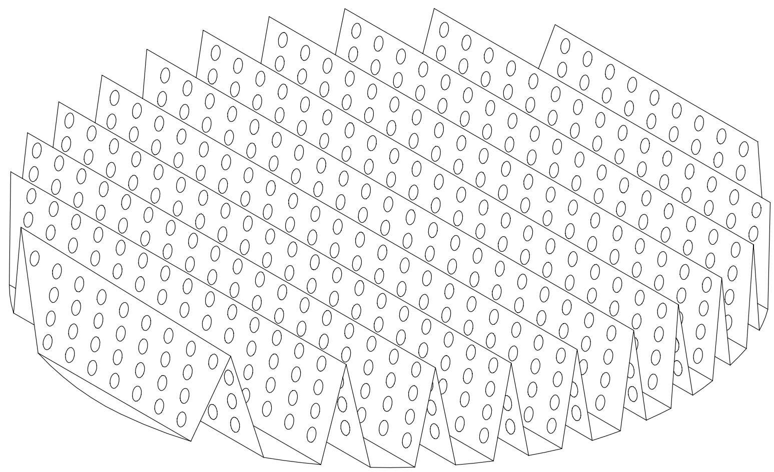


**Figure 5**
Random packing support

## *VAPOR DRAW INCREASES LOSSES AFTER REVAMP*

In a debottlenecking study of a Texas BTX unit, the stripper tower was determined to be the unit limit. The unit was built in the 1960's and had under gone several previous revamps to increase capacity. The unit included a liquid-liquid extractor, flash drum, stripper tower, raffinate water wash tower, aromatics clay treater and three aromatic distillation towers (Figure 6).

The Stripper tower contained three-inch cast iron bubble caps. Bubble caps have limited distillation capacity, but a positive liquid seal. The normal unit flow was feed to the middle of the liquid-liquid extractor, raffinate was the overhead product and the aromatics with the solvent was the bottom product. The rich solvent then went into the lower-pressure flash drum. The flash drum removed light non-aromatics and then the solvent with the aromatics was introduced to the stripper tower.

In the stripper tower, a secondary flash was included and the aromatics stripped by steam from the solvent into an aromatic vapor sidecut. The aromatic vapor sidecut was cooled, decanted and then sent to a clay treater to remove olefins.

The stripper was determined to be vapor limited and a decision was made to retray the tower with high-capacity valve trays. The high-capacity trays used hanging downcomers to include additional active area under the downcomers [3]. Because of the age of the unit, the original rings were cast iron and bolted in place. New tray rings were installed. The high-capacity trays were installed and tower was commissioned.

The stripper tower's main specification was to remove all of the aromatics from the solvent that was then recycled to the extractor tower. On this point, the retray was a total success. Increased stripping reduced the aromatics in the recycle solvent below the ability of laboratory detection.

However, a problem arose when the unit was commissioned. The aromatic vapor sidecut now contained a much higher amount of solvent. Independent review at this point discovered the causes of the problem. The valves did not have a positive liquid seal and they weeped slightly, letting some solvent drop into the vapor draw header (Figure 7).

Correct vapor draw design takes into account possible leakage [4]. Vapor product draws should include shrouds over the vapor inlet ports to prevent liquid entrainment (Figure 8).

Rather than shutting down for immediate repairs, modifications were postponed to a future turnaround.

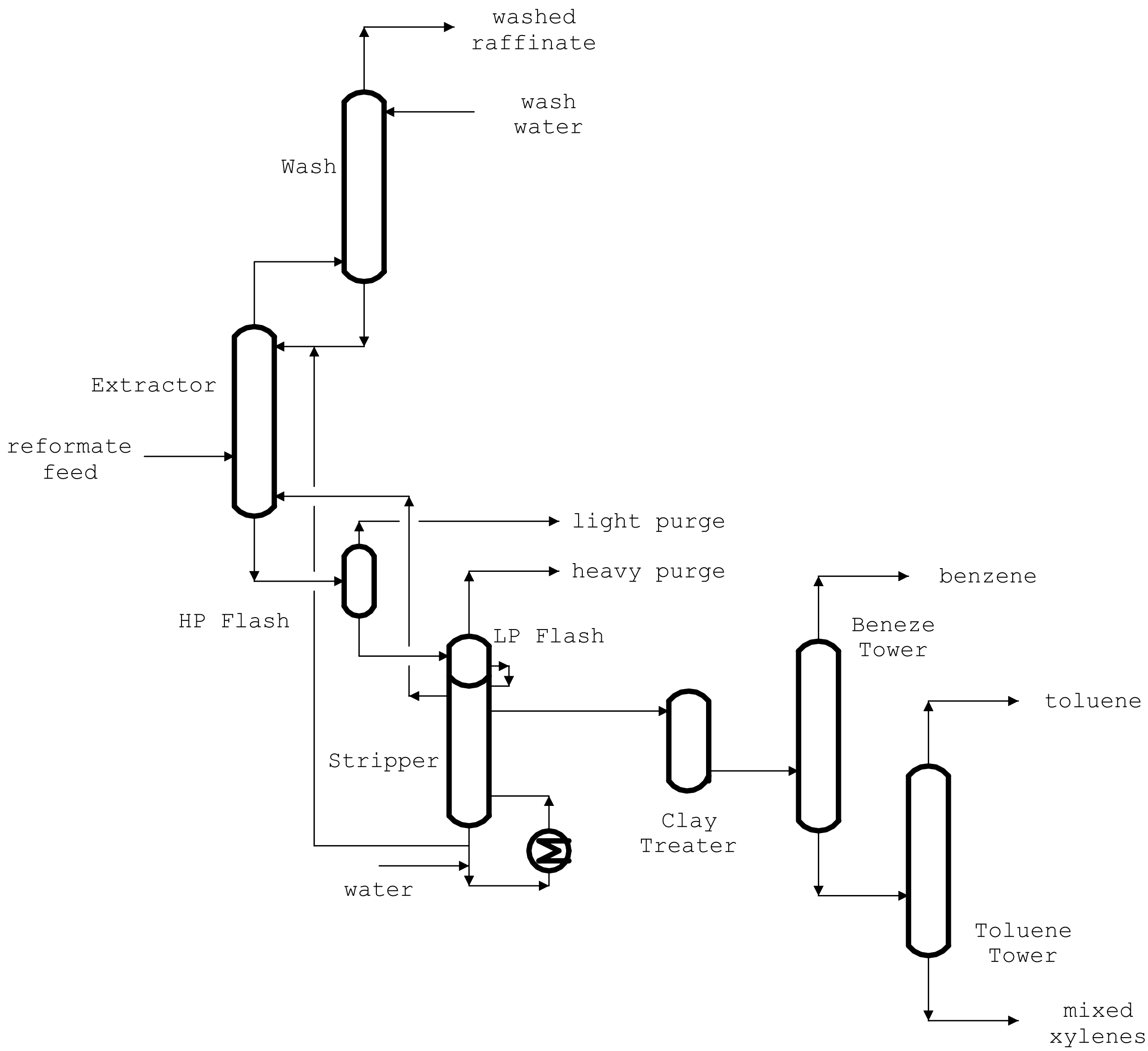



**Figure 6**
Aromatics Unit

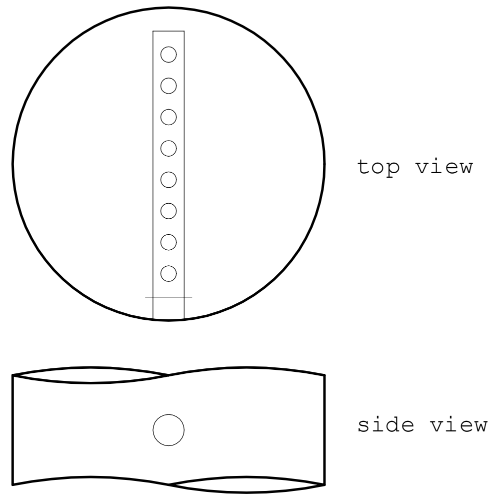



**Figure 7**
Unshrouded vapor side draw

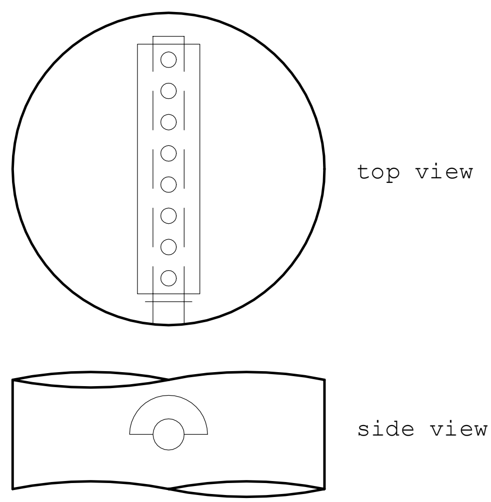



**Figure 8**
Shrouded vapor side draw

***POOR COLUMN OPERATION DUE TO A MALFUNCTIONING REBOILER LOOP***

***Before the Revamp***

A petrochemical column was revamped with new trays. The column was a high-pressure service column (greater than 100 psig, 690 kpag). The column used a once-through thermosiphon reboiler for column boil-up. Before the revamp, the existing column arrangement had a feed pipe inside the column coming from a sump through the boot and then out the tower bottom to the reboiler feed line (Figure 9).

***After the Revamp***

After the revamp, the column was started up and lined out. The column did not work. The problem was determined to be insufficient boil-up. This was found through analysis of field data and comparison of simulation work done before the column was revamped. After reviewing the revamp design of the column and inspection records, it was determined that the internal pipe feeding the reboiler and the tray sump had been removed. The liquid level driving flow through the reboiler was substantially reduced, thereby creating flow instability problems in the reboiler loop (Figure 10).

Additionally, removal of the vessel internals converted the thermosiphon from a once-through configuration to a recirculating thermosiphon. This removed approximately one theoretical stage from the tower. However, as the tower had more than sufficient stages, the loss of one stage was not significant.

***Thermosiphon Reboilers***

A once-through thermosiphon reboiler is classified as a natural circulation reboiler. Natural circulation is achieved by drawing liquid from the bottom tray and circulating it directly to the reboiler. The stream that returns to the column is partially vaporized. The net return liquid is recovered as bottom product (Figure 9) [5].

The performance of a reboiler is based on the equation:

$$Q = UA\Delta T_{CLM}$$

where,

Q heat transferred, btu/hr
U overall heat transfer coefficient, btu/hr-ft$^2$-°F
A exchanger heat transfer area, ft$^2$
$\Delta T_{LM}$ corrected log mean temperature difference, °F

A reboiler may have heating zones: liquid sensible heating to its bubble point, liquid vaporization, and vapor superheating. At low liquid rates, the liquid feed entering the reboiler is essentially totally vaporized. In this situation, the liquid heating zone is small and nucleate boiling occurs almost immediately. In the reboiler above the nucleate boiling zone, mist flow develops and the heat transfer is gas superheating. The nucleate boiling heat transfer coefficient is high. However, the gas superheating coefficient is very low. Net, high vaporization in reboilers reduces the effective heat-transfer coefficient.

