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# Trading silicon for carbon: how to reduce energy usage through automation

## The average plant can conservatively achieve 15% energy savings through this technology

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Energy is the largest variable operating cost after raw materials for most of the process industries and its efficient use is key to sustaining profitable operation. Natural gas is the most common incremental fossil fuel and its general increase in price and in price volatility over the past few years are well known with most experts projecting these trends to continue in the future. Table 1 gives typical specific energy usage (Btu/t product) for common processes[1] and the value of a 10% energy reduction in terms of increased financial operating margin at an energy price of $7 per million Btu (mBtu). This value is a significant portion of the total operating margin for most of these processes.

In addition to the direct energy price, it seems likely that the US will eventually adopt some regulations regarding greenhouse-gas emissions. If the regulations in other countries are a guide, these may take the form of a "cap and trade" on $CO_2$ emissions.

This will place an increased value on energy reductions since these reductions can be used to offset increases in other areas or can be sold under a cap-and-trade system. If the $CO_2$ reductions are valued at $20/ton, which is at the low range of recent prices in Europe, then reductions in natural gas or equivalent light hydrocarbon fuel usage would have an additional value of approximately $1.3/MBtu. This would add roughly 20% to the value of the energy savings above.

To identify energy savings it is important to understand how energy is normally used. In larger process sites energy distribution can be quite complex (Fig. 1). External purchased energy can be

**TABLE 1. Typical processes[1] specific energy usage**

| | Net process energy usage, MBtu/t product | Value, 10% energy reduction, $/ton |
|---|---|---|
| Petroleum refining | 4.4 | 3.1 |
| Integrated pulp/paper mill | 29.0 | 20.3 |
| Cement production | 7.9 | 5.5 |
| Chemicals | | |
| Ethylene (naphtha feed) | 15.0 | 10.5 |
| Polyethylene | 3.3 | 2.3 |
| Polypropylene | 2.0 | 1.4 |
| EDC/ VCM/ PVC | 10.2 | 7.1 |
| EO | 3.8 | 2.7 |
| EG | 4.5 | 3.2 |
| Ethylbenzene | 2.9 | 2.1 |
| Styrene | 38.8 | 27.2 |
| Chlorine | 25.8 | 18.1 |
| Ammonia | 10.1 | 7.1 |

Source: Reference 1

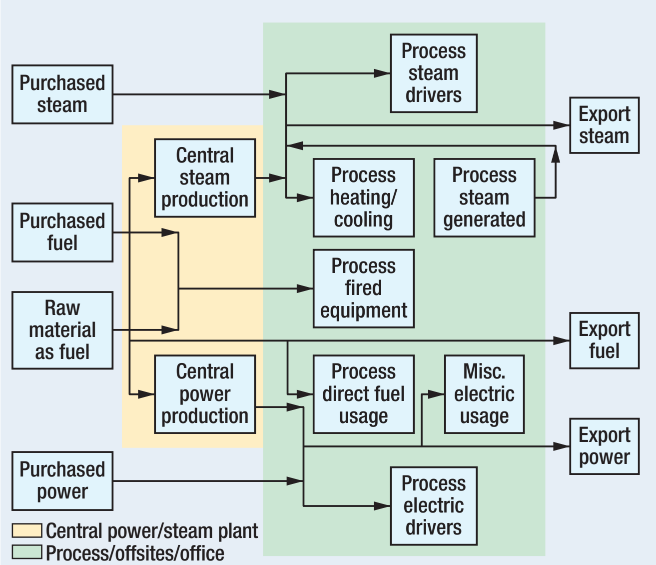


**FIG. 1** Typical industry energy supply/usage sources and uses.

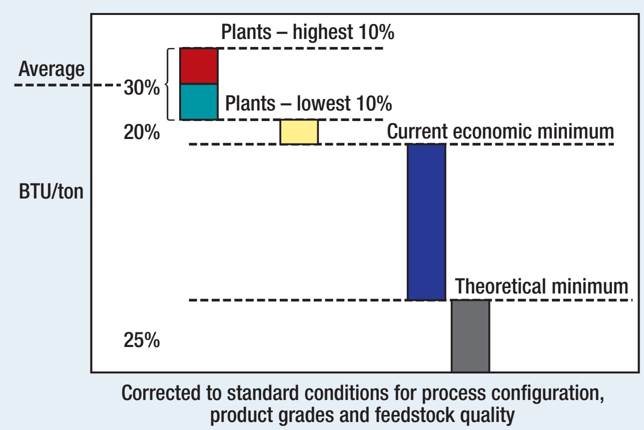


**FIG. 2** Typical process energy saving opportunities.

### TABLE 2. Typical North American process industry energy supply/usage distribution

| | Description | Approximate percentage of total input energy (Equivalent Btu basis, including losses) Chemical plants | Oil refineries | Pulp and paper mills |
|---|---|---|---|---|
| | *Energy inputs* | | | |
| 1. | Purchased steam | | 1 | |
| 2. | Purchased fuel | 63 | 25 | 34 |
| 3. | Raw material as fuel | (i) | 64 | 42 |
| 4. | Purchased electricity | 37 | 10 | 24 |
| | *Central utilities* | | | |
| 5. | Central steam production | 32 | 34 | 62 |
| 6. | Central Power production | 4 | 5 | 6 |
| | *Process energy usage* | | | |
| 7. | Process steam heating and steam drives | 22 | 19 | 39 |
| 8. | Process direct fired equipment | 27 | 52 | 5 |
| 9. | Process direct fuel consumption | | 6 | 4 |
| 10. | Process electric drives | 22 | 15 | 29 |
| 11. | Process miscellaneous electric usage | 18 | | |

Notes: (i) Included in purchased fuel above

in the form of steam, fuel, electricity or some portion of the raw material that is diverted to direct fuel use.

