# AIChE Paper Number 110f

# OLEFIN PLANT ENERGY SAVINGS THROUGH ENHANCED AUTOMATION

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Prepared for Presentation at the 2009 Spring National Meeting

Tampa, Florida, April 26 – 30, 2009

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**Abstract**: Energy is the single largest controllable cost for olefin plants and the rise in prices has caused most plants to examine even more closely their energy usage. Automation enhancements can significantly reduce energy usage across all areas of the plant. Some of these savings can be achieved with no investment, only changes in normal operating procedures. In other cases improvements to on-line analyses, measurements and control action are justified but generally require relatively modest investments. The management of the utilities at a major olefin site can be difficult with many daily operating decisions that must balance competing economic and production issues. Real time modeling of process and utility equipment and monitoring of the energy usage in plants permits allocation decisions to be made much more frequently and accurately, often resulting in substantial savings. In this paper, results from several installations are summarized to provide guidance to olefin plant staff on likely areas for savings. Olefin plants are complex and highly integrated and analysis of potential energy savings must recognize this complexity and integration. A systematic evaluation methodology is presented to insure that projected savings are both realistic and attainable with the proposed investment. Examples from several different plants are included.

## Introduction and Background:

Olefin plants are large energy consumers with energy the largest variable operating cost after feedstocks. Using energy efficiently has been and remains a primary goal for olefin producers. For most plants the marginal fuel is natural gas and the change in average US natural gas costs (1) over the last few years, as illustrated in Figure 1, illustrates both the general rise in energy prices and the volatility in these prices. At a price of $6 per million BTU's (mBTU), a 5% energy saving is worth approximately $4.5 million per year for a typical North American 500,000 T/ yr ethylene plant with naphtha feed.

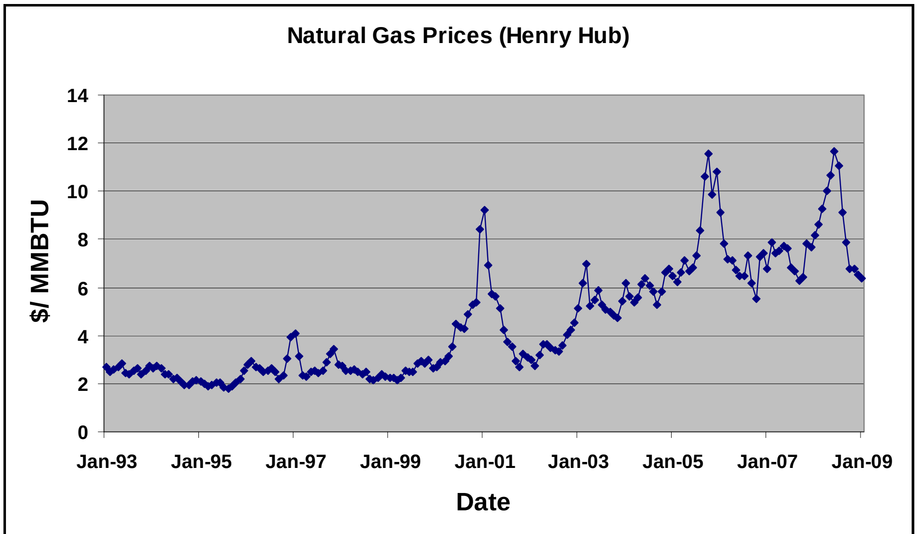


**Figure 1: Natural Gas Prices**

In addition to the direct price of energy, it seems likely that the US will adopt some regulations regarding greenhouse gas emissions in the next few years. If the regulations in other countries are a guide, these may take the form of a "Cap and Trade" on CO2 emissions or a carbon tax. 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 these CO2 reductions are valued at $20 / ton, which is at the low range of recent prices in Europe, then reductions in natural gas or equivalent fuel usage would have an additional value of approximately $1.3 / mBTU. This would add a value of approximately $1 million per year to the 5% energy savings above.

Although many olefin plants have been investing to reduce energy consumption, there is still a wide variation in energy usage, even after correction to standard conditions for feed quality, product composition and process configuration as illustrated in Figure 2. Surveys, such as reference 2, repeatedly show wide gaps between the most efficient plants and the least efficient. There is at least a 40% spread in energy usage between the most efficient plants and the least efficient, even after correction to standard conditions for process configurations, product grades and feed types. This variation is primarily due to the age of the equipment with older plants often having less efficient equipment and less heat integration than newer plants. The rise in average energy costs has increased the incentive for additional investments and there are typically still additional opportunities to be found even in better performing sites, generally estimated to be of the order of 20%. Note that the theoretical minimum energy for these plants is still well below current levels at roughly 25% of the current average value.

There are a large number of possible investments that can be made to reduce energy with different costs and different impacts as illustrated in Table 1. These range from low cost/ relatively low impact programs such as those to reduce steam trap and other steam leaks to installation of integrated gas turbines with very high costs and long implement times and correspondingly high impact. In this paper the focus will be on automation investments. These generally fall in the low to medium cost range with savings that are typically mid range. As a result, the expected return on investment for these programs can be quite high.

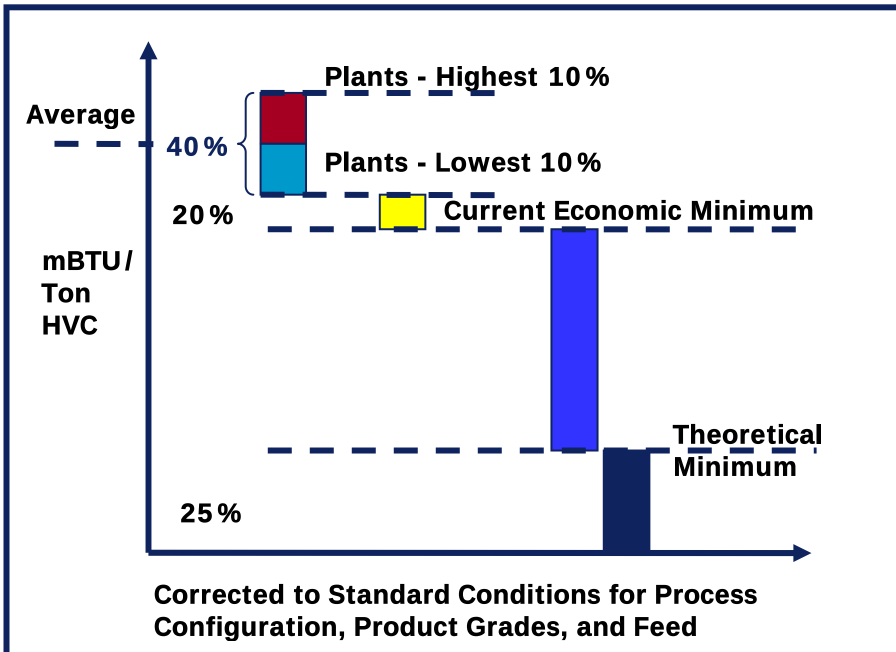


**Figure 2: Olefin Plant Energy Usage**

| **Table 1**<br>**Potential Olefin Energy Investments** | | | |
|---|---|---|---|
| | **Capital Cost (Time To Implement)** | | |
| **Potential Energy Savings** | **Low** | **Medium** | **High** |
| **High** | | | Integrated Gas Turbine: Cogeneration;<br>Furnace / Separation Process Upgrade;<br>Replacement Low Efficiency Process Equipment |
| **Medium** | Improved Automation;<br>Operating Procedure Changes;<br>Energy KPI Monitoring | Advanced Control/ Optimization;<br>Site Energy Management Systems | |
| **Low** | Increased Insulation;<br>Steam Trap/ Leak Management;<br>Exchanger Maintenance<br>Condensate Recovery | | |