The temperature difference in the liquid heating and the nucleate boiling zones is high and relatively constant because the low two-phase static head and hydraulic losses in the tubes and outlet piping do not raise the boiling point significantly. In the mist flow zone, the temperature rise can be significant because gas heating can result in a severe temperature pinch at the outlet for low liquid levels. The temperature rise in the mist flow zone results in a low overall integrated temperature difference that, when combined with the low overall heat-transfer coefficient, explains the poor reboiler performance at low liquid rates [6]. Low liquid rates in recirculating thermosiphons are caused by low liquid levels in the tower.

In order to increase the amount of liquid flow the liquid driving head must increase. The hydrostatic pressure difference between H1 and H3 has to be decreased (Figure 10). As the liquid driving head increases, the circulation rate increases, lengthening the nucleate boiling zone and reducing the mist flow zone. As this occurs the performance of the reboiler improves.

***Fixes for the Problem***

One option to repair the column was to redesign the bottom tray and reinstall the internal down pipe to feed the reboiler. This would result in extensive amount of downtime due to the amount of welding that required to complete the repair. Another option was to raise the liquid level of the column to within six to 12 inches (15 to 30 cm) of the reboiler return nozzle (Figure 11). To prevent tray flooding, the seal pan required modifications. The effective control range for the bottoms liquid rate was narrow, but still possible. This option required very little downtime. The amount of hot work to be performed in the column was minimal. The repair decided upon was option two.

Downcomer backup was checked at maximum operating rates. The total amount of downcomer liquid back up would be eight inches (20 cm) including three inches (7.5 cm) of downcomer submergence. This was acceptable and would not interfere with the operation of the column.

The modifications were made and the column was restarted. The seal pan was modified and the operating liquid level in the column increased to within six to 12 inches (15 to 30 cm) of the reboiler return nozzle (Figure 11). The column performed as expected. The column product was within specifications. There were no operational problems.

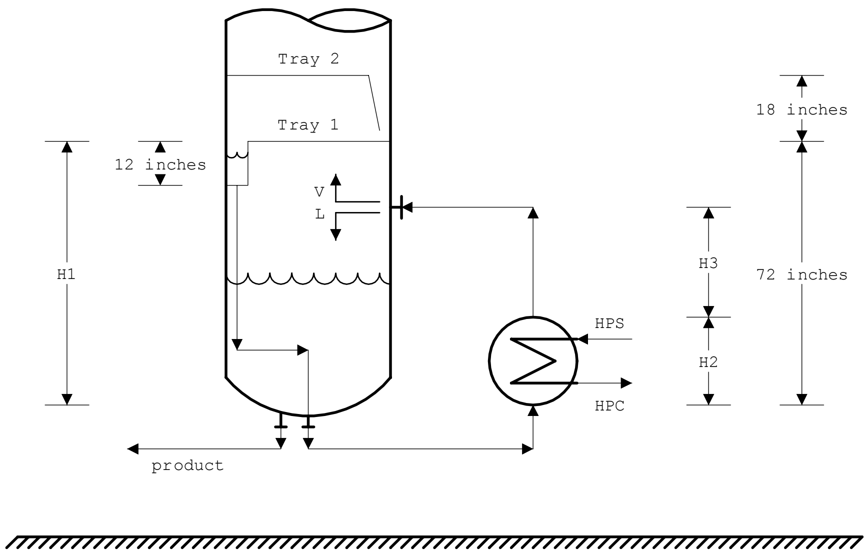



**Figure 9**
Once-through thermosiphon arrangement, before revamp

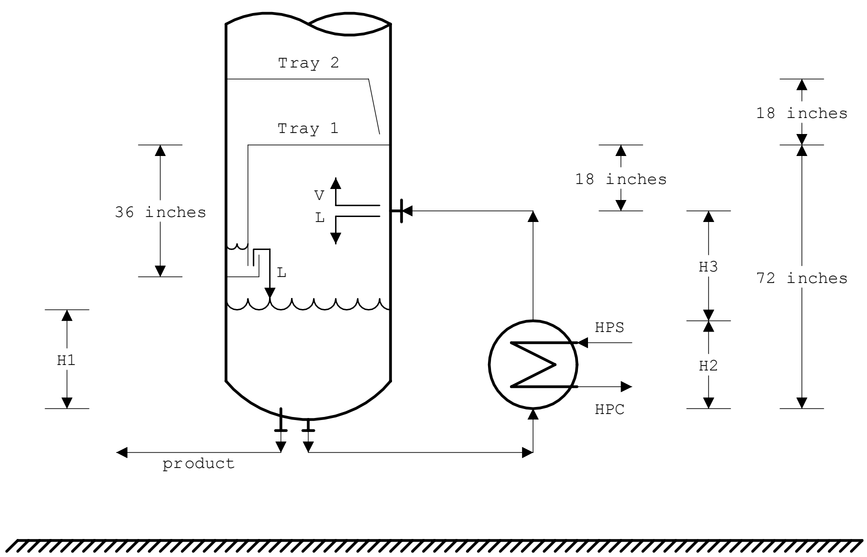



**Figure 10**
Recirculating thermosiphon arrangement, after revamp

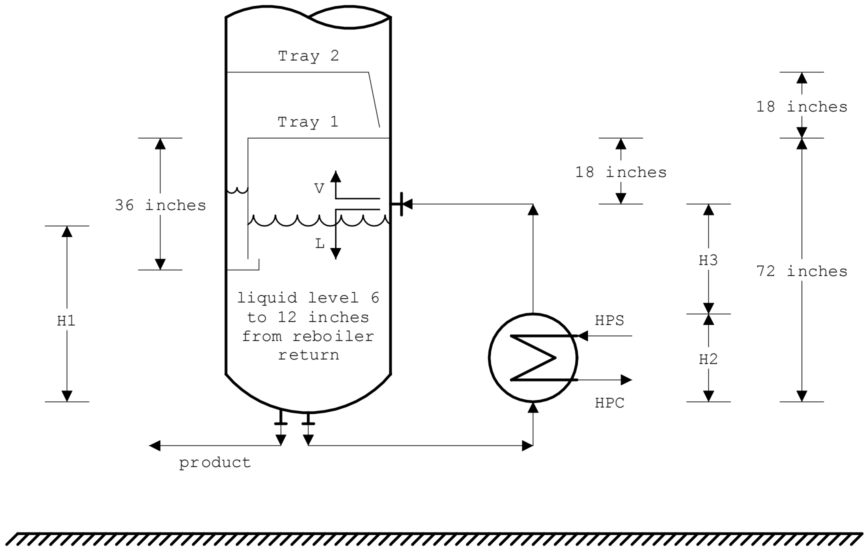



**Figure 11**
Recirculating thermosiphon arrangement, after fix:
sump modified and liquid operating level raised

## *IMPROPER DESIGN AND INSTALLATION LIMITS MALAYSIAN BENZENE-TOLUENE EXTRACTIVE DISTILLATION UNIT*

A new Malaysian extractive distillation unit was commissioned with limited success. The unit consisted of a reactor pretreatment section with two distillation towers, an extractive distillation (ED) column, a stripper column, and a benzene-toluene distillation column.

At start-up, a power failure caused a unit upset that dislodged trays in the ED column. The trays were replaced in kind. After tray replacement, the unit failed to meet design rates and recoveries. The ED tower has 4 structured packed sections in the rectifying section (top of the tower) and 37 trays in the stripping section (bottom of the tower). The top two packed sections are 12 feet (3.66 m) tall and the next two are 26 feet (7.32 m) tall. Typically, packed bed sections of above 20 feet (6.1 m) are not recommended.

The tray and packing supplier reviewed the operation and decided that new trays in the bottom would allow at guarantee conditions. The unit was shut down and trays replaced.

During the shutdown, maintenance personnel inspected ED (Figure 12) and stripper towers. In the ED tower packing was found to be up to 1/2 inch (12.7 mm) from the circumference of the wall. The packing also contained up to 1/2 inch (12.7 mm) openings in the middle of the bed. This can lead to channeling down the wall or through the openings in the middle of the bed with a loss in packing efficiency.

On the top of the third bed, inspection showed that the distributor was supported by 4x8 inch (10x20 cm) plates on top of the packing. The plates covered 8% of the packing. The bed would have had a higher capacity if the distributor was supported without the plates.

The bottom of the third bed did not have a liquid collector or redistributor. Liquid was allowed to freely rain down on the fourth bed. A liquid collector and redistributor has two functions. First is to collect the liquid and route it to the next tray. The second is to evenly distribute the vapor across the upper section.

Below the chimney tray above the second bed, the packing did not have a hold down grid. The packing appeared to be upset under the chimney tray.

The bottom 37 trays were replaced with enhanced liquid and vapor capacity trays and the unit recommissioned. The unit failed to meet design guarantees on either product quality or capacity.

Further work is underway to implement solutions to the problems found by the maintenance department. These repairs will be made during the next unit shutdown.

Some important lessons from this include:

1. Solutions to problem towers should be reviewed by independent parties to assure that the right problem is being fixed.

2. Packing should fill the vessel. Packing strips should fill gaps between structured packing bricks and between the structured packing and the vessel wall. This insures that the packing has no open areas. Typically, packing is “shoe horned” in place by thin sheets of metal that are expanded, packing is then inserted between the sheets, and then the shoehorn is removed.

3. Do not allow mechanical convenience to dictate over process requirements in equipment design.

4. Do not allow distributor supports to block vapor and liquid flow through packed beds. Have as high an open area for liquid and vapor flow as possible.

5. Packed towers need to have collectors, distributors, redistributors, and hold-downs at all appropriate points.

Engineers need to insure that packing sections are fully packed, contain hold-down grids, and have a collector and redistributor. Problems are easier to prevent than to fix.

