ADVICE FROM AN OLD-TIMER

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Process engineers, such as myself, who are approaching retirement or are at an age where "you can see it from here," probably received their undergraduate training in the 1950s. As we consider the prospect of retirement, or edge into it through part-time work or consulting jobs, it is interesting to consider how we differ from today's graduate.

There is little question that today's graduate is better equipped with the "tools of the trade" and better prepared to be immediately useful. But it can be argued that my generation spent an apprenticeship doing hand calculations that were more productive. For example, hours, if not days, were spent checking a heat exchanger design by hand, giving us a better feel for the variables involved than doing computer iterations would. But time soon evens everything out.

So, what can my generation pass on to new process engineers? Perhaps some rules-of-thumb that have been useful, or some guidelines for good practice, or some judgmental discernments that we have learned through unfortunate experiences. What follows are some of the rules and guidelines that have proven useful to me over the years.

STREAM EFFICIENCY

The stream efficiency, or annual on-stream operating time, is a key factor in successfully operating any chemical plant. Plants are designed to run on gallons per minute, or pounds per day, or barrels per stream day. But cash is generated and investors are rewarded from tons per annum, or pounds per year, or barrels per calendar day.

A key number to remember is 8760—the number of hours in a year. In a perfect world that is error and maintenance free, a plant would produce in a year 8760 times what it could produce in an hour. But allowing for a two-week annual maintenance shutdown and an unscheduled outage of one day a month, the operating hours in a year are actually 8136, for a stream efficiency of 93%.

In actual practice, a promise of more than 8000 operating hours in a year (or a stream efficiency of 91%) is highly suspect. Oil refineries that are well run and well maintained show that stream efficiencies in the 90s are difficult to achieve. The refining industry as a whole probably operates with a stream efficiency in the mid-to-high 80s.

Projected high stream efficiencies often stem from massaging the numbers to improve a project's economics, and "name plate" capacities are probably derived from a 72-hour test run, or whatever was contractually agreed upon at the beginning. Remember, though, that the real test run is the 8000-hour test. One should use 8000 operating hours per year as the goal, even when giving appropriate consideration to feed outages, power interruptions, changing product grades, etc.

ECONOMIC ANALYSES: TOP LINE VS. BOTTOM LINE

There are many excellent guides to preparing economic projections for a new project. For the most part, these guides focus on developing the "bottom line," *i.e.,* net profit, cash flow, payback, etc. That is what owners and investors want to see.

Much useful information, however, can be obtained from an analysis of the "top line," *i.e.,* the total sales generated. One should look at sales generated per dollar invested in much the same way as stock analysts look at a company's sales-per-share. How many times per year the sales "turn over" the capital invested can lead to a good appreciation of a project's risks and rewards.

Assume annual sales per invested dollar are substantially greater than one; unless the project is the proverbial license

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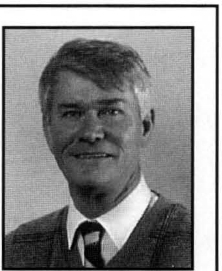

to print money, the margin per sales dollar is going to be thin. Examine the margin carefully. How firm is it? A small reduction in the margin could eliminate any profit. Is the operation a value-added one, such as refining crude oil where the margin is protected by a direct link between raw material and product prices? If a large volume of raw materials and products is involved, has sufficient attention been paid to materials-handling factors? Concentrate on the cost factors involved in the project.

HEAT EXCHANGERS

Without question, the most common mistake made in specifying heat exchangers is made by the conservative engineer anxious to supply adequate equipment who specifies too much area! By a unit that is too big, fluid velocities are reduced to less that 3 feet/second, transfer coefficients fall, and deposits build up in stagnant zones. Performance is poor and even, in some cases, inadequate.

Three feet/second is the absolute minimum velocity, shell or tube side, that should be considered. Providing the head required is a small price to pay for good heat exchanger operation.