## Automation Impact on Energy

Typical automation and related systems in olefin plants are shown in Figure 3. The ISA 95 specification groups them in Levels 1 to 4 as shown. Almost all of these systems will have some impact on energy usage but the focus of this paper will be on those in Levels 1 to 3. Economic values calculated in the text following for potential improvements will be based on a typical North America (NA) Naphtha feed plant producing 500,000 T/ Year ethylene with the assumed properties listed in the Appendix. No credit is taken for the value of greenhouse gas reductions.

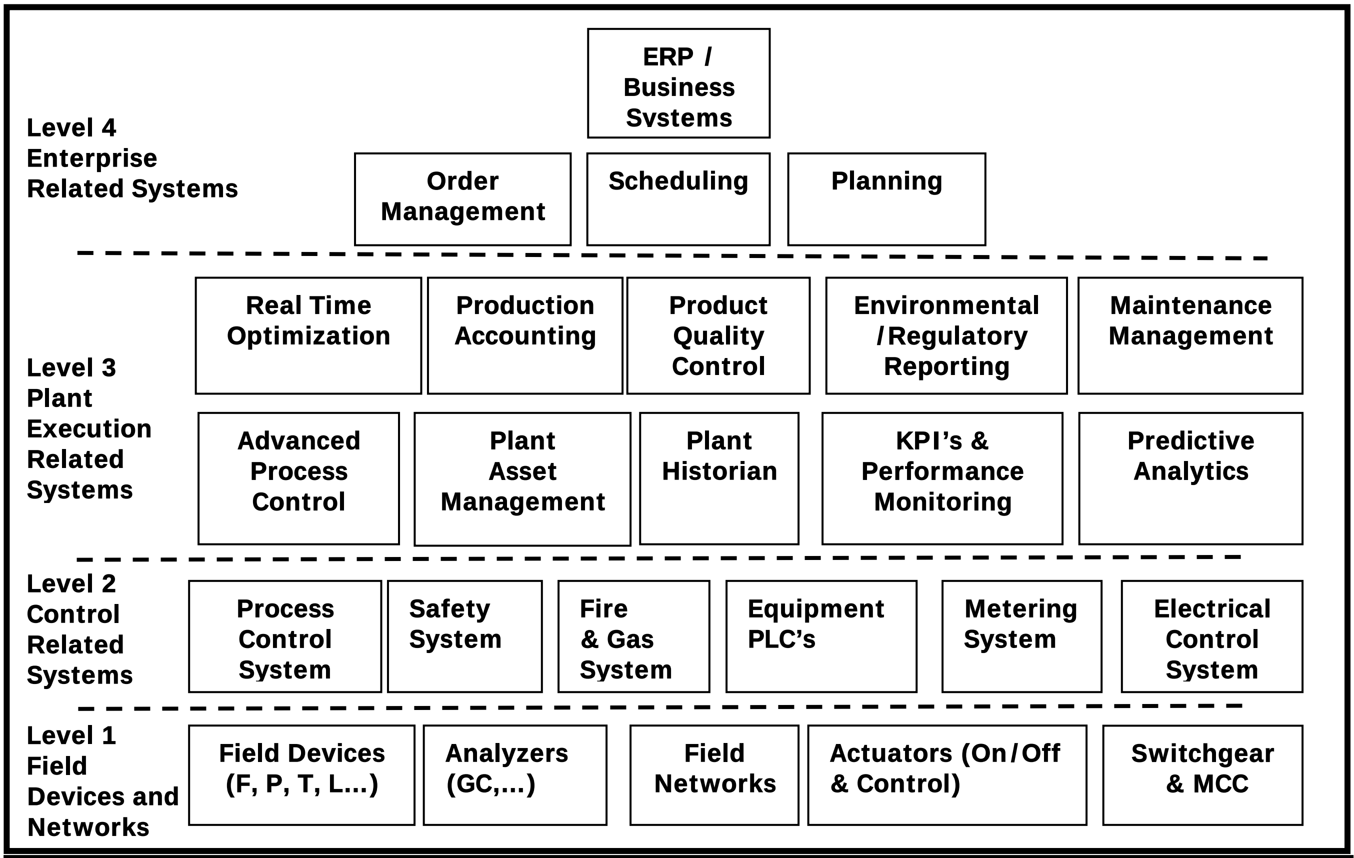


**Figure 3: Olefin Plant Automation Systems**

There are two primary mechanisms by which new investments can reduce energy usage at specified production levels - either by reducing the cost of supply or by reducing process energy demand. Reducing supply costs can be subdivided further into investments that reduce external purchase costs or those that increase internal energy production efficiency. Table 2 lists some of the specific ways that automation and related systems can reduce energy within this framework. There are a large number of impact areas, too many to be covered in any single paper. Olefin plant advanced control and optimization systems are one of these and can obviously have a large effect on energy usage. However, their general functionality and energy impact has been covered previously in many papers, for example reference 3, and will not be discussed further here. In this paper a few specific automation impact areas will be covered that, in the author's experience, are not as widely recognized as others.

**Table 2**
**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<br>  • Furnaces<br>  • Quench/ Frac<br>  • Compressors<br>  • Distillation<br>• Improved Control Loop Performance<br>• Maximize Process Heat Recovery /Minimize Losses<br>  • Maximize Recovered Steam<br>• Minimize Pressure Drop<br>• Minimize standby turbines and boilers<br>• Better control at low rates | • Improve Combustion Efficiencies<br>• Steam System Management<br>  • Steam Header Management (minimize venting, letdown, pressure)<br>  • Maximize Recovered Steam<br>  • Blowdown Control<br>• Steam vs. Electric Turbine Optimization | • Energy Management System<br>  • Power, Fuel, Steam purchase |

## Control Loop Performance Improvements

Efficient and effective execution of the basic control loops at the plant is obviously essential to successful operation of the plant and to other functions such as advanced control and real time optimization. Control loops are composed of a measurement element; an actuator, most commonly a valve; and an executed control algorithm. As suggested in Figure 4, improvements in each of these elements can lead to reduced energy usage. In the sections below, some specific improvement areas applicable to olefin plants are discussed.