There can be centralized power or steam production that consumes purchased fuel or raw material. Typically, there will be a variety of users in the process and offsites area including fired heaters; steam and electric drives; and miscellaneous hydraulic, heating and cooling equipment—often supporting distillation/fractionation operations. Energy recovery in the process can yield supplemental steam and power production. There will be a variety of miscellaneous electricity users including the office, shop and warehouse facilities. There can also be export of energy to other plants or external consumers.

The relative percentages in each of these areas, based on overall usage, will vary significantly among different types of plants (Table 2).[2] The percentages are based on an equivalent Btu basis including losses. For purchased power, this means that "average" primary electricity production inputs of fossil fuels per kWh are considered along with allowances for transmission losses to calculate the Btu requirements rather than just the Btu equivalent of the plant fence-line-delivered power. The process fired equipment percentage includes ambient losses as well as process heat.

### TABLE 3. Potential energy investments

| Potential energy savings | Investment cost (time to implement) Low | Medium | High |
|---|---|---|---|
| High | | | Cogeneration<br>Process redesign<br>Replaceing low-efficiency process equipment |
| Medium | General improvements in automation<br>Energy KPI monitoring<br>Operating procedure changes | Advanced control/optimization<br>Site energy<br>Management systems<br>Flare reduction programs | |
| Low | Increased insulation<br>Steam trap/leak management<br>Enhanced exchanger maintenance<br>Increased condensate recovery | | |

Overall, the major energy consumers in the process industries will be the centralized power and steam production, process fired heaters, distillation/fractionation and steam/electric drives.

For specific plants of the same process type, there is still a wide variation in specific energy usage even after correction to standard conditions for feed quality, product rate/composition and process configuration (Fig. 2). Surveys, including those by the author, repeatedly show wide gaps between the most efficient plants and the least efficient with a 30% spread common between the mean of those with the highest 10% usage and the lowest 10%. This variation is primarily due to the equipment age with older plants often having less-efficient equipment and less heat integration than newer plants. Investments in energy reduction have to be economically justified and the proper level for these investments will depend on the future projected energy costs (including greenhouse gas effects) and the cost of the investments. For each plant there will be a current economic minimum energy use that will vary with projected energy costs. However, there are often additional savings obtainable for even the most efficient plants and 20% additional potential savings can be considered typical for current economics.

Even with these investments most plants would be well away from theoretical minimum energy use which is typically about 25% of current average levels. This would imply that the average plant could conservatively target 15% energy savings to bring its operation to benchmark levels and 35% to reach the current economic minimum.

Those with higher-than-average consumption would have even larger potential benefits. Many plants with comprehensive energy programs have indeed done much better.

A large number of possible investments can be made to reduce energy with differing costs and impacts (Table 3). These range from low-cost/relatively low-impact programs such as reducing

**TABLE 4. Potential automation energy saving strategies**

| Reduce process energy demand | Reduce energy supply costs | |
|---|---|---|
| | Increase internal utility production efficiency | Reduce external purchase costs |
| • Advanced control/optimization | • Improve combustion efficiencies—boilers, heaters, kilns | • Energy management system |
| – Heaters | • Steam management | – Internal vs. external |
| – Compressors | – Steam header management | (electricity generation optimization) |
| – Distillation | (minimize venting, letdown, pressure) | – Electricity purchase optimization |
| • Improved basic control loop performance | – Boiler allocation | • Maximize cheaper fuel use |
| • Maximize process heat recovery/minimize losses | – Blowdown control | |
| – Maximize recovered steam | • Steam vs. electric turbine optimization | |
| • Minimize process recycle (including off-spec. product) | | |
| • Minimize process pressure drop | | |
| • Minimize waste/flare gases | | |
| • Minimize standby equipment | | |
| • Minimize overdrying | | |
| • Better low-production rate control | | |

steam trap and other steam leaks to installing cogeneration units with very high costs, long implementation times and correspondingly high impact. In this article the focus will be on automation investments. These generally fall in the low- to medium-cost range with savings that are typically midrange. As a result, the expected return on investment for these programs can be quite high. They can also typically be implemented in a relatively short time providing quick payback.

Automation can affect energy use in many ways (Table 4)—more than can be covered in any single article. Overall, they might be considered as a way to trade the relatively small amounts of silicon used in the automation computers and other electronic devices for the large amount of carbon burned to produce energy. Initially, the strategies can be considered in two categories: those that reduce process energy demand at constant production rate and those that reduce the supply costs. The latter category can be further subdivided into those strategies that increase internal utility production efficiencies and those that reduce external utility purchase costs. Within each of these categories there are multiple strategies that can be pursued. Advanced control and real-time optimization of processes can yield significant savings. Improving basic control loop performance can have a very high payback. Maximizing heat and steam recovery, minimizing pressure drop, and minimizing waste and recycle are all target areas for automation. Improved control of the steam boilers, power turbines and plant steam system are also likely high-priority areas.