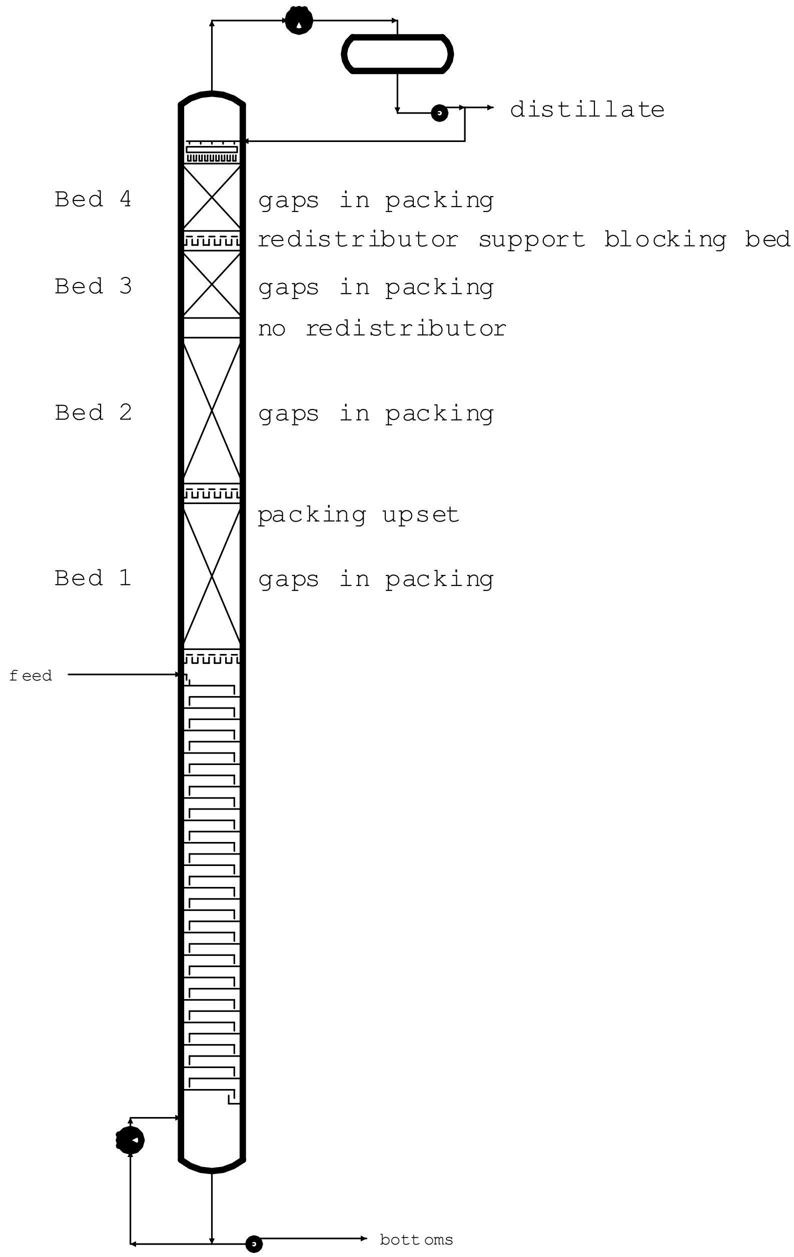



**Figure 12**
Inspection results on extractive distillation tower

## *AMINE STRIPPER UPSET TRACED TO FOAM*

A client had flooding problems with an amine stripper used to regenerate a proprietary amine blend. The unit worked effectively from start-up to the first unit turnaround. External events extended the turnaround into an extended shutdown. After several months the unit started up again. However, after restarting, the amine stripper flooded at 70% of previous reboiler duties. Symptoms observed were:

- Rapid loss of bottoms level in the stripper
- Rise in pressure drop across the stripper
- Eventual loss of amine into the stripper reflux drum

These are classic symptoms of tower flooding. For this tower, the entire reflux drum filled with amine and had to be purged at least once every five days.

Field investigation eventually located a problem with the bottoms level control system. Taking a sample from the upper level tap during operation showed liquid in the sample, even though only vapor should be at this vessel elevation (Figure 13). Manually changing bottoms rates by large percentages showed no change in bottoms level. When foam fills the distance between a DP cell's pressure taps, the DP cell measures the liquid density, not the level in the vessel.

Amine units commonly suffer from foam formation in the tower boot. Differential pressure (DP) cells used to measure the bottoms liquid level convert a differential pressure into a liquid level based on an assumed boot liquid density. An incorrect density assumption can easily flood towers in this service (Figure 14). Contamination of system amine inventory during the shutdown was the suspected source of the foaming problem.

Foam control chemistry was changed and stringent monitoring of amine system contaminants started. As a short-term fix, the level in the boot was deliberately kept at low indicated values to reduce the possibility of foam entrainment onto the trays.

### *Foam major source of unit upsets*

Foaming is a major cause of column problems. Typical petrochemical and refining services where foam causes level control problems include crude preflash, amine strippers, sponge absorbers, and extraction solvent regeneration. In general, any surface active chemistry system suffers from stabilized foam formation. Other services with a high potential for foaming upsets are thermosyphon reboiler systems in high relative volatility systems (absorber-strippers, for example) and stab-in reboiler services (HF alkylation, MTBE, and others). These services often suffer from upsets due to foam entrainment onto the trays causing tower flooding.

Understanding that DP cells only measures level if the density is known is important in troubleshooting these systems. Control calculations use an assumed liquid density to determine the liquid level inside the vessel. Obviously, if the assumed liquid density is correct, problems do not start. Problems begin when the vessel boot density does not equal the assumed liquid density.

Sight-glasses suffer from the same problem. A sight-glass reports back to the viewer the DP between the vessel taps as a height of clear liquid in the vessel. Sight-glasses, while valuable, must be used with care. Continuing the sight-glass analogy, looking at the exterior liquid reading gives us a lower liquid level than what the low-density foam inside the vessel really has (Figure 15). The control system has the same problem with DP cells. The density used to calculate a level is wrong and the controls allow foam to back over the trays in the tower.

Once foam rises above the upper level tap for the DP cell (Figure 16), then the DP cell (or sight-glass) only shows average density inside the column, not liquid level.

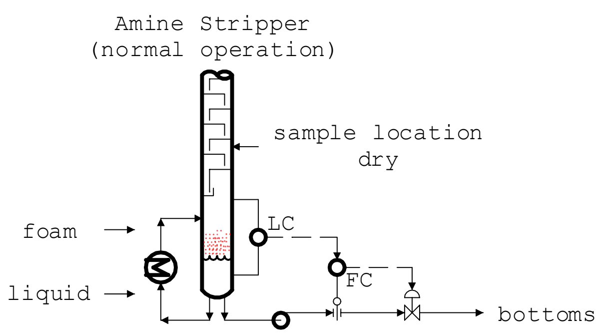



**Figure 13**
Flooding and liquid samples

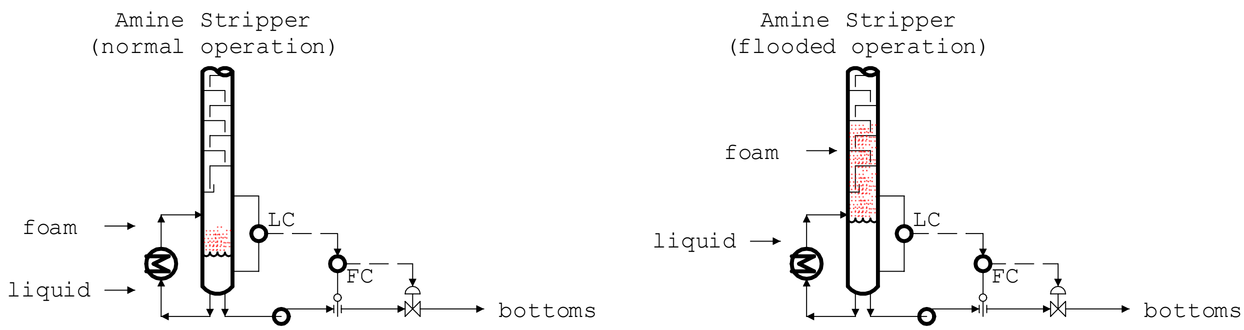



**Figure 14**
Flooded amine stripper operation

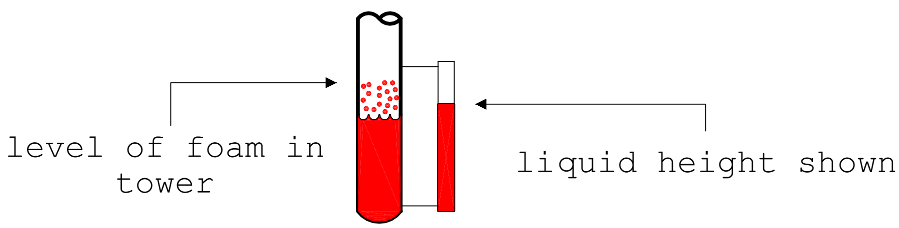



**Figure 15**
Foam and false liquid levels

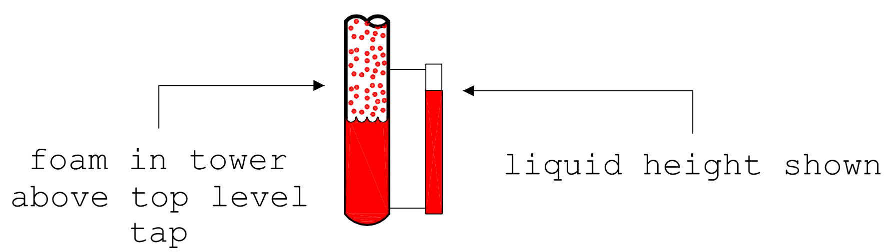



**Figure 16**
Foam and complete filled tower

### *POOR COLUMN PERFORMANCE DUE TO INEFFECTIVE LIQUID DISTRIBUTION*

#### *Liquid Distribution Critical for Packed Tower Performance*

A commercial petrochemical column was achieving poor separation efficiency while operating at normal operating rates. Operational data from the column was collected in order to evaluate the column's performance. The data showed that the column's separation efficiency had deteriorated significantly through the plant's run cycle. There was no indication of operational problems: excessive pressure drop, highly subcooled reflux, or poor reboiler performance.

A natural suspect for performance failure in any packed column is poor distributor performance [7]. Liquid distribution to packed bed columns is one of the most important aspects of packed tower design. Packed towers are more sensitive to liquid and vapor maldistribution than trayed towers. Therefore, it is critical that vapor and liquid enter packing evenly distributed. A number of studies have shown the impact of liquid maldistribution packed tower separation efficiency [1]. The performance of the packing depends heavily on the initial vapor and liquid distribution entering the packing. Poor vapor and liquid distribution to a packed bed can result in a loss of efficiency [8].

The column was gamma scanned and it was determined that severe maldistribution was occurring in one of the packed beds. The other packed beds looked like they were getting fair liquid distribution. Table 1 gives the operational data collected before the column was shut down and repaired.

**Table 1**

| Stream specifications | Column Feed | Overhead product | Bottom product | Reflux |
|---|---|---|---|---|
| Flow, lb/hr | 85,000 | 26,300* | 58,700 | 250,000 |
| Temp, °F | 173 | 110 | 205 | 205 |
| Press, psia | 2.0 | 1.4 | 5.75 | 5.75 |
| Product wt % | 62.0 | 6.5 | 86.9 | 86.9 |
| Impurity wt % | 31.0 | 87.0 | 5.9 | 5.9 |
| Inert material wt. % | 7.0 | 6.5 | 7.2 | 7.2 |

*Product stream leaving the condenser

#### *Severe Distributor Problems Limit Tower Performance*

There are many types of liquid distributors in commercial use. The most commonly used in fractionation services are the trough-orifice distributor (Figure 17) and the pan-orifice distributor

(Figure 18). Both can suffer severe maldistribution problems. Both require proper design, fabrication and installation.

Distributor errors can result in either systematic or random maldistribution problems. Random distribution problems are those where the liquid distribution variations are randomly distributed over the are the distributor covers. Systematic distribution problems are those where the distribution variations are located in a definite pattern related to the distributor geometry.