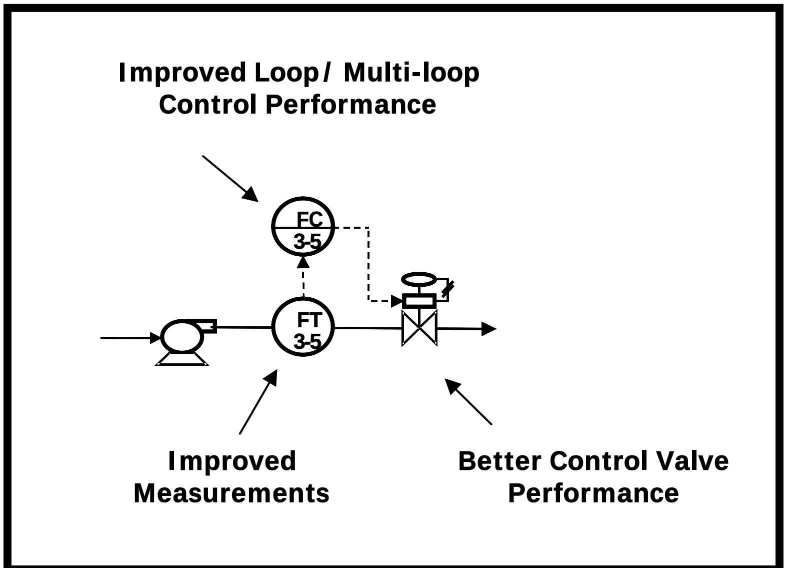


**Figure 4: Energy Saving Through Control Loop Improvements**

### Measurement

One of the first areas to evaluate for potential automation changes is improving the measurements of key plant variables in terms of accuracy, location, frequency and number. Physically relocating a measurement to reduce dead time in a loop can often do more to improve performance than any tuning or algorithm. Increasing the frequency of measurement, for example for a gas chromatograph, can be similarly beneficial for loop performance. Increasing the accuracy of the measurement can sometimes be used to reduce operating constraint margins and change operating conditions to save energy. In the past it has typically been very expensive to consider adding measurements after a plant has been built because of the permitting and wiring requirements. Wireless transmitters, which are now more widely accepted in plant use, significantly reduce the cost of adding additional measurements.

There are some specific measurement improvements that can have a significant impact. Many olefin plants experience wide variation in their plant fuel gas composition and the corresponding heating value. This variability induces variability in the cracking furnace combustion control and increases the required operating constraint margin above the optimum air/ fuel ratio.

In Table 3 below volumetric heats of combustion for standard fuel gas components are compared with their mass equivalents. Note that there is much less variability on the later 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.

| Table 3<br>Fuel Gas Component<br>Heating Value | | |
|---|---|---|
| **Component** | **Heat of Combustion BTU/ cu ft (gross)** | **Heat of Combustion BTU/ lb (gross)** |
| **Hydrogen** | 320 | 60957 |
| **Methane** | 1011 | 23861 |
| **Ethane** | 1783 | 22304 |
| **Propane** | 2563 | 21646 |
| **Butane** | 3374 | 21293 |

This suggests that fuel gas control on a mass rather than volume basis will eliminate much of the variability and that is the experience of plants that have adopted such control. With modern flow measuring devices that will directly measure mass and also provide a gas density measurement it is easy to implement such controls. The value of this improvement will depend on the normal variability of the fuel gas composition and its effect of the furnace excess air operating margin. However, for a typical NA plant this improvement might be 0.2% in furnace efficiency which would be worth approximately $250,000 per year, more than enough to give a one year simple payback on fuel gas mass meters.

Another key measurement improvement for cracking furnace combustion is the use of stack gas CO measurements. While some sites have installed these analyzers previously, their use has been limited by the perception that the sampling systems were high maintenance items and the analyzers themselves were somewhat unreliable. However, CO analyzers are now available for in situ installation with no sampling system and no calibration gas requirements and have proven to be quite reliable.

The recent research by The John Zink Company, as shown in Figure 5, and discussed further in reference 4, provides further incentive for improved online combustion gas analysis. The research shows that ambient air temperature and relative humidity changes have a surprisingly strong effect on combustion conditions, even with constant fuel composition. The first graph shows that a heater tuned to satisfactory combustion conditions at 80 deg F and 0% relative humidity (14.7 psia) with constant air/ fuel ratio control will experience incomplete combustion if the ambient temperature changes to 100 deg F and 80% relative humidity with an ambient pressure of 14.1 psia. Such changes are well within the range of ambient condition changes experienced on the US Gulf Coast. The second graph shows the specific effect of these changes on CO composition.

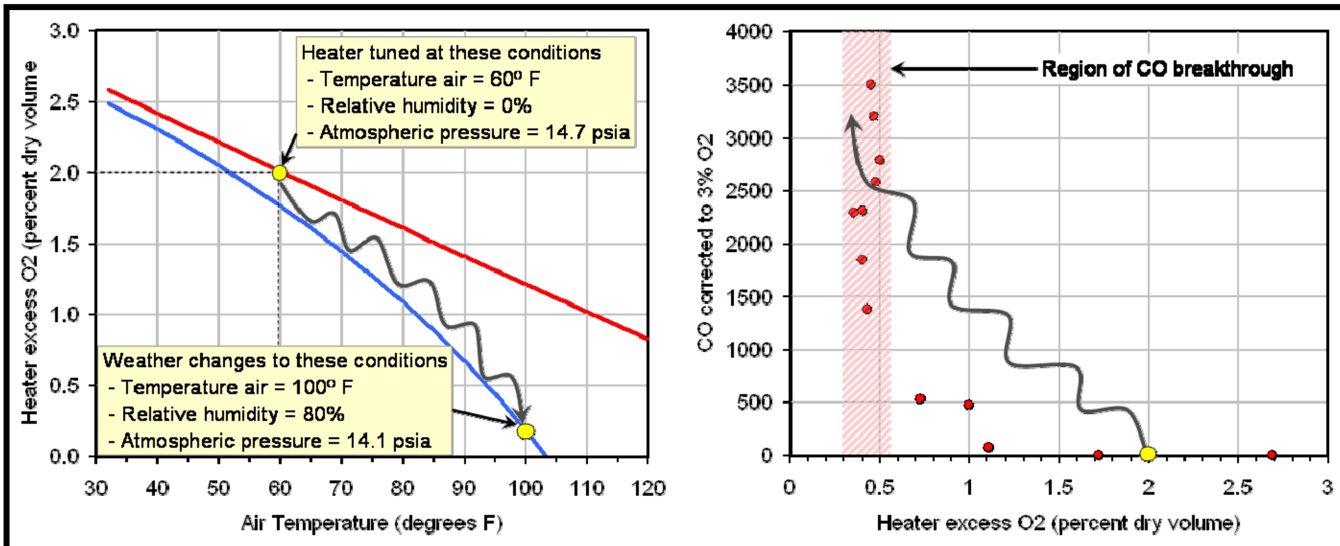


**Figure 5: Ambient Effects On Combustion:**
**From Reference (4):** ***Bussman and Baukal***

With varying fuel composition, direct CO measurement is the best way to monitor combustion efficiency and enable close control of the furnaces.