A rapid way to estimate the number of tubes in a shell and tube exchanger is with the formula

$$N = C(L/P)^2$$

where

N = number of tubes
P = tube spacing, inches
L = "outer tube limit," inches
C = constant, 0.75 for square pitch, 0.86 for triangular pitch

The "outer tube limit" is 5/8 in. less than the shell diameter for fixed tube sheet or U-tube construction, and 1 1/2 in. less for floating head construction.

Assume 20-inch nominal shell diameter (19.2-inch inside diameter)
16-ft tube length, fixed tube sheet, 3/4-inch tubes on 1-inch square pitch
$N = 0.75\ (18.575/1)^2 = 259$ tubes
Area = (259)(16)(0.196) = 812 sq. ft.

or assume As above, but with 1-inch tubes on 1 1/4-inch triangular pitch
$N = 0.86\ (18.575/1.25)^2 = 190$ tubes
Area = (190)(16)(0.262) = 796 sq. ft.

This formula neglects tubes lost due to multipass construction, impingement plates, etc.

PUMPS

The old adage, "All pump problems are suction problems," still applies. Design for low velocities in suction lines: 0.5 to 1 foot/second for boiling liquids, and 1 to 3 feet/second for non-boiling liquids.

Vortex breakers are often omitted or ignored. Cross or flat plate baffles, with a width of 2 to 4 times the nozzle diameter and a height of one-half the nozzle diameter are effective vortex breakers.

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PLOTS AND COUNTERPLOTS

A company I once worked for operated some acetylation kettles in which sheets of cellulose (wood pulp) were acetylated with acetic anhydride. The kettles were jacketed with a recirculating brine to remove the heat of reaction. Brine circulation was controlled manually to prevent rising temperatures from degrading the cellulose and falling temperatures from reducing the reaction rate.

Varying temperatures, humidities, and time in storage of the cellulose resulted in varying moisture levels in the cellulose. This resulted in varying reaction temperatures and the recurring question—were we over- or under-controlling? Were we looking at random noise when the temperature wandered, or had reaction conditions actually changed?

Routinely, we plotted kettle temperatures to see to what extent we were diverging from set point. Next, we initiated a new plot for every five minutes, showing the cumulative extent to which temperature diverged from the set point. If the temperature was cycling randomly around the set point, this second curve would make an exaggerated cycle but would return to zero (see Figure 1).

If, however, reaction conditions had changed and a new equilibrium temperature had been established, the second plot would rapidly indicate this by going outside any bound-

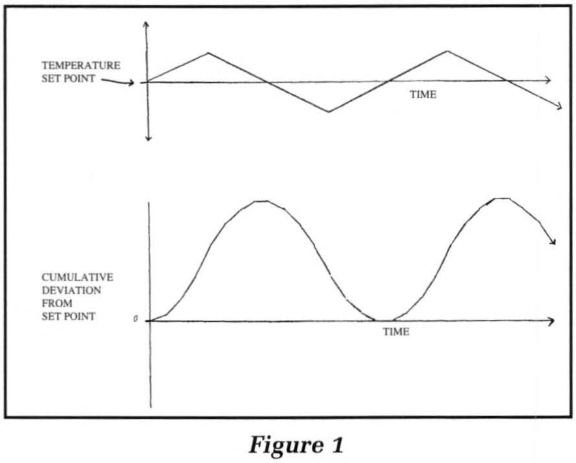

Figure 1

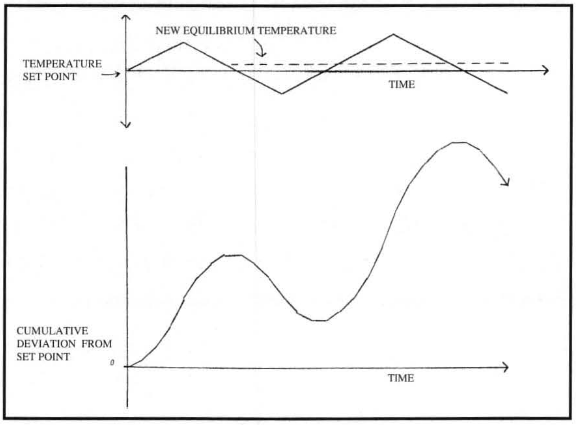

Figure 2

ary limits (see Figure 2). Boundary limits can be readily set after some operating experience.