Systematic errors normally degrade distillation efficiency much more than random errors [9]. Two major source of systematic liquid distribution problems are warped distributors and distributors installed out-of-level. Warped distributors can result from either poor fabrication techniques or temperature excursions in the tower.

***Preparing the Tower Fix***

Once it was determined that liquid maldistribution existed, the design of the liquid distributors was reviewed. Several problems were found with the design. It was determined that the design of the liquid distributor called for a drip point density of six drip points per square foot (65 per $m^2$). This would be considered a minimum number of drip points per square foot for good efficiency. The reason for the low number of drip points per square foot was to maximize the size of the distributor orifices to minimize risk of plugging. The system had a history of fouling. Drip-point layout versus fouling reduction is a trade-off often requiring engineering judgement rather than relying on definite calculations or standard engineering practice manuals. The packed beds in this tower were tall (over 30 ft, 9 m) and a higher number of drip-points would have helped performance. Approximately 10 per square foot would be preferred (110 per $m^2$).

Examining the drip point pattern of the distributor it was determined that liquid distribution at the periphery of the packed bed was marginal. Poor periphery flow is common to some distributor designs where concerns about wall flow caused designers to under-irrigate the outer area of the tower cross-section [1].

After an extensive analysis, two additional conclusions were reached. First, the column had never operated to its full potential. Second, the column performance had degraded over time.

Action items identified included replacing one of the distributors and cleaning fouling from the column. The new distributor was designed with a drip point density of nine drip points per square foot (97 $m^2$). Additionally, the periphery flow of the new distributor was improved.

***Fixing the Tower***

The tower was shut down for repair and the existing distributors were inspected. Several problems were found with the existing distributors. Many of the trough orifices were plugged. The orifices plugged were mostly at the outer area of the distributor. This type of plugging was observed in every distributor in the tower. Figure 19 illustrates the approximate location of the plugging problems.

It was also discovered that every distributor was out of level. Normally, when a distributor is out of level the efficiency of a column falls dramatically. There are many reasons for a distributor to be out of level: installation problems, lack of leveling features in the distributor design, inherent tilt in the fabrication of the distributor, or tray support rings out of level, among others. Figure 20 shows a typical trough distributor out-of-level.

The plugging problem was the cause of the worsening performance over the run. The out-of-level installation was the cause of the tower never meeting its full potential. Many towers fail to meet their operating targets for these reasons. Features can be added to the distributor to reduce the effects of fouling inside a tower [10]. Standard designs often do not include these features. Inappropriate application of standard equipment and practices is a major source of operating problems and lost profits.

The new distributor was installed and the remaining distributors were cleaned and leveled. The column was restarted and lined out. Table 2 gives the performance of the column after the revamp was completed. The overhead product purity improved significantly.

**Table 2**

| Stream | Feed | Overhead product | Bottom product | Reflux |
|---|---|---|---|---|
| Flow, lb/hr | 85,000 | 26,500* | 58,500 | 250,000 |
| Temp, °F | 173 | 110 | 205 | 205 |
| Press, psia | 2.0 | 1.35 | 5.75 | 5.75 |
| Product wt % | 62.0 | 2.3 | 89.0 | 89.0 |
| Impurity wt % | 31.0 | 92.0 | 3.4 | 3.4 |
| Inert material, wt % | 7.0 | 5.7 | 7.6 | 7.6 |

*Product stream leaving the condenser

After the column was revamped, the bottom product purity observed was between 89% and 91%. The overall pressure drop measured in the column was no different than it was before the column was shut down. The fix objectives were met.

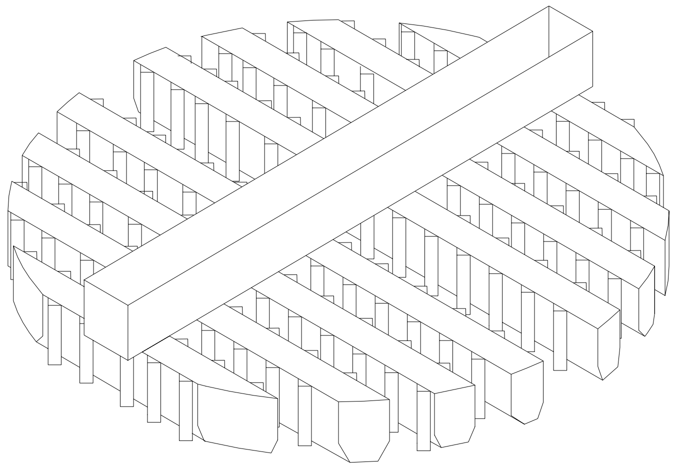


**Figure 17**
Trough-orifice distributor

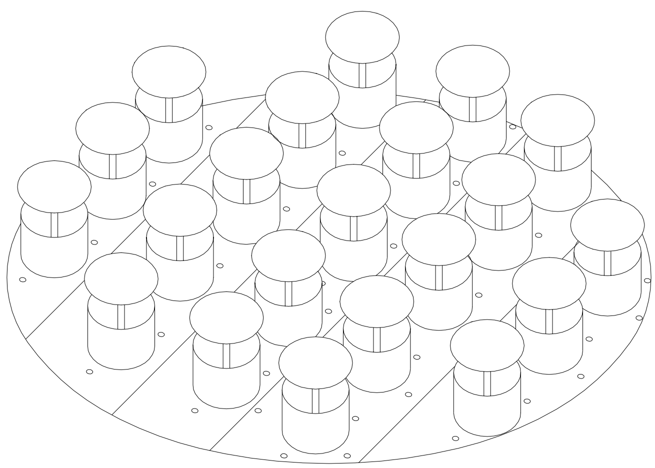


**Figure 18**
Pan-orifice distributor

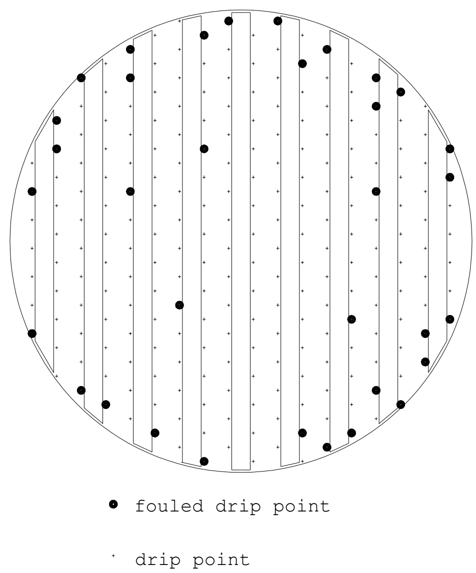



**Figure 19**
Fouling found on trough distributor

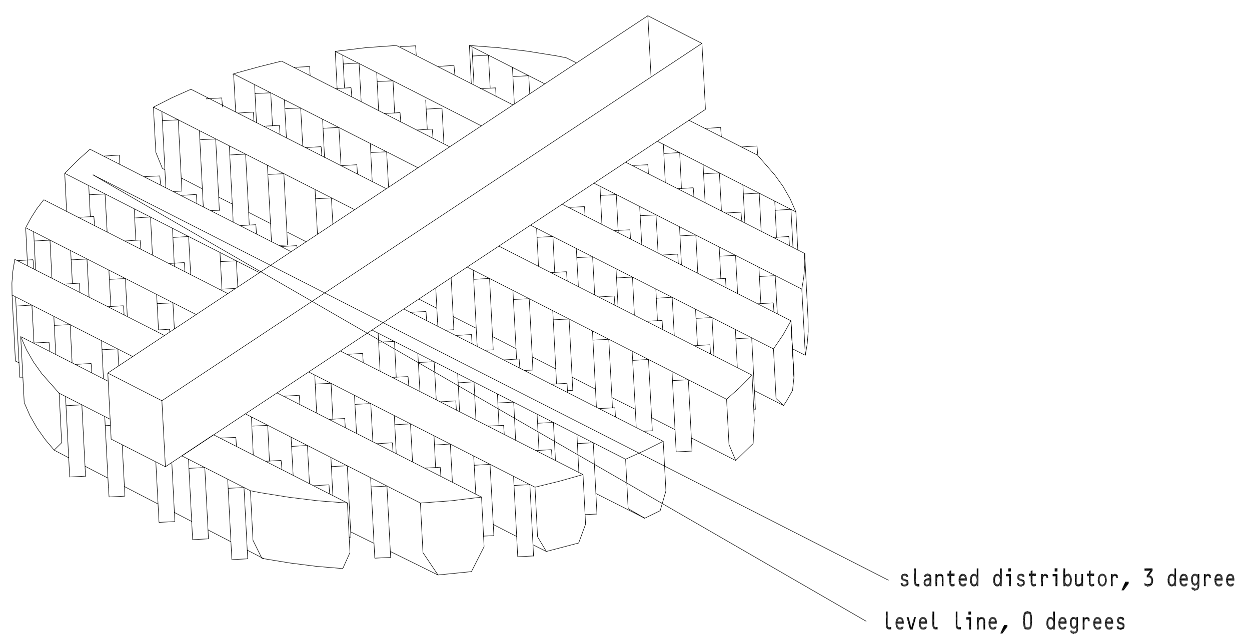



**Figure 20**
Distributor installed out-of-level

## *CONTROL PROBLEMS IN A DEISOBUTANIZER*

A petrochemical plant deisobutanizer suffered from control problems since start-up. The deisobutanizer had three feeds, a vapor distillate purge, liquid distillate product, vapor sidedraw product, and a bottoms liquid purge. The vapor sidedraw product was heat integrated against the unit feed. Figure 21 shows the unit configuration.

The original control scheme used:

- Overhead pressure on direct pressure control
- Air fin motor speed for overhead temperature control
- Hot vapor bypass for tower-to-distillate drum differential pressure control
- Reflux on temperature control in the top section of the tower
- Distillate product on distillate drum level control
- Bottoms and sidedraw vapor product on bottoms level control
- Reboiler duty on maximum required by overhead pressure control, sidedraw vapor product rate, or bottoms product rate

Figure 22 shows the overall control scheme for the tower. Feedforward and feedback connections to other parts of the unit are not shown. This complex control scheme for the tower never worked properly.

After considerable experimentation, the major contributing factor for control failure was found to be the bottoms level. Taking the bottoms level loops out of the control system stabilized unit operation dramatically. The actual control scheme derived from operator experimentation had the bottoms and sidedraw products on flow control. Flow rates were reset manually as the affects on downstream operation were seen. Typically, the bottoms rate was kept much higher than necessary to allow for stable operation of the rest of the plant. Figure 23 shows the configuration used.

However, one problem with this operation was continued difficulty in getting accurate level readings from the bottom level. Bottoms level appeared to be essentially uncontrollable. Wide swings in bottom level could lead to upsets in tower operation and the tray 13 normal butane vapor draw composition.