### Valve Performance

Olefin plants often operate at varying load conditions, due to market conditions, feedstock availability, furnace decokings, and process equipment availability. Control strategies need to be designed and implemented that reflect efficient performance over the full range of operating conditions and field devices need to be chosen and sized appropriately. The cracked gas and refrigeration compressors are major energy consumers in olefin plants and proper anti-surge control, as shown in Figure 6, is an important factor in minimizing energy usage at lower load conditions. The stage recycle or anti-surge valve is controlled by the surge control system which typically includes compensation for changes in speed, suction temperature and pressure, and gas composition, if measured. The recycle valve has to open very quickly and accurately to re-circulate gas from the stage discharge to the suction. The ability to stably and

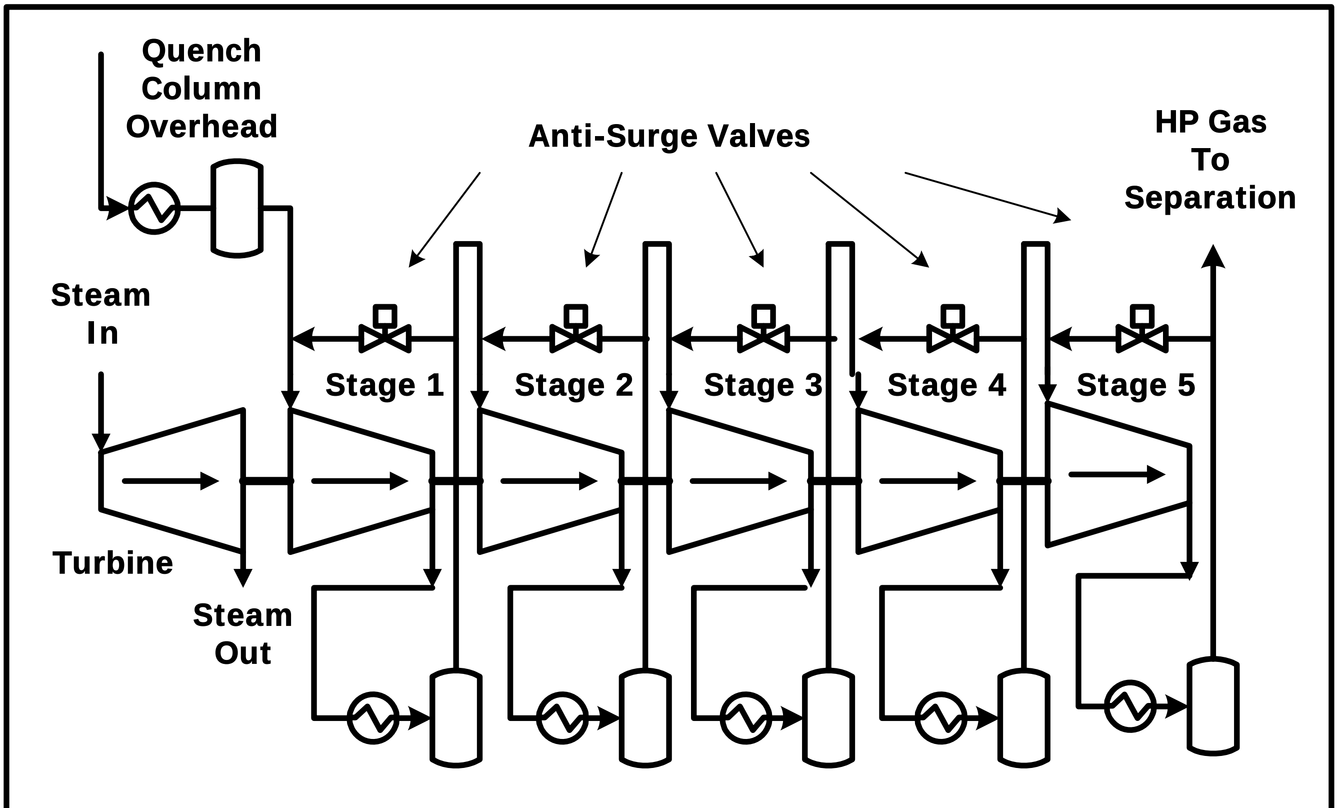


**Figure 6: Compressor Surge Control**

safely operate the compressor closer to the surge limit saves energy and the size of the required operating constraint margin is dependent on the response characteristics of the anti-surge valve. Developments in valve technology now permit even very large recycle valves to go from fully closed to fully open and vice versa in under two seconds.

In figure 7 a typical surge line for a compressor is shown. The actual normal operating surge limit will be set at some margin from the actual limit. By installing a higher

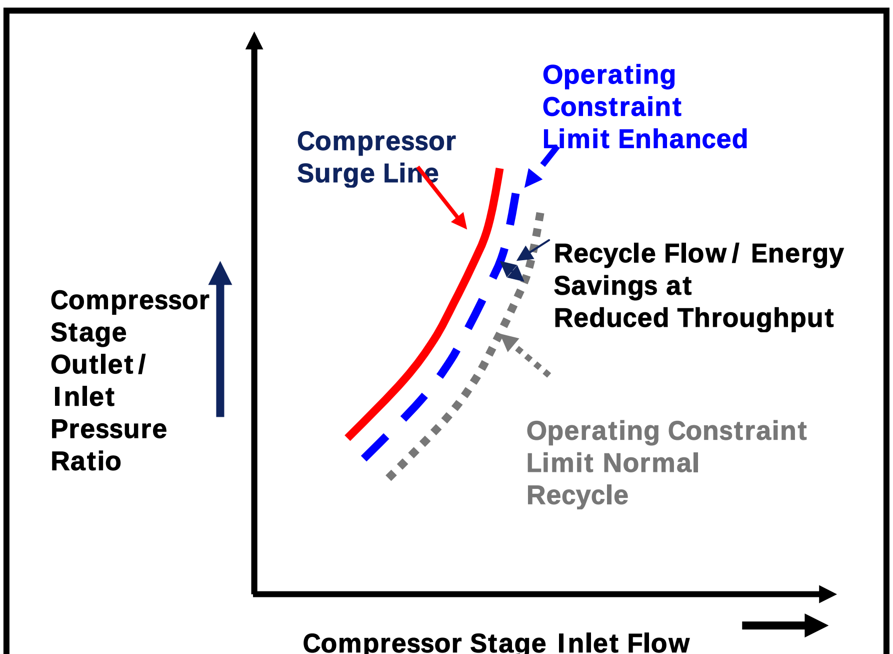


**Figure 7 – Compressor Energy Savings With Improved Valve Performance**

performing valve it will be possible to reduce the operating margin. The value of such improved control depends on the current operating margin and the percentage of time that the plant runs at lower load conditions. If the plant is in low throughput conditions one-quarter of the time, the value of the reduced recycle would be approximately $120,000 per year for the reference plant. In addition, at normal operating flow conditions, a higher performing valve would permit the cracked gas compressor suction pressure to be reduced, resulting in increased ethylene yield, if economically justified.