The same kind of plot can be useful in any situation where you want to establish if a status quo situation or an existing trend line had been breached. They can be used alongside moving average plots in financial quotations.

Another plot I have found useful is the probability paper, both normal and logarithmic. A probability paper is most suited to record a series of events distributed around a mean where one wants to design for a certain fraction of the occurrence of the events. These events could be temperature

TABLE 1

Daily Truck Loadings		*Daily Truck Loading in Ascending Order*	
# of Trucks Loaded	**Day number "M"**	**# of Trucks Loaded, in order**	**[M/(N+1)]**
60	1	28	0.048
40	2	30	0.095
85	3	34	0.143
30	4	38	0.190
67	5	40	0.238
46	6	40	0.286
60	7	42	0.333
42	8	45	0.381
90	9	46	0.429
51	10	50	0.476
53	11	51	0.524
62	12	53	0.571
34	13	60	0.619
73	14	60	0.667
80	15	62	0.714
50	16	67	0.762
38	17	73	0.810
28	18	80	0.857
40	19	85	0.905
45	20 (N)	90	0.952
53.7 Avg.			

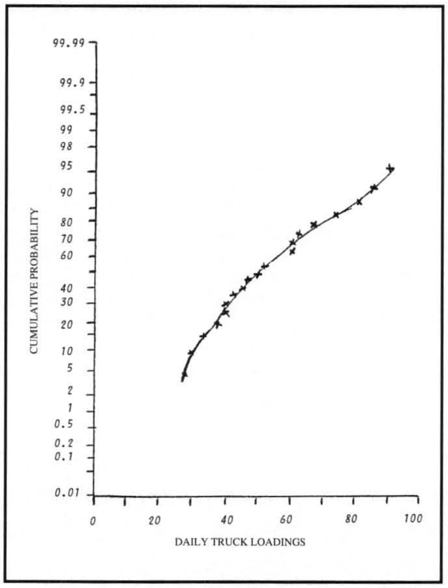

Figure 3

or wind levels, ship arrivals, river flows, etc. As an example, consider daily truck loadings at a terminal. The number of trucks per day is tabulated by day "M" in Table 1. Next, arrange them in ascending order and divide "M" by (N+1), where N is the total number of days, and plot "M"/(N+1) vs the number of trucks/day, or probability paper. The data aligns reasonably well (see Figure 3).

To satisfy the loading demanded four days out of five, or 80% of the time, inventory would be required to fill 70 trucks. If it was desired to satisfy the loading demanded nine days out of ten, inventory would be required to fill 85 trucks.

LINE SIZING

The guideline "1 pound per 100 feet" pressure drop can serve to size lines in a wide variety of situations. The tables in "Cameron Hydraulic Data," based on the Williams and Hazen formula, form a conservative standard. Two points worth mentioning here are:

1) In long, large-diameter lines, hold the velocity down to a walking pace of 4-5 miles per hour, or 6-7 feet per second.
2) 1 1/2-in. sch. 40 pipe is the smallest size that will span 15-to-20-foot pipe racks without intermediate support. Operators often climb on piping to take readings, etc., so 1 1/2-in. pipe is the smallest pipe size that should be used for routine use.

These are some of the guidelines and rules-of-thumb that have been of value to me during my career. Perhaps this article will prompt other "senior" process engineers to share some of the experience they have gained during their careers. We seniors owe a lot to a profession that has rewarded us well, both personally and professionally. ❒