Two attempts were made at improving bottoms level control. First, a level control algorithm [11] based on reboiler duty, reboiler size, column geometry and estimated froth density was used to estimate bottoms level. The value calculated varied from zero percent level to several hundred percent with stable operation. Operation would also go unstable at unpredictable calculated level values.

Second, a DP cell was added across the tower to attempt to predict bottoms level based on the pressure drop from the lower level tap to the top of the tower. The DP cell measured the liquid height in the boot plus the pressure drop across 135 trays (Figure 24).

Taken together, the two attempts to improve level control helped make a small improvement in tower operation. Further work was postponed for a planned shutdown.

Before the shutdown, a more thorough review of the problem source was conducted. Figure 25 shows the expected configuration of the tower bottoms based on using a chimney tray to hold liquid on the stab-in reboilers used. Figure 26 shows the configuration as installed and shown on the tower and internal equipment drawings. Inspection photographs found in the files verified this configuration.

Only two rows of tubes in the reboiler bundles were wetted by the liquid held on the chimney tray by its outlet weir. With no liquid on the reboilers, heat could not be put into the tower. At startup, the operators had quickly discovered that deliberately running the tower above 100 percent of level allowed the tower to work. Figure 27 shows the minimum liquid level required to make the reboilers work.

Obviously, the minimum liquid level was above the upper tap of the level control range. Several solutions were checked for use at the turnaround:

- Raise the height of the chimney tray outlet weir to hold a liquid level on the exchangers
- Modify the level control instrumentation to use another type of instrument instead of a DP cell.

Due to the very low bottoms rate, less than one percent of the liquid traffic to the reboiler, the second choice, modify the level control instrumentation was selected. At the high liquid rates and with turbulence in the boiling pool around the reboiler, liquid overflowing the weir could be much more or much less than the bottoms rate required to purge heavies from the unit. The tower will continue to operate flooded to above the reboiler bundle levels. A gamma ray level detector is being installed to allow for full advanced control of the unit.