## Loop Dynamic Analysis and Tuning

One of the most cost effective measures that can be taken to improve energy control performance is to analyze the dynamic response of key basic control loops, identify problems and correct them. If poor performance exists, the problem(s) could be in the measurement, the control valve or other actuator, or in the loop dynamic tuning. It is also important to identify when the problems are external to the loop either in terms of process disturbances or due to multi-loop interaction. The analysis can be facilitated by use of a hardware/ software system that permits high speed loop data capture and automates more sophisticated analyses. Assuming that the field automation equipment is performing adequately, the next step is to tune the loops with a rigorous approach that clearly recognizes the required tradeoffs among stability, robustness, performance and total system dynamics. The Lambda tuning method (5) is a well proven, mathematically sound methodology that deals explicitly with the tradeoffs above. The top set of graphs in Figure 8, from reference 6, shows the performance of the

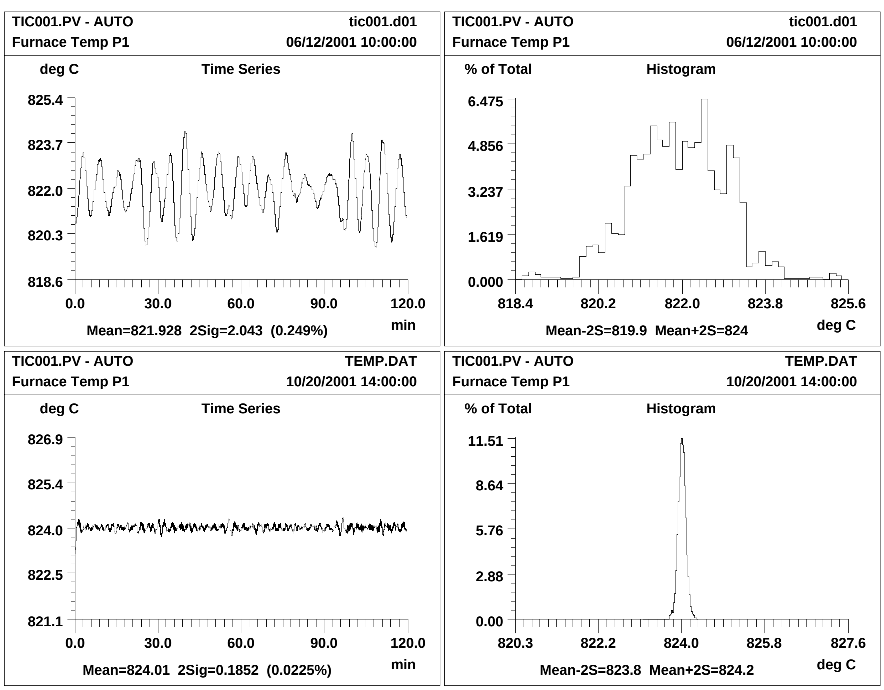


**Figure 8 – Cracking Furnace Outlet Temperature Control Tuning Effect; From Reference 5; *Filho***

outlet temperature controller on a cracking furnace when tuned using older methodologies, in this case Ziegler-Nichols Quarter Amplitude Damping. The right hand graph is the histogram of the temperatures. The lower graph shows the performance of the same control loop, with the same valve and measurement, after tuning using the Lambda methodology. The reduced variability leads to reduced energy consumption at the same target setpoint. Since furnace fuel consumption is a non-linear concave function of outlet temperature, excursions above the setpoint consume more fuel than is saved by excursions below the setpoint, at the same mean. In addition, tube coking rates are similar non-linear functions of temperature and reduced variability leads to reduced coking rates, again at the same mean temperature. As the furnace tubes coke, the heat transfer efficiency is reduced, resulting in increased fuel usage. Alternately, the reduced variability can be used to increase the target outlet temperature setpoint at a constant coking rate equal to original value. Such an increase will increase once through conversion on the cracking furnaces. For ethane and propane feed furnaces this will decrease overall recycle rates at constant net conversion with a potential net energy savings. The proper balance between these two possibilities for energy saving can be calculated from an overall plant economic analysis.

## Steam System Management and Control

Control of the overall steam system in an olefin plant is one of the most challenging automation areas. Figure 9 shows a typical system though every site tends to have a unique configuration, particularly sites that are older and have experienced several debottlenecking projects and equipment upgrades. The current control systems tend to have evolved over time with new controls added to handle a particular issue, sometimes without full consideration of their effects over the complete range of expected operation.

There will be normally be multiple steam headers at different pressure levels, each with multiple users and suppliers. Since most of the VHP steam is supplied by heat recovery on the cracking furnaces, this supply varies with furnace changes in rate, feed type, and severity. Demand at each level is similarly changing with changing plant conditions and changes in one header can induce disturbances in the others. There is often varying steam import or export to other plants at the site further adding to the disturbances to the system. The large steam turbines for the cracked gas and refrigeration compressors are the largest utility consumers on the site after the cracking furnaces but there are typically many other users. In some plants there will be on site power generation as well.

A further complexity with regard to the control of steams system is the hierarchy of objectives as shown in Table 4, which lead to different control responses depending on the state of the system. The lowest priority is utility optimization with the objective of determining the most economic operating conditions for the current plant demand. They are normally calculated by a steady state optimization model. Typical control actions at this level might be to specify targets, such as the turbines' extraction steam flows, which would minimize venting and steam letdown from one level to the next lower level and maximize power production. More sophisticated optimizer models might actually calculate new header pressure targets, within a limited range, which would increase unit capacity or increase efficiency. The next higher control level is the normal dynamic operating region control which responds to typical disturbances in supply and demand and adjusts conditions to compensate and maintain stable operation. There are many constraints on operation, many manipulated variables and lots of interaction among the variables leading to use of multivariable control for this region. Feedforward action on measured disturbances is used to further improve dynamic performance. The adjustments required for dynamic compensation can be in short term conflict with the optimization objectives. For example, assume there are two manipulated variables with roughly equal steady state gains with regard to header pressure. Assume further that one has a small coefficient in the optimizer objective functions but a large time constant with associated slow dynamic effect while the other has the opposite characteristics. From an optimization point of view it would be desirable to primarily manipulate the first to control header pressure since changes would have a small effect on the objective function while dynamic compensation priorities would favor the second. This leads to control in priority levels 2 and 3 with short term dynamic considerations governing while the optimizer is setting targets for these variables that are only able to be satisfied in relatively steady operation. .

