Sour water strippers: design and operation

Simpler, more cost-effective sour water stripper designs can outperform more generally accepted designs in the refining industry

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In 1965, while working as a process design engineer with the American Oil Refinery in Whiting, Indiana, I designed a sour water stripper to remove ammonia, hydrogen sulphide and phenols from coker and FCC effluent sour water. Many years later, in 2012, I was working in two refineries in India, where I was troubleshooting sour water strippers in both plants. None of these three strippers had worked properly, in that the residual NH₃ in the stripped sour water was excessive. All three strippers suffered from low tray efficiency, as well as reboiler and feed preheater exchanger fouling. While conventional, all three stripper designs were needlessly complex. This excessive complexity contributed to their poor performance in the field. As a result of these deficiencies, the stripper bottoms were diverted to the effluent water treatment plant rather than being reused for their intended purpose as crude unit desalter wash water.

There are three general ways to design refinery sour water strippers, but only one of these three designs reflects correct process engineering principles. The two sour water strippers that I was troubleshooting in India both reflected common but incorrect process engineering design.

Purpose of sour water strippers
Refinery sour water originates largely from delayed cokers, hydrodesulphuriser reactor effluents, fluid catalytic cracking units and visbreaker fractionators. The main contaminants are NH₃ and H₂S. Sour water stripper bottoms are reused in two places:

- Wash water for the crude desalter on a once-through basis, to remove chloride salts that promote HCl evolution in the crude tower overhead condensers
- Make-up water for hydrotreater effluent recycle wash water to remove ammonia sulphide salts that plug the downstream condensers

When sour water stripper bottoms are used in the crude desalter, the NH₃ content should be about 10 to 20 ppm. Higher NH₃ levels interfere with crude unit corrosion control due to chlorides.¹ One hundred ppm of NH₃ is excessive for desalter operation.

When sour water stripper bottoms are used as make-up to the hydrotreater effluent wash water, an NH₃ content of a few hundred ppm is acceptable. After all, the recirculated wash water has over 10 000 ppm of NH₃, so the NH₃ content of the make-up water is irrelevant, as long as over 90% of the NH₃ in the sour water stripper column is stripped out.

Modern conventional stripper design
Figure 1 shows the current conventional design used at most locations. The operating conditions shown are typical of many refineries. Feed is heated by exchange with the stripper bottoms in E-1. The reboiler duty (E-3) is required for three purposes:

- Heat feed from 180°F (82°C) to the 250°F (120°C) bottoms temperature
- Generate internal reflux at tray 32
- Break the chemical bonds between water, NH₃ and H₂S.

¹ One hundred ppm of NH₃ is excessive for desalter operation.
Older conventional stripper

Figure 2 is based on older conventional sour water stripper designs. In this version, there are only 16 trays rather than 40. Field observations indicate that 15 or so stripping trays are sufficient to remove over 99% of the NH$_3$ from the sour water and an even greater percentage of the H$_2$S. Note that reducing the reflux drum temperature much below 165-170°F (74-77°C) results in salt plugging. The 190°F (88°C) reflux drum temperature is likely conservative (high) and results in increased water in the sulphur plant feed and thus more condensation in the sulphur plant feed NH$_3$ gas knock-out drum (and thus water that has to be recycled back to the sour water stripper).

The design shown in Figure 2 was common in the 1960s. However, it also suffers from the same heat balance drawbacks and needless complications as seen in Figure 1. In other words, a lot of equipment is added to generate reflux when no fractionation is required between the feed and overhead product. Again, the only purpose of the tower is to strip out the NH$_3$ and the H$_2$S.

Correct stripper design

In 1969, while working for the now vanished Amoco International Oil Company in the UK, I designed a sour water stripper that eliminated the unnecessary features of the unit shown in Figure 2.