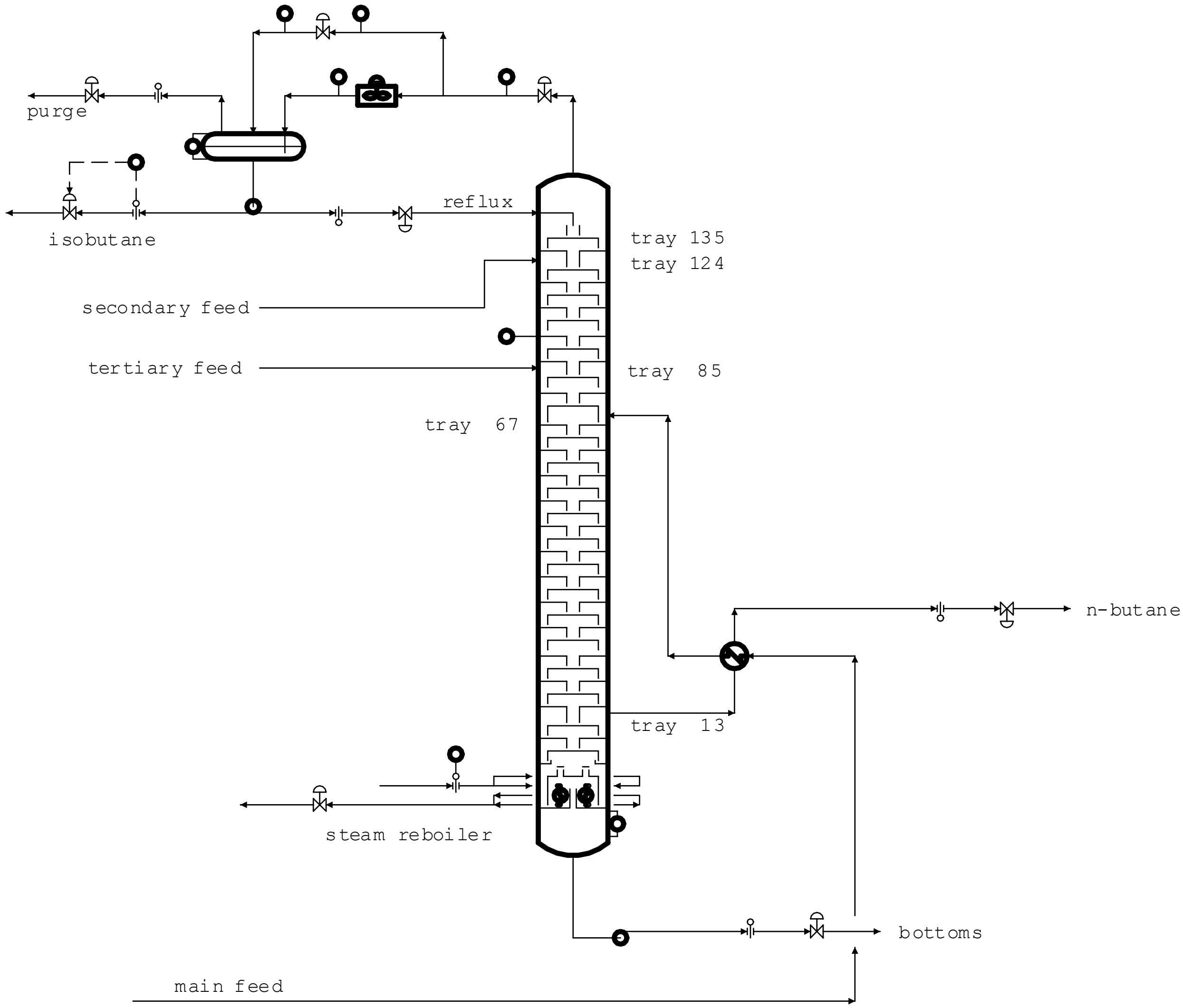



**Figure 21**
Deisobutanizer configuration

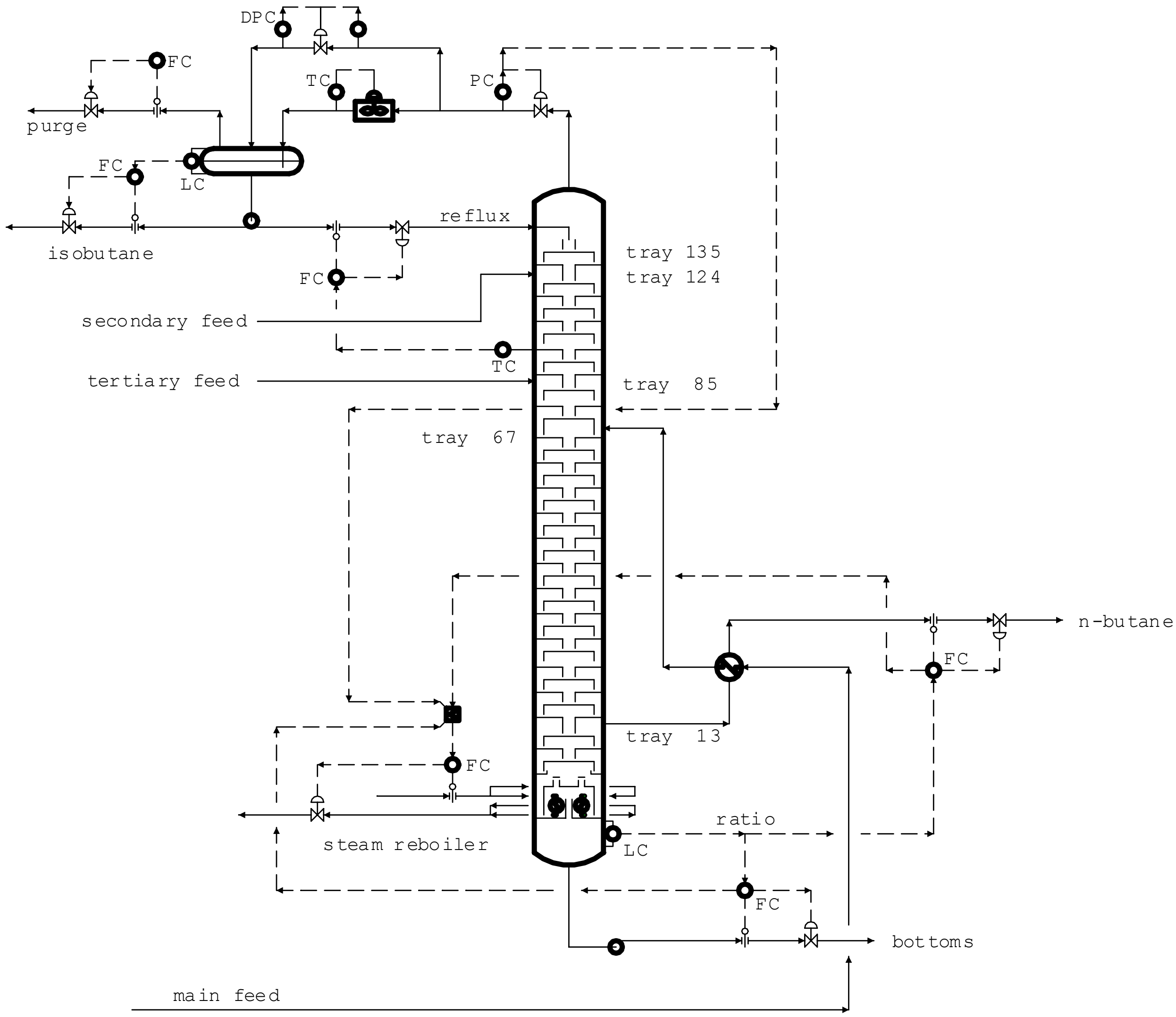



**Figure 22**
Original control scheme

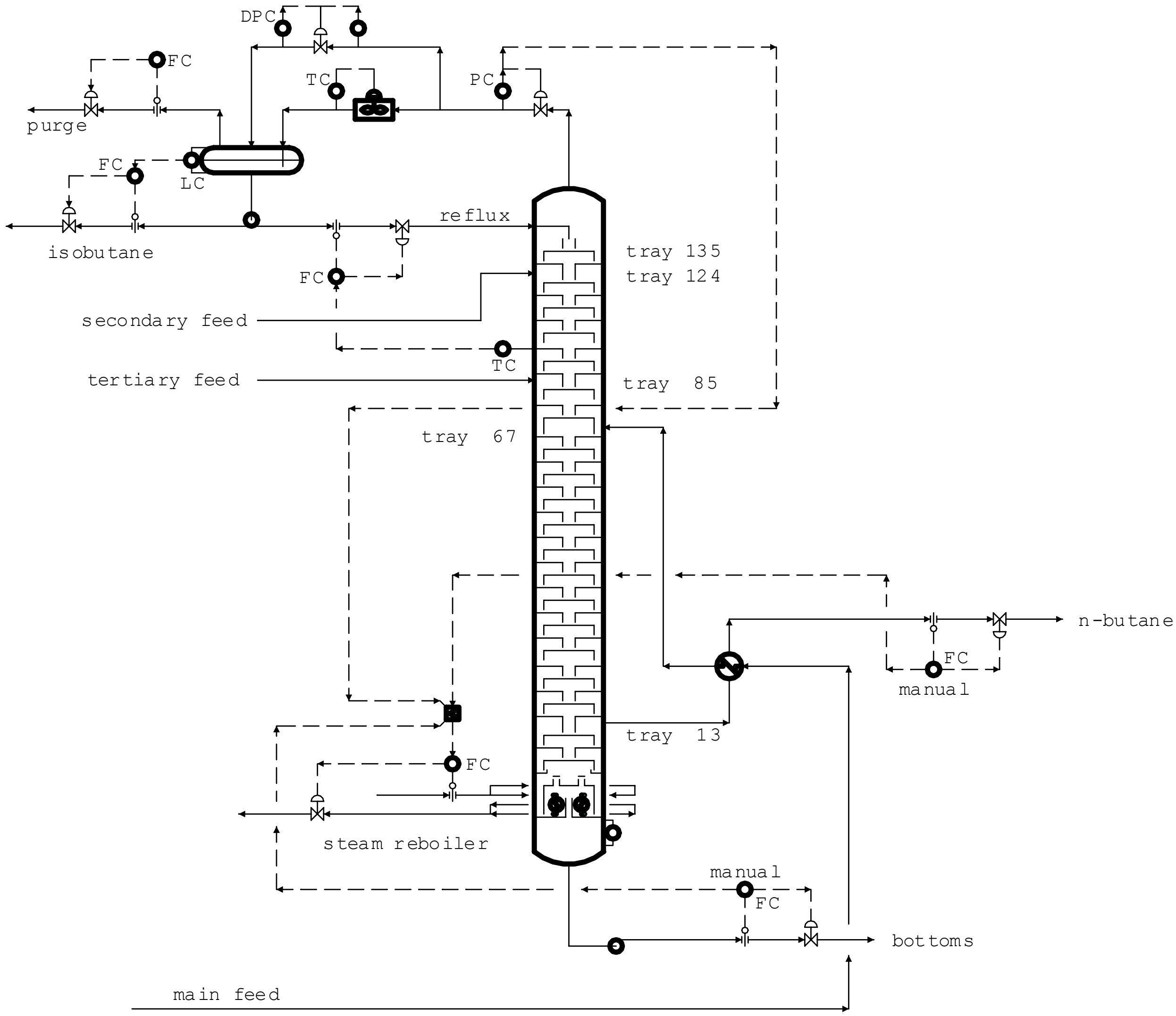



**Figure 23**
Control scheme derived by operator experimentation

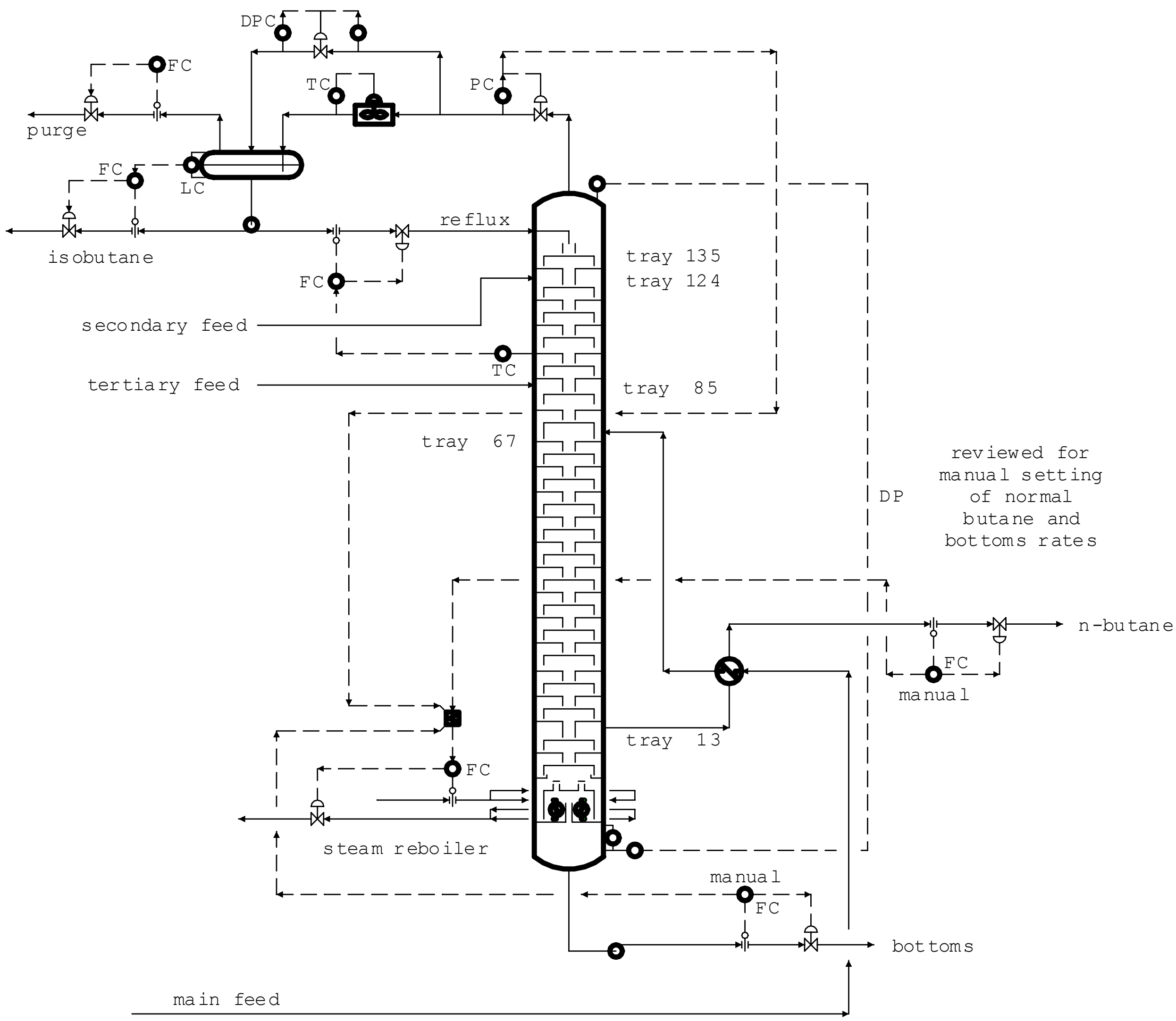



**Figure 24**
DP cell added in attempt to improve bottoms level control

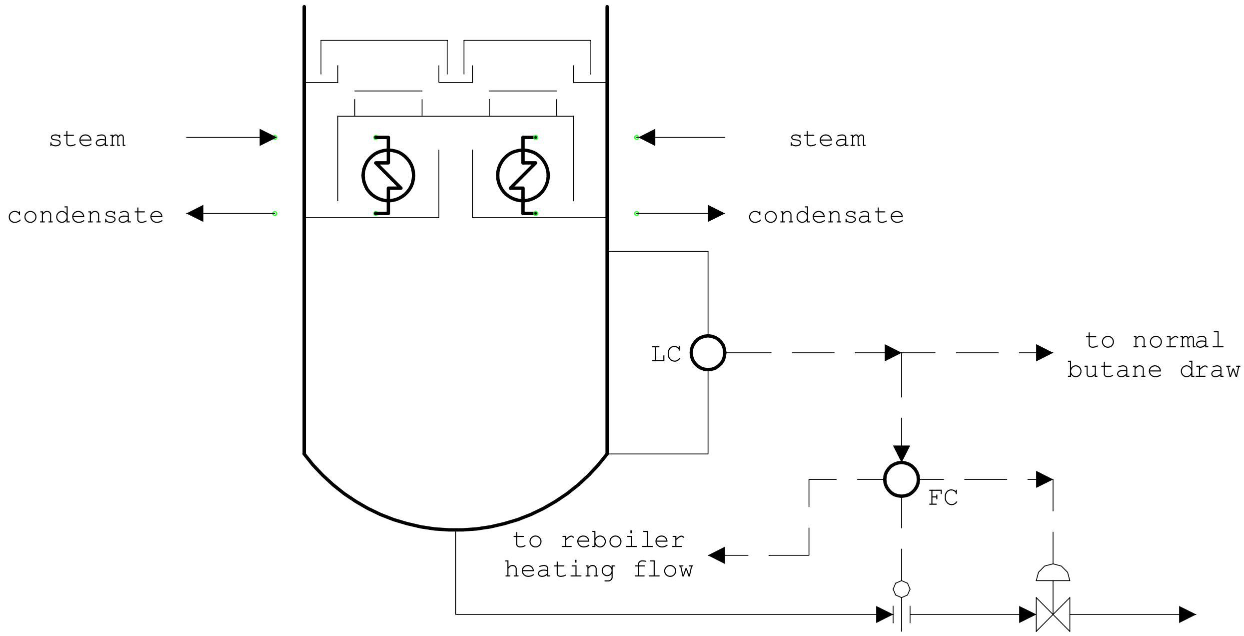



**Figure 25**
Deisobutanizer boot with stab-in reboiler expected (correct) configuration

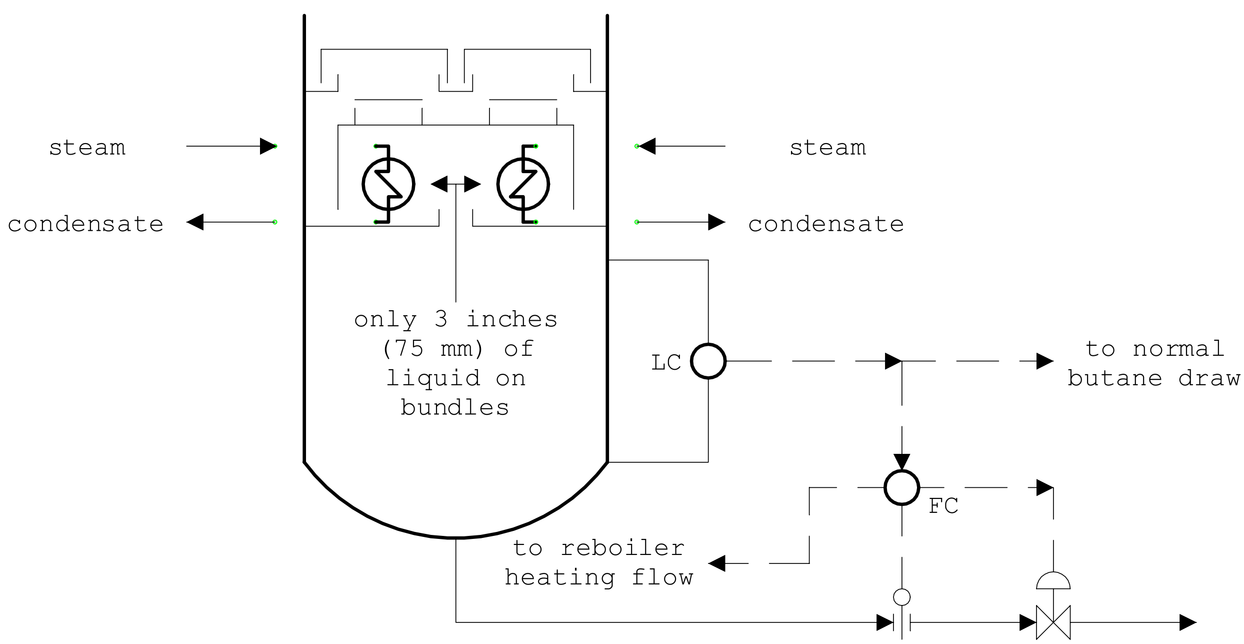



**Figure 26**
Actual deisobutanizer boot with stab-in reboiler configuration

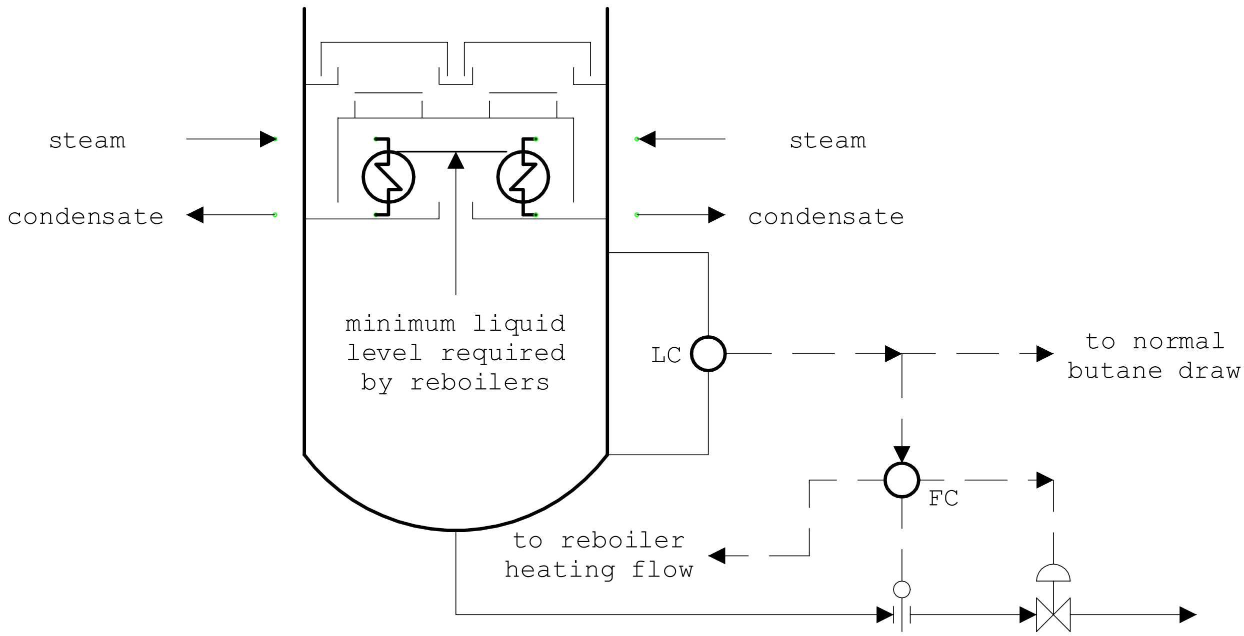



**Figure 27**
Minimum liquid level required to make reboilers work

***MALFUNCTIONING COLUMN LEVEL CONTROLLER***

***Background***

After a revamp of a commercial petrochemical column the column was started up and lined out. A test run was performed to evaluate the column's performance. The test run lasted two days. The column was pushed to 116% of original design during the test run. Table 3 gives the before revamp operational data of the column. Table 4 gives the after-revamp test run operational data of the column.

**Table 3**
Before revamp

| Stream | Column Feed | Overhead Product | Bottom Product | Reflux Stream |
|---|---|---|---|---|
| Flow, lb/hr | 155,520 | 150,500 | 5,020 | 72,240 |
| Temp, ºF | 270 | 150* | 250 | 150 |
| Pres, psia | 65 | 24 | 34 | 60 |
| Product wt % | 98.67 | 99.60 | 70.7 | 99.60 |
| Impurity wt % | 0.7515 | 0.0750 | 21.1 | 0.0750 |
| Inert material wt % | 0.5785 | 0.325 | 8.2 | 0.325 |

*Product stream leaving the condenser

**Table 4**
After revamp

| Stream | Feed | Overhead Product | Bottom Product | Reflux |
|---|---|---|---|---|
| Flow, lb/hr | 180,400 | 173,00 | 7,420 | 87,400 |
| Temp, ºF | 261 | 150* | 245 | 123 |
| Press, psia | 65 | 24 | 37 | 60 |
| Product wt % | 98.60 | 99.56 | 76.2 | 99.56 |
| Impurity wt % | 0.781 | 0.091 | 16.9 | 0.091 |
| Inert material, wt % | 0.619 | 0.349 | 6.9 | 0.349 |

*Product stream leaving the condenser

The objective of the test run was to operate the column at 110% of design rates for the two-day test period. At the midpoint of the first day of the test it was decided that the column could be run harder. Therefore, the test limits were raised. The feed rate to the column was raised to 116% of design.

The after-revamp overhead product purity was slightly less than the requirement of 99.6%. An overhead product purity of slightly less than 99.6% was acceptable as long as the percent impurity in the overhead product stream was less than 0.1%. Nevertheless, with the overhead purity being so tight it made the operation of the column extremely difficult.

The product concentration in the bottoms product stream was less than before the revamp. Under normal conditions, the product concentration would be 70% in the bottom product stream. During the test run, the bottom product purity observed was around 75% to 76%. The purity observed in the bottom of the column presented a problem to the operation department.