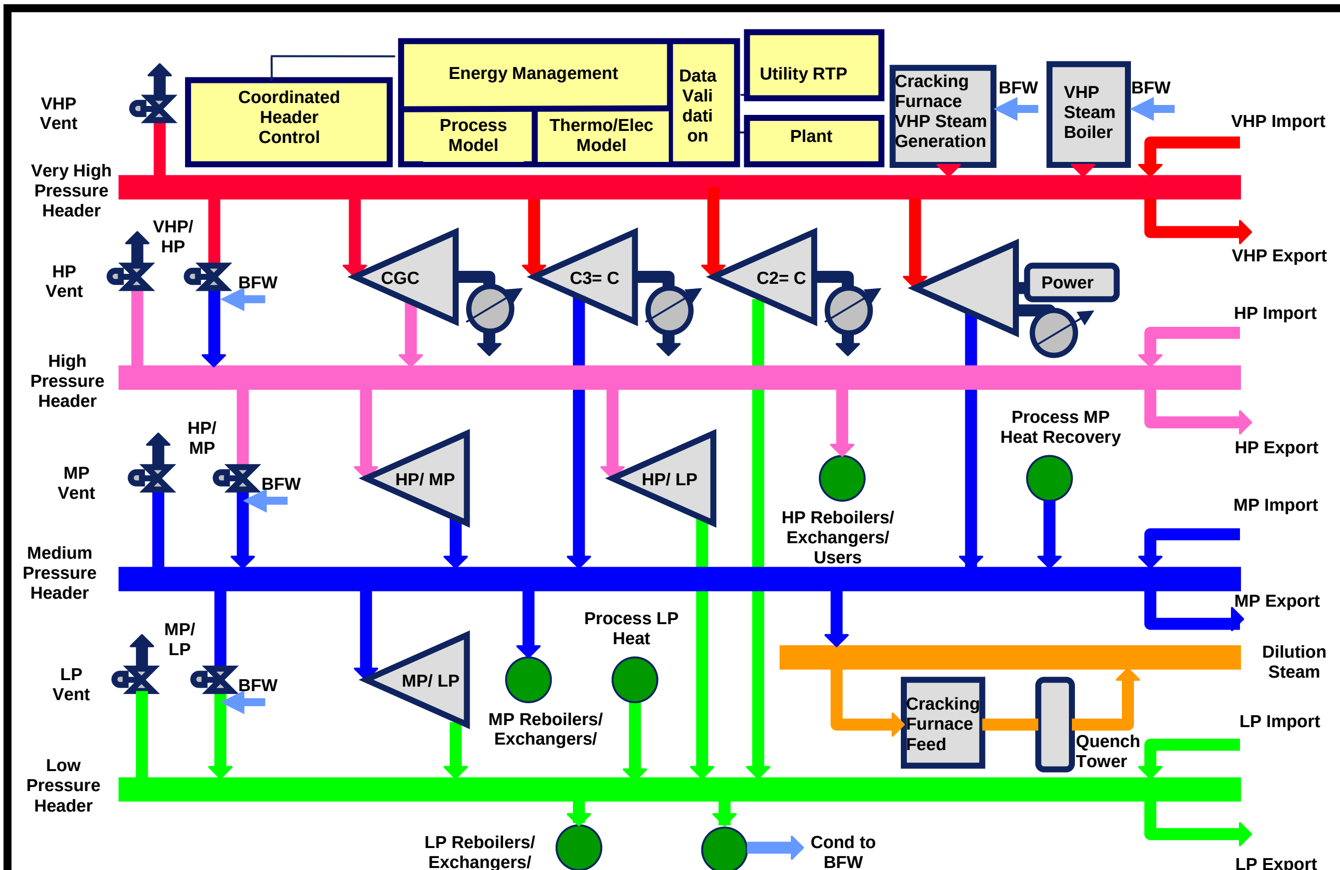


**Figure 9: Typical Olefin Plant Steam System**

| Table 4<br>Steam System Control Hierarchy | | | | |
|---|---|---|---|---|
| **Priority** | **Objective** | **Typical Issue** | **Typical Response** | **Control Function** |
| **1.** | Insure Safe Operation | Boiler Trip<br>Compressor Trip<br>Furnace Trip | Load Shedding;<br>Maintain dilution steam flow;<br>Provide steam to flare<br>Maintain flows to main turbines | Pre-programmed Sequence Control |
| **2.** | Correct Abnormal Operation | High / low header pressure outside range<br>Furnace decoking | Steam Letdown Control;<br>Venting | Overrides |
| **3.** | Respond to Normal Disturbances | Meet varying turbines steam Requirements<br>Meet varying user steam requirements<br>Compensate for varying furnace VHP steam supply | Control Turbines' Extraction Steam<br>Steam Letdown Control<br>Supplemental Boilers<br>Load Controls<br>Steam Import/ Export<br>Vary power generation | Cascaded Multivariable Control |
| **4.** | Optimize Utility Supply | Economics | Maximize internal power generation<br>Minimize letdown<br>Manage Import/ Export Steam<br>Steam/ Electric Turbine Selection | Steady State Optimization Model |

However, there are relatively common situations where larger disturbances move the controlled variables outside their normal operating region into what might be termed an abnormal state. These disturbances can be either on the demand side or the supply side, for example a decoking of a furnace where there is an increased demand for steam for decoking simultaneous with a loss of the furnace's VHP steam production. The objectives in this control region are to smoothly return the steam system back to the normal control range. One of the characteristics of this state is that the control actions are non-symmetric, i.e. there are different desired control actions if the variable is below its operating range than above. For example, if the pressure in the HP header was above the control range the control action might be to open the HP/ MP letdown as long as the MP header pressure was not above its target range. If the pressure went higher then the action would be to open the HP vent to atmosphere. Conversely, if the pressure was below its range the action would be to open the VHP/ HP letdown, assuming the VHP header was in range. If the pressure continued to drop the action might be to shut the HP/ MP letdown, if open. If not open, then the action might be to reduce feed to the plant. Again this assumes that normal control actions were at their limits. There are further complications because one of the headers can be in an abnormal state while the others are in a normal control situation. Generally this control action is handled via override functionalities with some obvious complexities.

Priority one control is assuring safe operation and activation of this level typically is a response to a furnace or other equipment trip. The header controls have to be prioritized in order of importance in the event that a severe upset occurs. This means that the headers with the highest priority will be kept as close to the targets as possible while other header targets will be relaxed as required in the order of their priority. The priorities might be to keep enough MP steam in the dilution steam headers to prevent decoking of the furnace tubes, provide dispersion steam to the flare, and also enough VHP steam to allow the main turbines to be safely dropped to minimum rates or shut down. This is clearly a challenging control issue requiring careful dynamic coordination and is often handled via pre-programmed sequences along with manual intervention.