Kenney's book[3] is a good introduction to the subject of process industries' energy conservation and Shinskey's book[4] on automation energy savings covers many important areas. In this article some selected topics from Table 4 will be covered that, in the author's opinion, deserve increased emphasis.

**Improved basic control loop performance.** Efficient and effective basic control loop execution at the plant is obviously essential to successful plant operation and to other functions, such as advanced control, that are controlling energy. Control loops are composed of a measurement element; an actuator, most commonly a valve; and an executed control algorithm. Improvements in each of these elements can lead to reduced energy usage. These improvements can often be made very quickly at relatively low investment costs.

**Measurement.** One of the first areas to evaluate for potential automation changes is improving the measurements of key plant energy-related variables in terms of accuracy, location, frequency and number. Accurate measurement of, and accounting for, energy flows is the first step toward controlling usage.

There are some specific measurement improvements that can have a significant impact. Many plants experience wide variation in fuel gas composition and the corresponding heating value. As discussed in previous papers[5,6] mass-based heats of combustion for standard light hydrocarbon fuel-gas components have much less variability than those on a volumetric basis. Even hydrogen, which is one of the major causes of volumetric heating value variability, has a ratio on a mass basis much closer to the other components than its volumetric equivalent. This suggests that fuel-gas control on a mass rather than volume basis will eliminate the effect of much of the composition-based variability and that conclusion is supported by the experience of plants that have adopted such control. With modern flow measuring devices that directly measure mass and also provide a gas density measurement, it is easy to implement such controls.

Hoglen and Valentine[7] discuss a specific example of the improvements obtained with mass measurement on a reformer for hydrogen production. These units may use as feed-site fuel gas, external natural gas or a mixture. Steam-to-carbon ratio control against short-term feed composition variation is important. Too little steam can reduce catalyst life while too much incurs extra energy costs. In older plants, this control is often done with orifice-plate measurement of the gas flow combined with a gas chromatograph or a mass spectrometer for composition determination. From experience a typical operating margin of 0.2 above the desired steam/carbon ratio target was used to allow for short-term composition fluctuations. Installing direct mass measurement of the feed showed a maximum error of 0.02 in the ratio in tests on the actual plant. A reduction in the operating margin ratio of 0.1 was stated to be worth approximately $500,000 per year in energy savings for an 80-mscd hydrogen plant with fuel valued at $6.50 per mBtu.

**Pressure-drop reductions.** Excessive pressure drop in plant equipment represents unnecessary energy usage. Current operating conditions are often different from those when the plant was

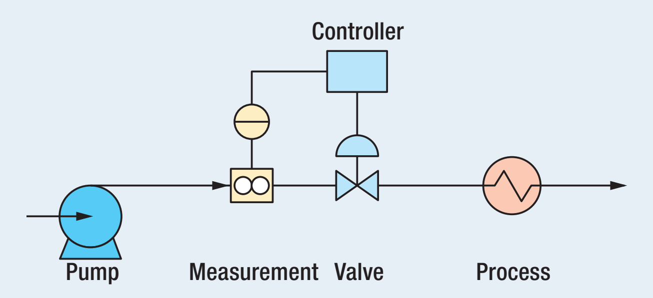


**FIG. 3** Typical configuration—pump and control valve.

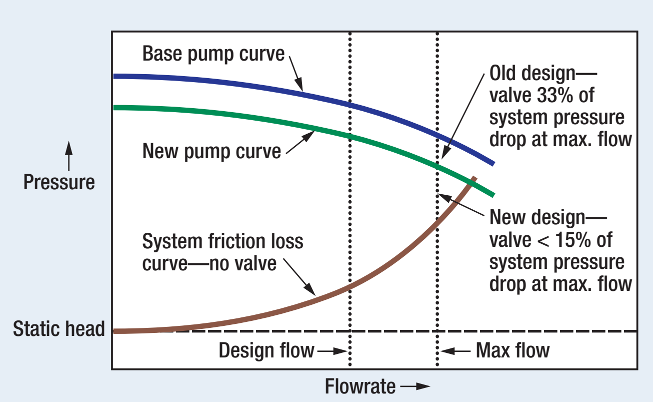


**FIG. 4** Reevaluating system pressure drop.

constructed or expanded and the plant hydraulic profile is different. In addition, the design may well have been done when energy costs were substantially lower. A systematic critical review of the profile can often reveal many areas where pressure drop and the corresponding energy input can be reduced. The energy savings can be substantial but they are usually composed of many small savings as opposed to a single large opportunity.

There are many pumps in the plant and it is worthwhile evaluating when they are oversized. Fig. 3 illustrates a typical centrifugal pump configuration with a control valve downstream controlling the flow into a process area.