The bottom product purity specification is 70%. In addition to those variances, the overall pressure drop measured in the column was 12 psi (83 kpa). This was somewhat higher than the calculated overall pressure drop of nine psi (62 kpa). With the pressure drop across the column at 12 psi (83 kpa)the column did not operate properly.

There was a noticeable loss of efficiency. Usually when the pressure drop is high and there is a noticeable loss of efficiency this would indicate that a flood condition exists. Before the revamp the column would normally operate with a nine psi (62 kpa) pressure drop across the column.

***Troubleshooting***

A simulation was used to model the performance of the column. The vapor and liquid loadings generated by the simulation were used to evaluate the performance of the tower internals.

A gamma scan test further confirmed that the tower was at or very near flooding conditions. The trays below the feed were very near flooding conditions. The scan indicated that liquid was being entrained from one tray to another in the bottom of the column.

After analyzing the data collected from the test and the gamma scan it was found that there was a level controller in the bottom of the column that was out of calibration. The liquid level was even with the reboiler return nozzle. The high level resulted in liquid being entrained to the bottom tray, Figure 28. If the liquid level in the bottom of the column rises above the reboiler return nozzle vapor from the reboiler can entrain liquid into the first tray in the bottom of the column. The condition can lead to many column operational problems.

***Liquid Level Errors are a Major Source of Operating Problems***

Liquid level measurement errors are a major source of operating problems [12]. High levels cause many problems. These problems may include:

1. Tray or packing damage
2. Premature column flooding

***After the Problem Fix***

Once this problem was corrected the column operation improved. The column pressure drop went from 12 psi (83 kpa) to 10.5 psi (72 kpa). The product purity in the overhead rose above 99.6%. The impurity in the overhead did not change significantly. The decrease in pressure drop is an indication that premature flooding was taking place, Figure 29.

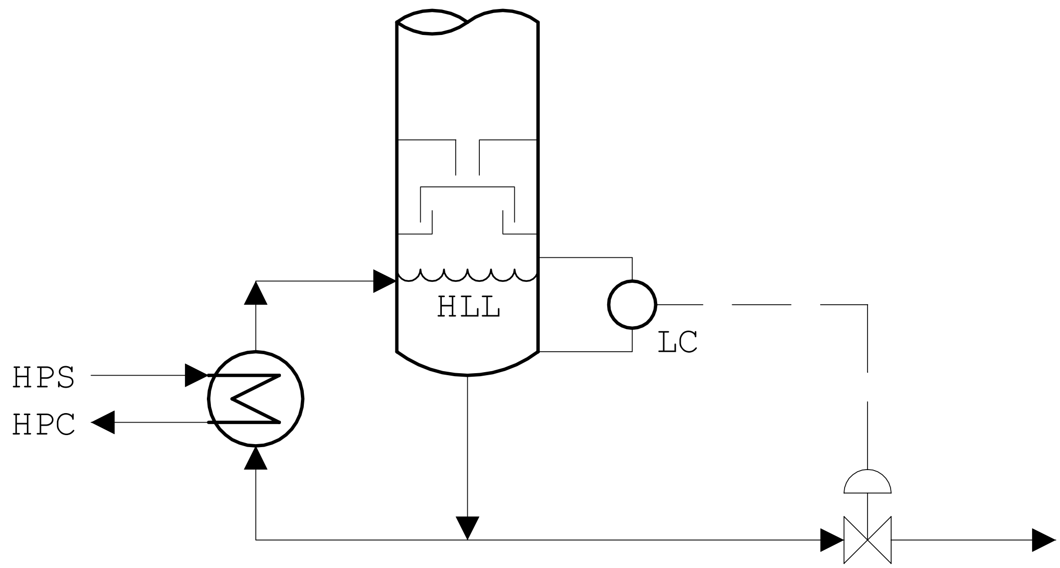



**Figure 28**
Liquid level at reboiler return

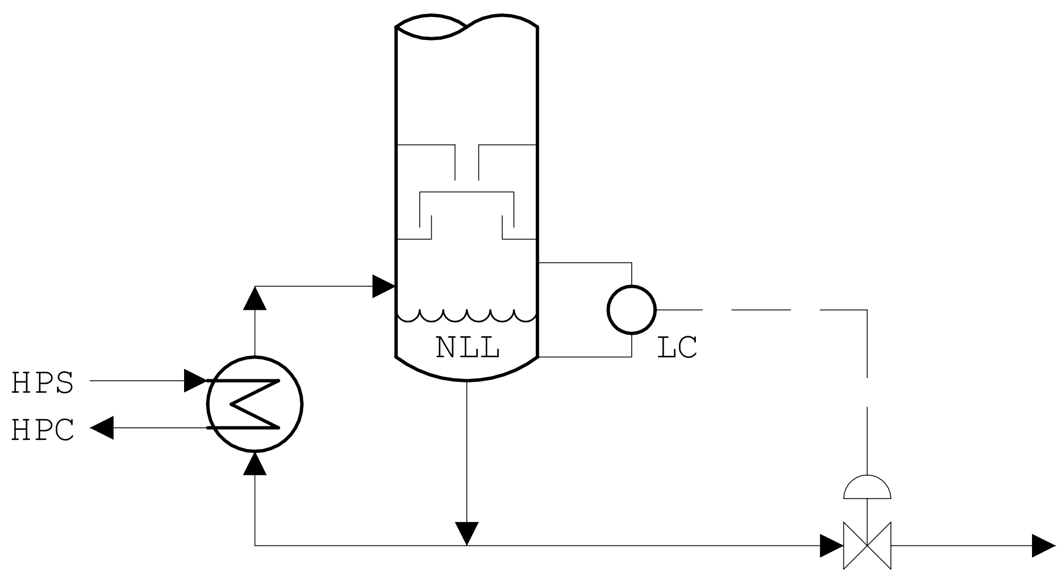



**Figure 29**
Liquid level normal

## *FOULED LIQUID DISTRIBUTOR LIMITS WASH COLUMN PERFORMANCE*

Water wash columns are often used in hydrocarbon systems to remove solvents or other hydrophilic material from a hydrocarbon stream. Figure 30 shows a water wash tower used in a petrochemical plant to remove excess reactant from a hydrocarbon stream. Hydrocarbon was the dispersed phase and water the continuous phase. Control used hydrocarbon inventory in the tower top to set the water bottoms rate and system pressure for the hydrocarbon overhead rate. Extraction was across four packed beds of random packing. A wire-mesh pad in the top acted as a water coalescer to reduce water carry over in the hydrocarbon.