Successful control design and implementation then involves integration among these four priority levels with smooth transition from one level to the next as shown in Figure 9. The number of manipulated variables, constraint variables, their interactions, and the frequency and magnitude of disturbances provide significant potential benefits for improved automation. Improved control loop performance via improved measurements, better actuator response, and better tuning are all beneficial.

Modeling the overall steam and utility system and using this model for realtime economic decisions can have a large benefit. Figures 10 and 11 from reference 7 show the results from implementation of such a modeling system on a large petrochemical site. Figure 10 shows the calculation of the optimum operation and 11 the difference in the objective function if the optimum is implemented. It is common to find that the benefits from such an installation can range from one to three percent of the ongoing utility costs. This

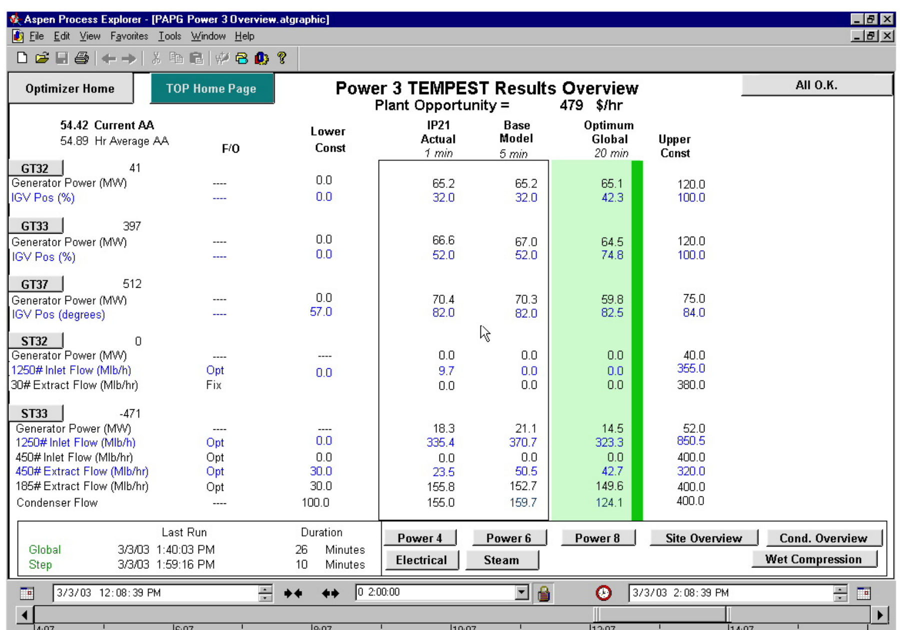


**Figure 10: Utility Modeling and Optimization; From Reference (7); *Bearden and McNally***

would yield a potential value of approximately 400,000 $/ Yr to 1,200,000 $/ year for our reference plant.

In addition, the use of the model for off line "What if" analysis permits evaluation of investment options including improved measurements and other operating equipment configurations. Calculating incremental energy costs for each level of steam at current conditions can be helpful for setting priorities.


**Figure 11: Optimization Results From Reference (7): *Bearden and McNally***

Ongoing equipment energy use monitoring is provided by the system and is a key to sustaining reductions in energy usage. Having a centralized single point of calculations reduces internal disagreements over actual usage figures.

## Developing An Automation Energy Saving Program

Most North American olefin plants have plans to reduce energy use. However, a structured program to assess, implement and sustain automation energy savings can still have high value.

### Assessment

A typical automation assessment would have the steps shown in Figure 12. The objective of the assessment is to develop a detailed action plan with a timeline for

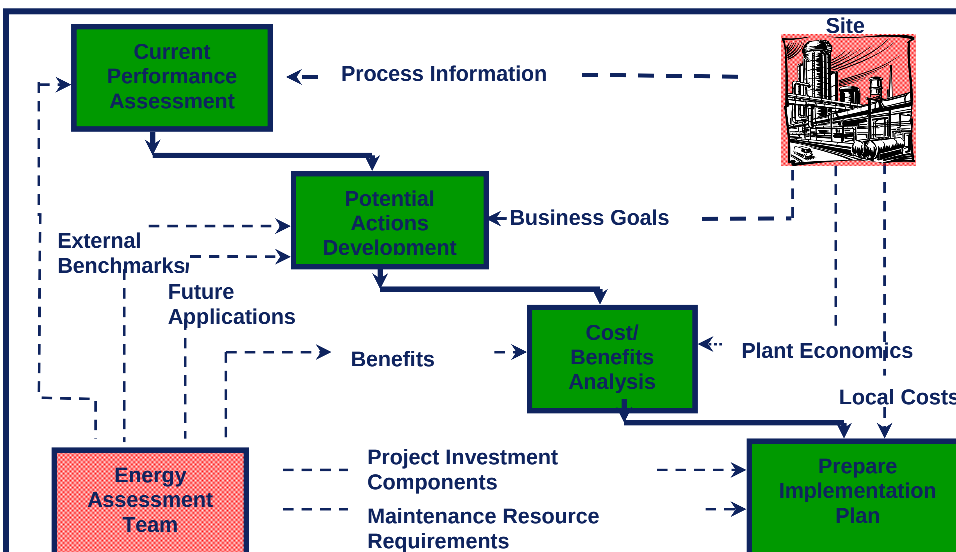


**Figure 12 – Assessment Phase**

execution. The first step in the program is then to identify where the plant is today with respect to functionality, use, and integration of existing energy monitoring and control automation and information infrastructure.

Part of this assessment is an identification of control loops, including advanced control loops, which have a major impact on energy usage. These loops will be the initial target for an ongoing performance analysis program. To identify these loops it is helpful to categorize them as shown in Table 4. First the control loops are ranked relative to the impact of significantly degraded performance of the loop on incremental energy consumption, generally expressed in monetary terms. This ranking can be done either using engineering knowledge of the energy relationships or by a correlation analysis of plant data or some combination of both approaches. Next the historical frequency of significantly degraded performance is determined, generally from the site data historian and maintenance records. From this analysis, loops are prioritized based on the product of their impact and problem frequency. A possible result of the analysis is also shown in Table 4.