In Fig. 4, a typical hydraulic profile for such a system is shown. For well-designed equipment in the past, the expected system process pressure drop as a function of flowrate would be estimated, without the valve, and the valve and pump would be chosen to provide a controllable pressure drop at estimated maximum flow conditions—often set at 33% of the maximum flow system pressure drop. The valve type and trim would be set to give good control over the operating range and ideally to linearize, as much as possible, the combined process gain over this range. However, with modern valves, and most particularly valves with modern digital positioners with direct position feedback, it is possible to have precise control at much lower pressure drops. To obtain the energy savings, it is necessary to reduce the pump impeller size that is normally possible for standard-size pumps. In the figure the effect of a 15% system pressure drop reduction at maximum flow is illustrated. For a 1,000-gpm pump with an electric drive at 100 psi head operating at 70% overall efficiency, the savings would be $6,000 per year with electricity priced at $0.075 per kwh. This would normally repay the impeller change

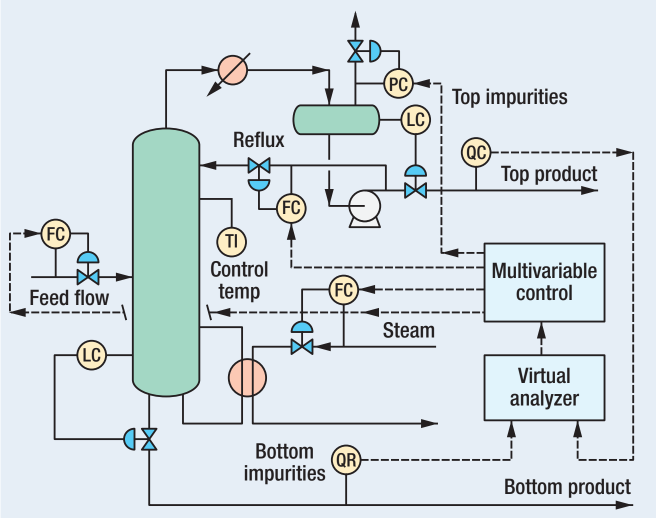


**FIG. 5** Typical two-product distillation column.

cost in a few months. In such a review it is, of course, necessary to evaluate startup and shutdown conditions for rangeability and determine if it is also necessary to change the valve trim.

In surveys at a number of plants, we have found many cases where valves are operating at less than 20% open at current maximum plant throughput. These should be a priority target for possible change since the savings can be proportionally much higher than those estimated previously.

**Distillation.** It is estimated that there are over 40,000 distillation columns in North America and that they consume about 50% of the energy usage in the refining and bulk chemical industries.[8] A typical two-product column is shown in Fig. 5. Improving energy efficiency in this unit operation is obviously an important target area for achieving overall energy savings.

The basic tradeoff is shown in Fig. 6. As reflux and the corresponding reboiler duty are increased, the separation in the column increases and the product value increases. But this increase is not linear. Initially, the reflux has a strong effect on the separation but the effect decreases as more and more is added. Energy costs, however, are approximately linear with reflux/reboiler duty. If the operating margin is then calculated as the difference between the

