Avoid advanced control project mistakes

Basic frontline, advanced regulatory, constraint and inferential controls must work well before additional sophistication is attempted. A control technology audit of five refineries highlights mistakes made

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On-line process optimization is worth working for but without robust advanced controls it will never happen. I recently evaluated how well advanced controls worked in five refineries. Having spent money on such projects, the refineries faced a situation in which there was no measurable improvement in overall plant performance. These refineries are owned by different companies, yet they share a pattern of mistakes in administrating advanced controls. Highlighting these mistakes shows ways to improve the organization of advanced control technology, to avoid obvious pitfalls.

Tuning basic controls. Everyone knows that building advanced controls on poorly tuned basic controls is akin to building skyscrapers on quicksand. But, in the audited refineries, this was theoretical knowledge. The art of loop tuning demands a combination of skills not often found in one individual: process dynamics and control theory; process engineering and understanding of interloop interactions; and familiarity with the economic driving forces and operating objectives. The skill requirements, and also the fact that tuning is a time-consuming activity, lead to tuning practices such as cutting or adding to the gain whenever someone complains, without conducting a test to identify the loop dynamics and without appreciation of the tuning objective for the loop.

The most commonly mistuned loops in the audited refineries were level controllers, where the operating objective is usually to keep the flow steady and accept level swings, but where tuners often prefer the opposite. At best, poorly tuned level controllers will introduce unnecessary disturbances in downstream equipment. At worst, the flow swings will interact through heat exchangers and recycles to drive the whole unit unstable without anyone knowing how to discriminate between cause and effect. The tuner would then respond by detuning the wrong loops adding sluggishness to the instability. In one of the refineries tuning was so bad that it was impossible to test any of the advanced control schemes because none of the units were stable.

It is not always easy to improve the tuning in a refinery because of union rules and availability of people, but as a start, it would be beneficial to consider the following steps:

• Identify the people permitted to tune loops and train them well.

• Improve communication between the instrument technician who tunes the basic control loops and the process control engineer who tunes the advanced controls.

• Keep a history of tuning changes and reasons for changes. Hopefully, this will eliminate perpetual tuning and detuning of the same loops.

• Create a guideline for tuning to cover the most common tuning objectives for flow, pressure, temperature and level loops.

• Create a standard for predictive control for analyzer loops with excessive deadtime.

• Do not shy away from spending time on tuning loops. An expert may be able to tune 10 loops per day with the usual refinery mixture of loops. Advanced control loops are slower and even more time-consuming.

Analyzer maintenance. Some refineries believe analyzers are necessary and allocate the manpower required to maintain them. Others are of the opinion that analyzers are too complex to keep working and prefer to avoid them. Interestingly, there is also a third breed who spend the money to buy analyzers, but do not allocate the manpower to support them.

In audited refineries, the number of analyzers kept working correlated closely with technician time. One refinery was divided into two process areas where maintenance was done by two different people. Both sections had about 20 installed analyzers. In one area the average service factor was over 95% consistently, while in another area two-thirds of the analyzers did not work at all, and the other third had a service factor of 70%. There was no difference in the quality of people or complexity of analyzers. The only difference was that in the well maintained area there was a dedicated analyzer person, while in the other, the analyzer technician also had responsibility for maintaining the instruments, which took most of his time.

Experience regarding distribution of benefits from advanced process control is shown in Table 1. On average, 40% of the benefits rely on our ability to measure product qualities on-line. Some analyzers can be replaced by inferential controls, though the latter are not always available and not on every unit. State-of-the-art models can capture perhaps half the product quality control ben-

Table 1. Approximate distribution of advanced control benefits

Good tuning of basic control loops	20%
Additional stabilization of the unit by feedforwards, etc.	15%
Adding schemes which maximize use of critical equipment	20%
Feedback of analyzer readings to eliminate product quality giveaways	40%
Analyzers can be replaced by good inferential controls or engineering models	(10% to 30%)
Other programs: ramping, switching, etc.	5%

efits without analyzers. Still, at least 20% of advanced control benefits depend on analyzers being properly specified, installed and maintained.