Figure 3 shows the essentials of a correct sour water stripper design. Feed is brought in at ambient conditions (70-100°F,

![Figure 1 - Conventional sour water stripper design used in newer units](image_url)

![Figure 2 - Older conventional stripper design](image_url)

![Figure 3 - Correct stripper design](image_url)
21-38°C) from the sour water feed tank. To heat the feed from 90°F (32°C) to 250°F (120°C) requires about 16 wt% steam flow, or about 1.3-1.4lb of steam per gallon of stripper bottoms, which is close to a typical design stripping steam ratio for sour water strippers. The E-1 feed preheater, reflux pump (P-2) and the reflux cooler (E-2) shown in Figure 2 are all eliminated. How, then, does one know that the design shown in Figure 3 will work? Because it was built this way (at the Amoco refinery in Milford Haven, Wales, UK) in 1970, where it worked just as well as the conventional design shown in Figure 2.

**Two-stage sour water stripper**

Figure 4 shows a sour water stripper with a side draw-off. The partly stripped sour water is extracted from tray 8 and directed to the hydrotreaters for use as make-up water in the salt (NH₄HS) removal step of the reactor effluent. Completely stripped water from the sour water stripper bottoms is sent to the crude desalter. While apparently non-conventional, this design is just a direct application of the “lean, semi-lean amine regeneration design” that is often used in the natural gas industry to save energy (reboiler steam).² The design of such two-stage rich amine regenerators is fairly conventional technology. The objective is to produce a very low residual H₂S content in the lean amine, at the expense of a higher residual H₂S content in the semi-lean amine.

**Tray efficiency**

For one of the sour water
strippers in India, discussed earlier, the NH₃ in the bottoms product was about 200 ppm, even though the NH₃ in the feed was only 1000–2000 ppm. Refinery personnel felt, and correctly so, that the 200 ppm of NH₃ precluded the use of these sour water stripper bottoms as crude unit desalter wash water. Thus, the water flowed directly into the refinery effluent treatment plant.

The problem was not a lack of stripping steam or the excessive pH of the stripper bottoms, as refinery operators suspected. It is quite normal for the stripper bottoms to have a higher pH than the feed if stripping efficiency is bad. That is, the acidic H₂S is stripped out of the sour water more easily than the basic ammonia. The problem of excessive NH₃ in the stripped water effluent was low tray efficiency.

**Field observations**

To troubleshoot the stripping deficiency, I made the following field measurements:

- The delta P across the bottom 32 trays was zero, meaning no intact trays
- The delta P across the top eight trays was 7 psig (or about 16 ft of water, which equaled the vertical height of the tower above the chimney tray shown in Figure 1), meaning trays 33-40 are totally flooded.

I then advised as follows:

- Do not use valve or float trays. The valve caps stick to the tray deck and promote high delta P and flooding. Use a grid tray with a half-inch cap lift. Pro-valves are a good option that I have successfully used in fouling services
- Bolt the trays to the tray ring or use back-to-back trays with shear clips⁴ to avoid tray failure due to pressure surges caused by high bottom liquid levels.

One acceptable tray design is sieve trays with half-inch holes, which seem to work fine. Avoid packed towers. They are subject to vapour-liquid channelling and poor fractionation efficiency due to sloppy installation, fouling or poor liquid feed distribution. When calculating the required hole area for the trays, do not forget that the vapour loads and hence the required tray hole area substantially diminish as vapour flows up the column from the reboiler to the feed tray.

A recent publication⁴ discussed the various methods available to calculate tray efficiency in sour water strippers:

- Murphree vapour efficiency
- Overall efficiency
- Mass transfer efficiency.

This article suggests that the reboiler steam rate may be reduced by roughly 20% if the number of stripping trays is increased from 18 to 32, based on a computer simulation using mass transfer tray efficiency. However, in reality, the tray efficiency for perforated tray decks (valves, sieves or grid trays) is mostly a function of:

- Tray deck levelness
- Weir levelness.