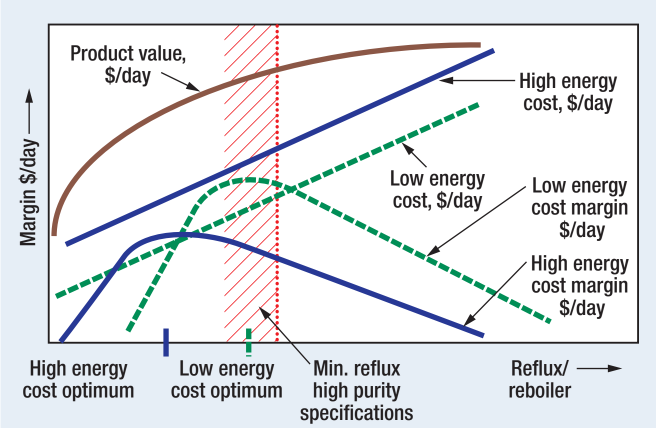


**FIG. 6** Column energy optimization.

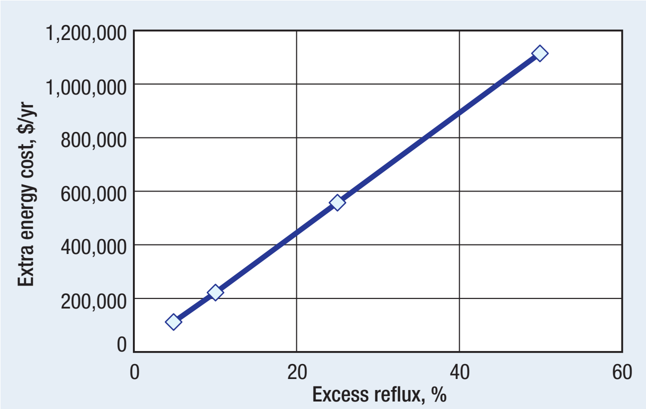


**FIG. 7** Excess reflux cost—20,000-bpd stabilizer column.

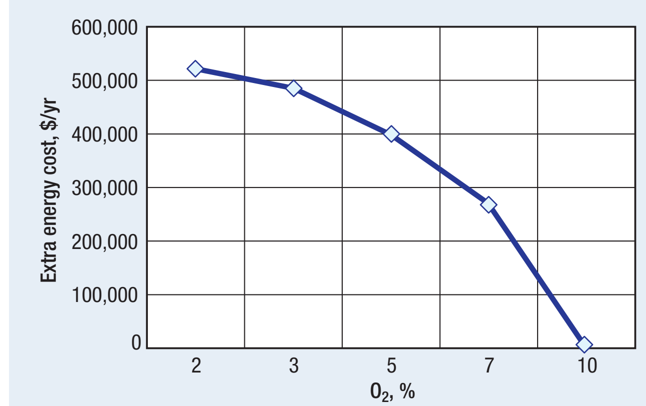


**FIG. 9** Excess $O_2$ cost—100-mBtu/hr heater.

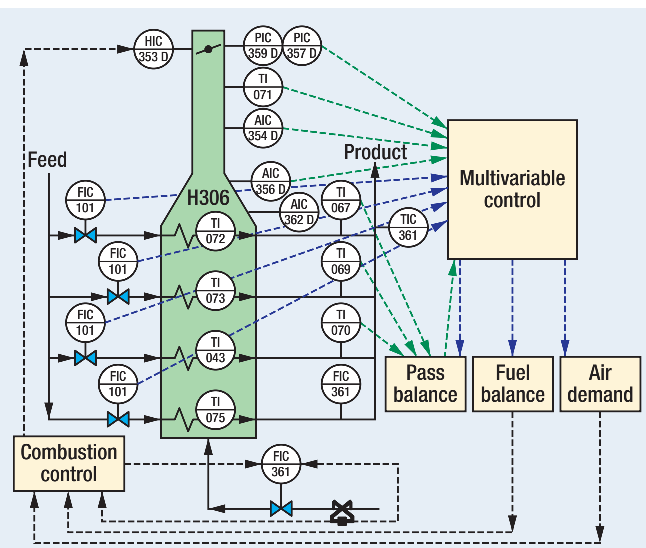


**FIG. 8** Process fired heaters

product value and the energy costs, there will be a maximum value and this is the optimum operating point. If energy costs increase while product values are constant, the margin curve shifts and the optimum moves to a lower reflux/reboiler point. The specific operating targets then should be a function of energy costs rather than a fixed number.

For the case of relatively high-purity specifications in both the top and bottom of the column, a minimum reflux ratio will be required for specific column tray numbers and configuration and feedstock composition. In the author's experience, many columns operate with reflux ratios far in excess of this minimum required for separation. Fig. 7 shows the extra costs incurred for a standard refinery stabilizer column (20,000 bpd) as a function of excess reflux (steam at $10/mBtu) over the minimum. Note the savings can be hundreds of thousands of dollars per year on a single column. Additional considerations concerning energy use in distillation columns are presented in reference 9.

As illustrated in Fig. 6, it is common to use multivariable control and virtual analyzers for distillation column control, particularly those that are large energy consumers or those with high-purity specifications. A large number of papers have been published on this subject with Blevins et al[10] providing a good introduction. Energy-saving results have been very positive with common savings in the 10–30% range.

**Process heaters.** Process heaters are major energy consumers in the refining and chemical industries (Table 2). Many of the opportunities for process heater energy savings are also applicable to steam boilers so proper combustion control is quite important. Heaters and boilers come in many different configurations ranging from very simple package boilers to exceedingly complex chemical cracking furnaces. They can have a single fuel with relatively stable composition or multiple fuels with highly variable compositions. Combustion air and draft control can be natural or forced with multiple control points combined with heat recovery equipment such as air preheaters. Process demand can be relatively stable or highly variable. The prescriptive automation level will depend on the equipment complexity and size with more complicated measurement and control justified on larger and more complex installations.

After ensuring safe and environmentally compliant operation, the next most important control objective is to meet the required heater load demand at the highest possible combustion efficiency, which normally translates into minimum excess-air control. In simple heaters/boilers with stable fuel composition, this is usually translated into air/fuel ratio control based on a load characterization curve. However, as shown in Bussman et al,[11] ambient condition changes can have a surprisingly large impact on combustion conditions, enough to move from full to partial combustions conditions under limited ambient changes. When fuel composition is varying as well, the problem of proper combustion control becomes even more complicated.

In Fig. 9, the cost of poor $O_2$ control is illustrated for a 100-mBtu/hr heater with a 400°F stack temperature rise above ambient and fuel costs at $7 per mBtu. Savings through improved control can be several hundred thousand dollars a year for a single heater.

As shown in Fig. 8, multivariable control is an option and is well-justified for larger heaters with varying loads and varying fuel composition. Flue-gas $O_2$ analyzers are commonly installed on larger heaters with CO analyzers being less common. In the past, CO analyzers were often problematic because of maintenance requirements. However, a new generation of analyzers has appeared with much higher reliability and lower installed costs. Fig. 10 shows actual CO readings from an industrial heater. Even with relatively stable $O_2$ readings, there are significant

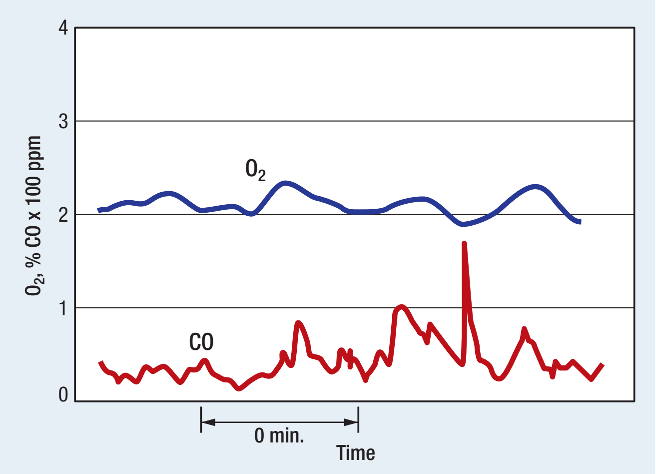


**FIG. 10** Carbon monoxide versus oxygen in combustion.

CO excursions indicating incomplete combustion and less than ideal efficiency. To maintain highest-efficiency operation, direct CO measurement and incorporation into the control scheme is recommended.