**Table 4**
**Critical Energy Loop Analysis**
**Priority For Monitoring / Maintenance**

| | **Historical Frequency Significantly Degraded Performance** | | |
|---|---|---|---|
| **Energy Loss Consequences Significantly Degraded Performance** | **≤ Once A Month** | **From Once A Month to Once A Year** | **≥ Once A Year** |
| **≥ $500k per year** | **Priority 1**<br>- Fuel Gas Pressure Control<br>- Combustion Air Control<br>- Steam Header Letdown Control | **Priority 2**<br>- Coil Outlet Temperature Control<br>- Fuel Gas Flow Control<br>- Steam Header Pressure<br>- Turbine Extraction Steam Control | **Priority 3**<br>- Dilution Steam Ratio Control<br>- CGC Suction Pressure Control |
| **$50k to $500k per year** | **Priority 2**<br>- Steam Drum Level Control | **Priority 3**<br>- Coil Differential Temperature Control<br>- Quench Column Overhead Temperature<br>- Separation Column Product Quality Control | **Priority 4**<br>- HC Feed flow controllers<br>- Separation Column Pressure Control<br>- Reboiler Flow Control |
| | | | |

| Table 4<br>Critical Energy Loop Analysis<br>Priority For Monitoring / Maintenance | | | |
|---|---|---|---|
| | **Historical Frequency Significantly Degraded Performance** | | |
| **Energy Loss Consequences Significantly Degraded Performance** | **≤ Once A Month** | **From Once A Month to Once A Year** | **≥ Once A Year** |
| **≤ $50k per year** | **Priority 3**<br>- BFW Flow Control | **Priority 4**<br>- Acetylene Converter Outlet Temperature Control | **Priority 5**<br>- Fuel Gas Pressure Control |

For the higher priority loops, further analysis is done to identify the source of the problems, i.e. measurement, actuator, tuning or external factors such as disturbances or interactions with other loops. From this analysis, a set of potential actions to correct the problems is produced. At this stage it is also important to include analysis of possible new improvements to the automation such as increased measurements of energy critical variables including additional analyzers, advanced control implementation, infrastructure improvements such as networking to the motor control center, and resizing of pump rotors and associated flow valves for current operating conditions. It is common to identify savings that can be obtained with just a change in operating procedures, for example relaxing an operating constraint that was set based only on historical experience rather than real process limits. In addition, it is useful to benchmark the plant against industry leaders to help identify areas for potential improvements.

Next is the cost / benefit analysis to provide a screening for potential project evaluation. It is important is to match the uncertainty in both the cost and benefits. For screening purposes, it is not necessary to have a cost estimate +/- 10% when the benefit estimate is +/- 50%. Recommended projects are evaluated on the basis of return on investment. The recommended sequence is then set by:

- Size of investment and schedule duration to achieve benefits
- Required infrastructure changes needed
- Cash flow
- Available resources
- Leveraging existing applications
- Project dependencies

With this information, a funding program request is prepared and submitted for internal approval. At this point it is necessary to refine cost estimates to the +/ - 10% accuracy level. It is generally advisable to choose some relatively quick low investment projects for initial implementation. Even if there are larger investment items with higher ROI, quick success and actual benefits help build internal support for these larger projects. After these initial activities, it is typical to choose some projects for re-assessment in more detail. When funding is authorized, implementation proceeds.

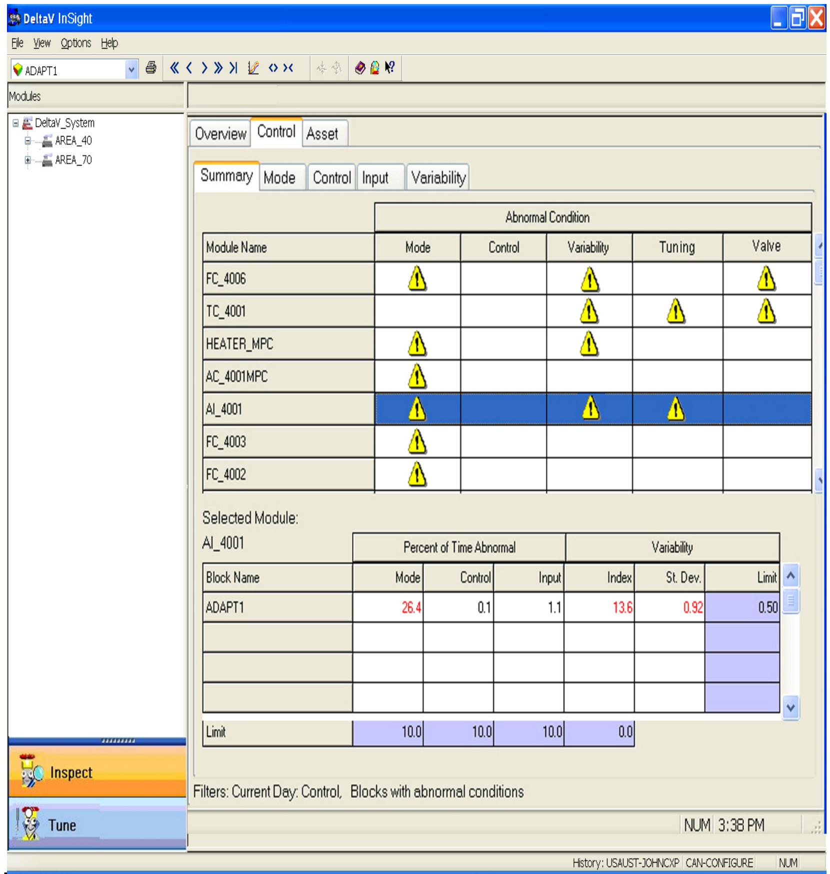


**Figure 13 – Loop Monitoring**

## Implementing and Sustaining the Program

After the program is approved, initial implementation proceeds. However, it is important to build into the program the tools and procedures to maintain the savings. There have, unfortunately, been too many programs that were initially successful but the savings degraded over time with changes in personnel and management attention. For success, there needs to be management "buy-in" and assignment of specific accountability to individuals with regular review. Most companies find that this is aided by routine calculation, tracking and reporting of automation metrics. Figure 13 shows a modern control performance monitoring package that identifies when key loops are operating outside their desired range, provides an alert to operations and maintenances of the degraded loop performance and provides suggested areas for further investigation with regard to the cause of the conditions. Providing visibility of results to operators and supervisors helps to keep the performance at the desired levels.

## Conclusions

High performing automation systems are a key to maintaining minimum energy usage in olefin plants while meeting other operating targets. Enhancements to automation can provide some of the highest payback items to consider for energy saving process upgrades. In this paper a review of some improvement areas is presented. The potential benefits from improved automation will, of course, vary with the current state of the plant. However, these benefits could approach five percent of the energy usage with a value of six million dollars per year for the reference plant considered, comparing a plant with a low utilization of automation with one with current best in class. Greenhouse gas savings would be additional. This certainly provides sufficient incentive to consider investments in this area as plants strive to continually reduce their usage.

## References

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## Appendix

### Reference Plant - Basis For Economic Calculations

The following assumptions were used for economic calculations in the paper:

- Olefin Plant with Naphtha Feed
- 500,000 Tons/ Year Ethylene Production
- Total Energy Usage - 31 mBTU/ T Ethylene (16 mBTU/ T HVC)
- Incremental Fuel - Natural Gas at $6/ mBTU
- No credit taken for value of CO2 reductions
- Cracking Furnace Base Case Thermal Efficiency – 88%
- Compressors; CGC – 60000 HP; Propylene – 40000 HP; Ethylene – 15,000 HP.
- Total Steam Usage in plant – 11 mBTU/ T Ethylene