What can be done then to improve benefits associated with quality measurements? The following guidelines should help:

- Purchasing an analyzer implies an investment of $100,000 per installed analyzer and $10,000 per year maintenance. Only those analyzers which are easily justified should be purchased.
- Analyzers whose mechanics of working deviate significantly from laboratory procedures should be treated with suspicion, unless there is a body of evidence to support good correlation between analyzer and lab.
- Avoid analyzers which are a known maintenance headache, such as distillation analyzers working above 700°F or optical analyzers which require frequent cleaning.
- Sometimes product specifications call for difficult analyzers measuring impurities at the ppm level. Often in such units though, there is another stream much easier to measure whose quality trends with the ppm level product quality. It pays to consider these easier, simpler options rather than blindly specifying analyzers on the final products.
- Take a close look at the sample loop designs. Analyzers frequently fail because of dirty, moist or corrosive sample systems, or sometimes simply because of an inadequate fast sample pump or insufficient pressure drop across the fast loop.

Control scheme input validation. When we say that measurements are 99.9% reliable, we accept that they will be off about one hour per month. But when we deal with advanced control schemes measuring from 5 to 50 variables, the off time increases from one hour per month to one hour per week or even one hour per day, becoming a real nuisance for the operator. Worse yet, if every time a measurement becomes problematic the control scheme is turned off, that will result in unreliable unrobust controls causing more trouble for the operator than they are worth. He or she would surely turn your schemes off the minute you walked out of the control room.

The audited refineries attempted to deal with the problem of measurement validation, except it was done in a sporadic, nonsystematic way that resulted in problems:

- The operator never knew for sure which inputs were validated and what were the validation tests. In fact, neither did anyone else. You had to read lines of code to figure it out.
- The operator did not know what, if any, action would be taken upon a signal failure or a signal recovery from failure. Some schemes were turned off, some were left on. The operator always feared a bump in the controls upon a signal recovering from a suspect status.
- When the instrument technician calibrates a flowmeter, the operator knows to set that flow tag to manual, but does not really remember all of the advanced controls which read that flow. These control schemes then bump during flow measurement calibration, causing the operator to panic and turn all advanced controls off.

If you wish to save the operator the trouble of kicking you out of the control room, you have to come up with a reasonable standard for signal validation. For example:

- All pertinent validity tests and test limits should be shown, and always in the same way.
- Noncritical measurements should freeze at the last good value upon failure. The control scheme should continue to work.
- Upon a recovery of that measurement, i.e., it being trouble-free for a certain length of time, it should be accepted and then all schemes reading it should initialize.
- Critical failures should render the scheme dormant for a set time, i.e., outputting stops but the scheme is not turned off yet; it can come alive when the signal recovers. If recovery has not taken place within that time, the scheme should be shut off. From then on only the operator can turn it on.
- The operator should have the ability to turn measurements off and set an estimate instead. Every time that is done, all schemes reading that value should initialize.
- Validity tests and consequences of failing the tests should be established during an operability review of each scheme.

Control technique standards. Most engineers will insist that any piece of equipment installed in the refinery must conform to some design standards, such as the ASME code, TEMA or others. Nevertheless, these same people would readily accept control designs which do not conform to any standard. While there is no official book of advanced control standards blessed by API, there still is a need to create a set of standards or else there is no guarantee of safety, operability or maintainability.

In the audited refineries, there was no documentation or design practices that would even remotely resemble a standard way of doing anything. In one site, there are five process furnaces and each one has a different advanced control scheme. When I asked the lead control engineer the reason for creating five different versions, his answer was typical: "We started out without much experience, but we gradually got better at it. By the time we did the third furnace, we really understood the furnace application. The fourth and fifth furnaces are in a different area and were done by another person. Nowadays, we consider the third furnace the most desirable standard."

We then observed the furnace advanced controls in the control room. Only furnaces three and five were on advanced control. The operators informed us that the other furnace advanced controls have been off for some time because they would occasionally drive the fuel in an unpredictable direction. This was later traced to feedforward signal validation and initialization problems. The working furnaces actually had the same bug, but it had not surfaced yet, so the schemes were kept on.

I then asked to see the standard documentation which would explain, at least for furnace number three, how furnace control was to be done and why, how to tune it, how to display it, etc., but there was no such document. To implement a sixth furnace control scheme, the engineer would need to clone the scheme for furnace three and make necessary modifications. Perhaps some think the cloning procedure is satisfactory, but I consider it a good way to create even more confusion and yet a sixth way of accomplishing the same thing.