**Overall plant control.** In Fig. 11, a simplified but representational process plant configuration is shown. There is a reaction section with energy input and a separation section, also with energy input. In the reaction section, feed is converted to a high-valued product, $P_2$, and a low-valued byproduct, $P_1$. They are separated in the separation section along with unreacted feed that is recycled back to the reactor. Again, typically, as the "per pass" feed conversion is increased, selectivity, which is the ratio of high-valued product to total product, decreases. Process plants with these characteristics include ethane cracking and VCM, among many others. These plants are often operated at conversion targets that are not often changed. However, it is worth evaluating whether energy cost changes justify changing plant conditions.

If the energy input in the reactor section is assumed proportional to the combined feed rate and the amount of combined feed converted to product, and the energy usage in the separation section to its combined feed, then it is possible to calculate the optimum conversion. Overall, as conversion increases, the energy use decreases because of the reduced recycle, but the product values also decrease. The optimum will then depend on the energy cost relative to the conversion selectivity coefficient.

The results are shown in Fig. 12 for typical costs, selectivities and conversions. Details of the equations and coefficients are provided in the appendix. At a cost of \$5 per mBtu, the optimum conversion is 38%. As energy costs increase, the optimum shifts to higher conversions and at \$10 per mBtu it is 53%. Not changing the conversion to the correct value results in approximately a 10% lower operating margin for the higher-energy case. Again note that this change can often be made for low or no cost, only a change in the automation setpoints.

**Central utility control.** Efficient and effective control of the overall steam and power generation systems in a large process site are challenging automation problems. Fig. 13 shows a typical