Decks that are out of level promote vapour channelling through the tray deck. Weirs that are out of level promote
liquid channelling across the tray deck. Either type of channelling causes loss of vapour-liquid contacting and thus a severe reduction in tray efficiency. Experience has shown that the 15 stripping trays shown in Figure 3 are adequate if sufficient steam is used to heat the bottoms effluent to the boiling point of water at the stripper bottoms pressure, provided that the trays are properly installed and designed correctly to prevent dumping and channelling.\textsuperscript{1-3}

**Use of caustic to improve stripping**

I was working on a smaller crude unit in Alabama during 2012 to improve desalter efficiency. Their problem was the periodic carryover of emulsified brine. The depth of the emulsion between the crude and the water phases in their desalter was excessive. The pH of the wash water from the sour water stripper was 8.5 to 9, even though the NH\textsubscript{3} content of the stripper bottoms was zero. Discussions with the operators revealed that they added NaOH to the sour water stripper feed to diminish the NH\textsubscript{3} content of the stripped water.

Caustic contamination of the desalter wash water is known to promote high emulsion levels. There are many references to this problem in NPRA Q&A sessions. With this source as support, I asked the operator to stop the NaOH addition to his sour water stripper feed. Two hours later, at the same reboiler duty, the NH\textsubscript{3} content of the stripped sour water had gone from 0 ppm to about 5 ppm and the emulsion layer in the crude unit desalter had been reduced enough so that brine carryover was no longer a problem.

Caustic addition to the sour water stripper feed will increase NH\textsubscript{3} stripping efficiency by reducing the solubility of NH\textsubscript{3} in water. This may be good practice if the stripper bottoms is flowing into the refinery effluent treatment basin. But, it is better to have 10-20 ppm NH\textsubscript{3} in the sour water stripper bottoms than to contaminate the desalter with NaOH, which promotes the formation of brine-crude oil emulsions.

**Reboiler corrosion**

One of the questions that I am often asked is what can be done to mitigate reboiler tube failures in sour water strippers and the consequent steam leaks into the stripped sour water?

One suggestion is to avoid the use of kettle reboilers or once-through thermosyphon reboilers in favour of circulating thermosyphon reboilers. That is, feed the reboiler directly from the bottom of the stripper and design for about a 10-20\% vaporisation rate (see Figure 3). This will reduce, but unfortunately not eliminate, reboiler tube failures.

There is a more helpful suggestion if one can meet the following two criteria:

- Stripped water flows to the crude unit desalter and/or the hydrotreater effluent wash water make-up
- External water is required to supplement stripped sour water at the above units.

The suggestion is to use open steam in the stripper bottoms and abandon the reboiler. This will increase the stripper bottoms flow by 10-12\%. But this will be offset by the diminished fresh water make-up rate to the desalter or hydrotreater effluent wash.

About 50\% of refineries use open stripping steam and have discarded their reboilers. Or they are operating with massively leaking reboiler tubes, which is much the same as open stripping steam.

Referring to Figure 3, for many refinery operations, the circulating thermosyphon reboiler shown can be eliminated and replaced with the same quantity of stripping steam, thus further simplifying the design of the sour water stripper installation.

**Conclusions**

Generally accepted sour water stripper designs in the refining industry are not necessarily correct. The application of basic chemical engineering principles such as the stripping factor can often lead to simpler, more cost-effective designs. Avoid the use of flutter cap-type valve trays in fouling services. Avoid preheating sour water stripper feed normally serves no purpose. Avoid the use of caustic if the stripped water is to be reused as crude unit desalter wash water. A refinery sour water stripper should have about 20 trays, not 40. It is best to use grid or sieve trays rather than a random type (rings) in the tower. If high turndown rates are required, use bubble caps rather than...
valve tray decks. For many locations, the bottoms reboiler may be replaced with open stripping steam.

References
2. Lieberman N, Troubleshooting Natural Gas Processing, PennWell.

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