The overhead product went to stabilization to remove light hydrocarbons generated in the upstream reactor. The bottoms product went to a distillation tower for reactant recovery from the water.

The plant had run seven years between start-up and the first shutdown. The wash column operated with no problems during this period. At the first shutdown, the hydrocarbon liquid distributor was rotated 180 degrees (Figure 31). The object of this change was to improve wash column performance by improving hydrocarbon distribution. The idea was that by having the hydrocarbon injected down, then having it rise, that the distribution of hydrocarbon to the packed bed would improve. Improved distribution would increase removal efficiency of the excess reactant. After the plant re-started, no improvement was seen with column performance.

The column operated effectively for approximately three years. At that point, intermittent downstream upsets started to occur in the downstream reactant recovery column. The problem was traced back to hydrocarbon entrainment in the wash water to the reactant recovery column. Even small amounts of hydrocarbons in water columns can cause dramatic problems due to two liquid phases inside the column.

As the run progressed, problems became more frequent and more severe. To prevent hydrocarbon entrainment in the wash water column, the water rate was dropped. Lower water wash rates reduced wash column effectiveness. Contaminants were left in the hydrocarbon feed to the stabilizer. Special facilities had to be added to post-treat the hydrocarbon product. Eventually, after 18 months, the water wash rate had to be restricted to 50% of the original water wash rates.

Various theories were proposed to explain the source of the problem. As part of the troubleshooting effort, a pressure gauge was placed on the hydrocarbon feed line to the unit (Figure 32). The pressure differential found between the feed line and the vessel showed a pressure drop across the distributor approximately five times the expected value. Further investigation showed that the hydrocarbon distributor had partially fouled during the first run of the plant.

The high pressure drop for the hydrocarbon flow was evidence of fouling in the liquid distributor. The partially blocked distributor had a higher than design hydrocarbon velocity out of the distributor. The high velocity decreases droplet size and increases the velocity of the hydrocarbon jet into the water. When pointed down, both factors increased the risk of entrainment of hydrocarbon in the water. Some fouling was tolerable, but after three years, entrainment began. Hydrocarbon entrainment caused operating problems. At the next shut-down the plant will return the hydrocarbon distributor to its original position.

Normally, the orientation of a liquid-in-liquid distributor has minimal to no effect on extraction unit performance. Process engineers should always be open-minded about changes to improve operation. However, they should check the basis for changes and make sure that any risk from the proposed change does not outweigh its benefits. In this case, a change caused an operating problem and no benefit was gained. Changes should be made based on fundamental understanding of the unit, not on a random basis.

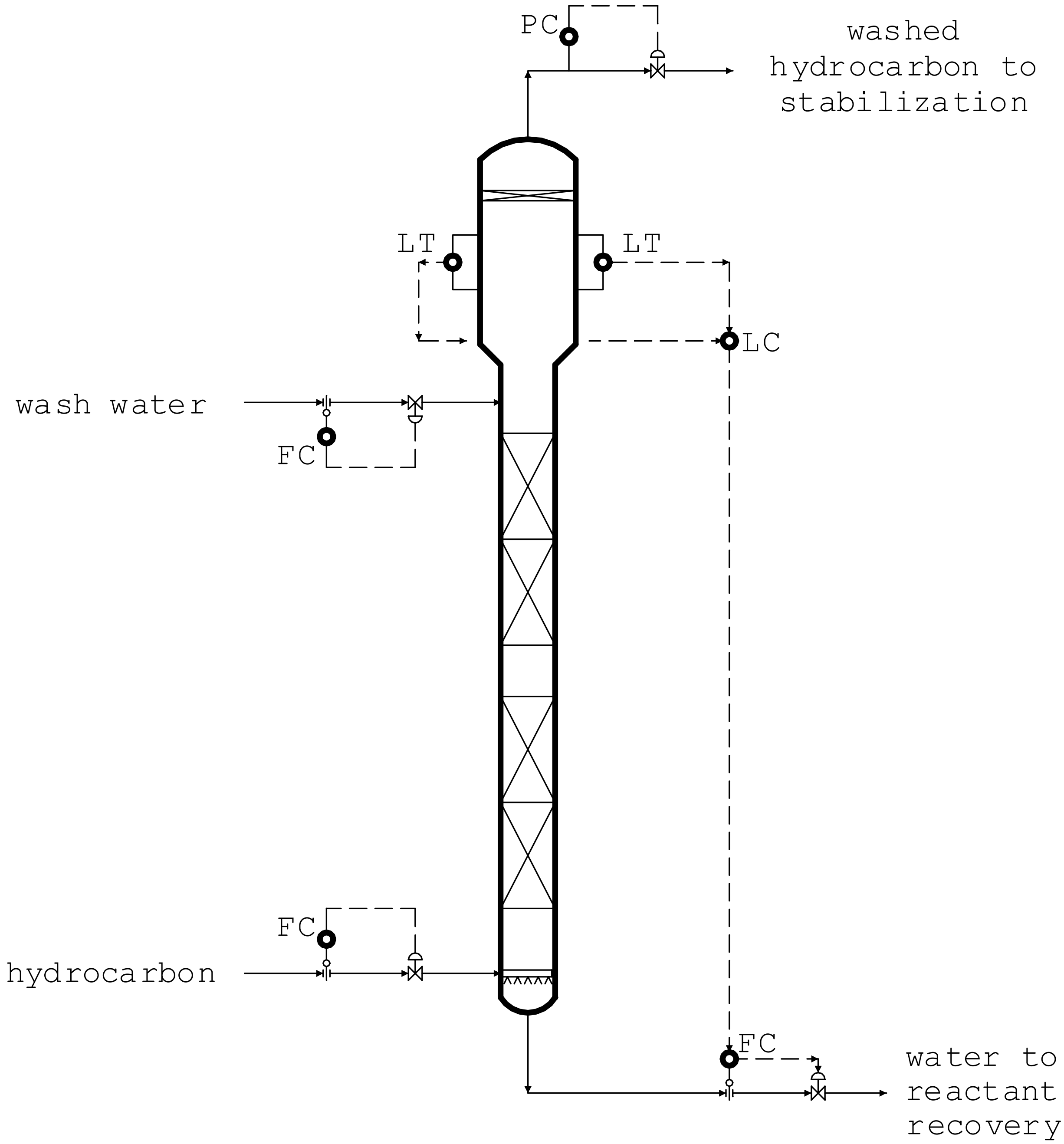



**Figure 30**
Water wash column

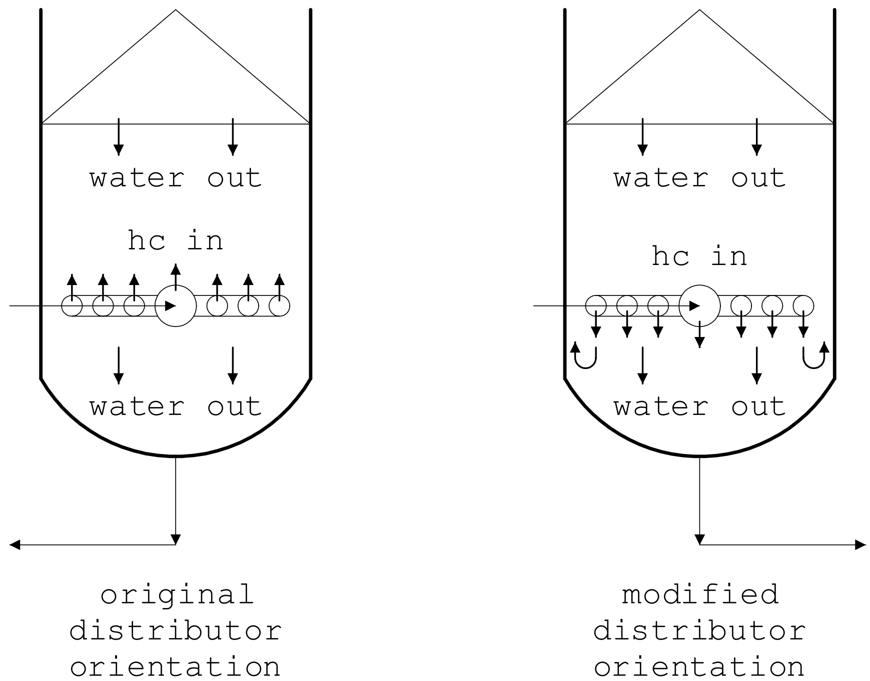



**Figure 31**
Rotated distributor

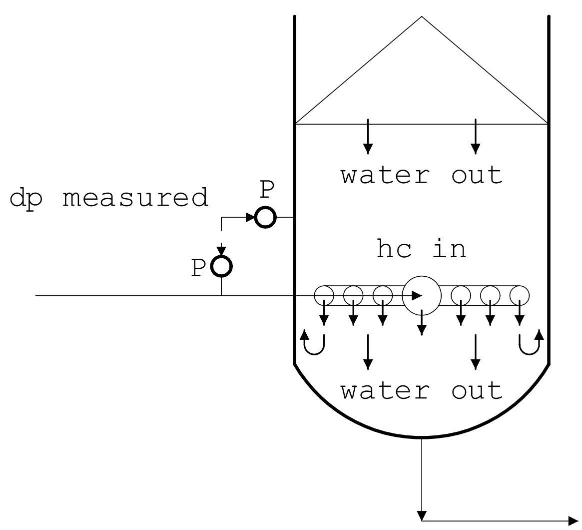



**Figure 32**
DP measurement

## *CONCLUSIONS*

Troubleshooting maintains plant profitability. Troubleshooting requires knowledge of the process, the equipment, chemical engineering fundamentals, good field technique and experience. The only way to learn to troubleshoot is by troubleshooting. Case studies have been presented to give ideas on many different distillation problems that can occur. Each includes information on the problem, troubleshooting approach, and solution.

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