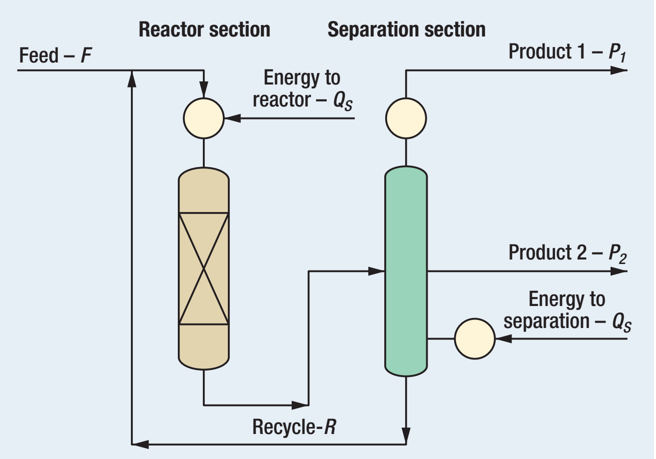


**FIG. 11** Energy and plant conversion.

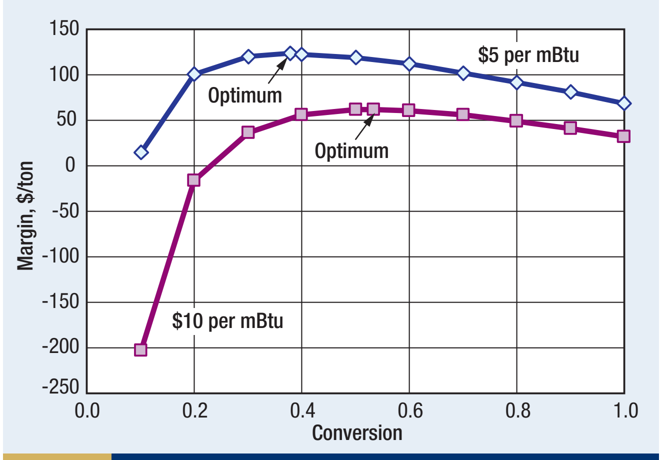


**FIG. 12** Optimum conversion versus energy.

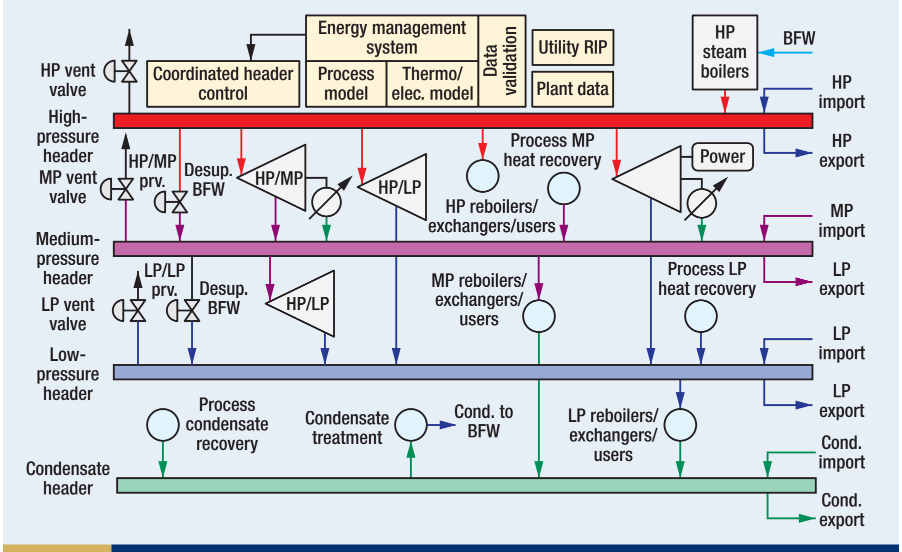


**FIG. 13** Process site utility systems.

system, though individual sites tend to have unique configurations, particularly older ones that have experienced multiple debottlenecking projects and equipment upgrades. There are normally multiple steam headers operating at different pressure levels, each with multiple process users and suppliers. Demand and supply at each level can change with changing plant conditions, and disturbances in one header can propagate to the others. There is often varying steam import or export to other plants at the site further adding to the disturbances to the overall system. In some plants there will be traditional steam turbine power generation on site as well as newer cogeneration units producing both power and steam.

As discussed further in reference 6, there are four interrelated objectives for steam system automation systems. The first, and highest priority, is to ensure safe operation and react to major and rapid demand/supply disturbances such as a major consumer trip. The second is to respond to abnormal operations, again generally relating to the supply/demand balance at a particular steam level, and bring the system back to the region of normal dynamic control. The third objective is that of standard dynamic control, i.e., to reject typical disturbances and hold the system close to desired targets. The fourth objective, and lowest priority, is to operate the system at the economic optimum targets, which generally means meeting desired process utility demands at minimum cost. The overall control system needs to move smoothly between these four objectives and control levels as conditions change.

The figure illustrates the typical advanced controls providing this control. The coordinated header control provides the non-symmetric action required for abnormal and normal dynamic control while minimizing unneeded steam letdown from one level to the next lower one. Feedforward actions on supply/demand changes are incorporated, often in an overall multivariable control framework. The energy-management system includes plant data capture, validation and reconciliation to establish a reliable base case representing current operation. Equipment models are incorporated into an optimization structure that calculates loading targets that maximize financial margin. These targets are then passed to the lower control levels.

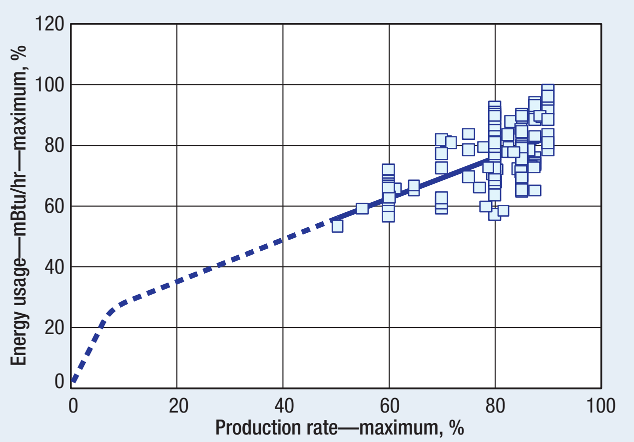


**FIG. 14** Typical plant energy use versus production rate.

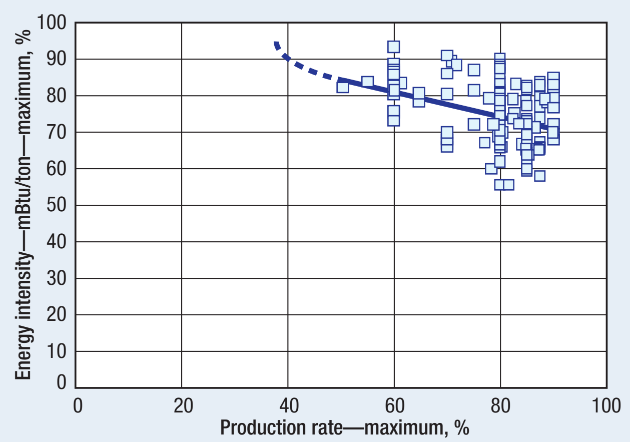


**FIG. 15** Typical plant-specific energy usage versus production rate.

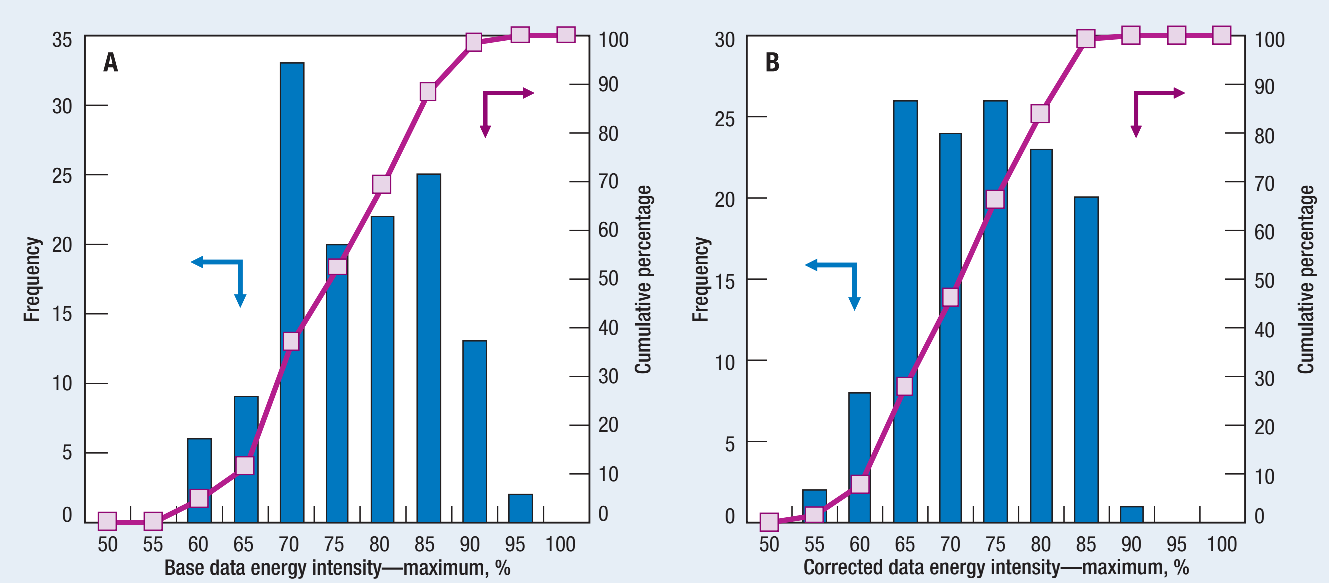


**FIG. 16** Specific energy usage histogram, uncorrected (a) and corrected (b).

In addition to reduced plant incidents, improved automation in the utilities can generally result in reduced energy usage in the central utilities area in the range of two to five %—though many plants have observed much larger savings.[5] This reference also discusses incorporating site environmental emission limits into the optimization framework—a subject of continued interest.

**Analyzing plant energy usage.** A key part of any energy-reduction program is analyzing current plant operation to determine current average energy usage, the variability in this usage and the reasons for it. In analyzing these data, it is important to bring the usage to standard conditions. In Fig. 14, observed energy use from a typical process is plotted versus production rate. Note that if a linear trend line for the normal operating range is projected back to zero production rate, it does not intersect with zero-energy usage. This is normal. For most equipment there is a sharp rise initially in energy use and then an inflection point where the energy use changes at a much lower slope.

One implication of this pattern is that it is misleading to use an uncorrected energy intensity index number (i.e., Btu/ton product) to benchmark a plant. The observed energy intensity index for the plant is shown in Fig. 15 versus production rate, and the substantial slope of the trend line can be seen.

When evaluating actual data, it is then necessary to correct measured energy intensities to a standard production rate. The first histogram, (Fig. 16a) shows the frequency distribution of the uncorrected energy data. The second, 16b, shows the same data corrected to a standard production rate using the trend line coefficient for the energy usage. The reduction in variability is observable and conclusions based on these data are likely to be more reliable than those on the uncorrected data.

The same comments apply for feed type and product grades, i.e., corrections are required to bring the data to a standard set of conditions. Generally, this is done by evaluating usage under differing conditions and applying observed differentials. In some locations usage varies significantly with ambient conditions and corrections may be required for seasonal effects, i.e., summer versus winter usage.

When considering an individual equipment item, the evaluation can be more complicated. Consider the energy use for a major steam turbine/compressor combination. With the surge controls there will be significant production range where the energy use is relatively insensitive to rate. No single energy intensity factor, independent of production rate, would adequately model this usage. **HP**


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## APPENDIX

### Plant model details:

Nomenclature:
$F$—Fresh feed rate, tons per hour
$R$—Recycle rate, tons per hour
$P_1$—Low-valued product rate, tons per hour
$P_2$—High-valued product rate, tons per hour
$x$—Once-through conversion = $(P_1 + P_2)/(F+R)$
$Q_R$—Net energy input, reactor section, mBtu per hour
$Q_S$—Net energy input, separation section, mBtu per hour
$S$—Selectivity = $P_2/(P_1 + P_2)$
$C_F$—Fresh feed cost, \$/ton
$C_{P1}$—Product $P_1$ value, \$/ton
$C_{P2}$—Product $P_2$ value, \$/ton
$C_Q$—Energy cost, \$/mBtu
$M$—Operating margin, \$/hr
$a, b, c, d, e, f, g$—Coefficients

Equations:
$R = (1-x)/x$
$Q_R = a + (b + c\,x)(F+R)$
$Q_S = d + e\,(F+R)$
$S = 1 - f - gx$
$M = P_1\,C_{P1} + P_2\,C_{P2} - F\,C_F - C_Q\,(Q_R + Q_S)$

Coefficients:
$a$— 0.01
$b$—0.03
$c$—0.05
$d$—0.01
$e$—0.05
$f$—0.01
$g$—0.04
$C_F$—600
$C_{P1}$—500
$C_{P2}$—850
$C_Q$—Case 1—\$5/mBtu; Case 2—\$10/mBtu

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