Use of engineering models. I have seen functional designs written in the form:

$$CV = A_1\,PV_1 + A_2\,PV_2 + A_3\,PV_3$$

Where CV is some important control variable, A_1, A_2 and A_3 are correlation coefficients, and PV_1, PV_2 and PV_3 are measurements, such as temperatures, pressures and flows. The assumption is that there is an equation that uniquely describes CV, and that while the correlation coefficients are not yet known, it is a trivial regression analysis exercise to find them.

Nothing can be further from the truth. Before considering such a model for control, one needs to verify that:

- CV is only a function of PV_1, PV_2 and PV_3. If there is an additional unpredictable and unmeasured variable, PV_4, the prediction will never work.
- Having proven that CV is only a function of PV_1, PV_2 and PV_3, verify that the linearity of the model holds for the operating range.
- Most engineering calculations are not linear and models which rely on chemical engineering equations are to be preferred (but are not easy to come by).
- It is best to verify that the model actually works before accepting it as a valid functional design.

In the audited refineries there were many models coded as control schemes but few worked. Most model-based schemes were simply turned off and labeled as "problematic," with the exception of one of the refineries which had models turned on and labeled as "working." Being suspicious, I looked at trends of these models and found that several of them were unstable. That instability continued even after detuning related basic loops. I then turned off some models and that stabilized the unit. It seems that two or three models were competing and driving the unit unstable. The remaining models kept the unit stable but there were no data to show that the product quality variability had improved. Among the models which appear useful at first but are actually useless, the most common culprit attempted to predict distillation column bottoms quality from a measurement of bottom temperature, pressure and reflux ratio. Such models are problematic:

- First, the bottom temperature is insensitive to changes in the light key composition. A change of one degree may mean a huge change in quality.
- Second, the bottom temperature is very sensitive to heavy key composition in the feed. That composition then is the unmeasured disturbance, PV_4, which renders the inferential model unpredictable.
- Third, the relation between temperature, pressure, reflux ratio and composition is highly nonlinear and a linear correlation stands no chance of working within the entire operating envelope.

As an antithesis to the oversimplified linear regression model, people sometimes install elaborate on-line simulations which require data to be input manually into the computer. It has been shown statistically that there is one mistake out of 50 numbers inputted, which means such models will never work properly.

And besides, who will type that data? The busy operator? The control engineer? And who will perform the lab tests needed to produce the model data? The busy lab technician? Will they input the data in time to be useful?

The only successful models I have seen are ones that read process measurements and apply simple engineering calculations to come up with a prediction of product quality, preferably without any operator or lab inputs. It is the type of model which treats the process equipment as an "analyzer." These calculations are usually standard API (nonlinear) engineering procedures. Example models which had a high success rate in the audited refineries are:

- True boiling point cut models predicted from column temperatures and partial pressure
- Rvp prediction models.

But even then one has to be careful and avoid the use of models which compete. A second most common model-based control failure was a scheme applying two models to predict the top and bottom purities of distillation column products. The top quality is predicted from rectifying section temperatures, and the bottom quality from stripping section temperatures. This amounts to two cut models which, of course, cannot agree precisely. Because of that disagreement, they sometimes demand an infeasible temperature profile on the column and thus drive it into high reboiler duty plus high reflux and flooding. The tighter the distillation specs, the more likely this is to happen.

Control scheme operability review. It pays to ask hard questions at the functional design stage of an advanced control project. These questions reduce the agony of discovering operational problems by trial and error. All of the refineries audited were in the habit of coming up with scheme designs, implementing and commissioning them, only to find out that these schemes were problematic and often hazardous. They would then modify the design, reimplement and recommission to find the schemes lacking again, etc., going through three or four cycles and ending up with an "afterthought" instead of a proper robust control application.

A committee should be formed to review the operability questions. Review meetings should be conducted twice: first to review the functional design and second just before a scheme is commissioned. The first meeting should consider the refinery economic objective and confirm that it can be reached with the proposed design. The second meeting is for discussion of specific constraint limits, actions to be taken when encountering unusual disturbances, and what precautions should be taken during testing of the new scheme.

If the objective of a scheme is maximization or minimization, the committee must understand the constraints for that piece of equipment. There is often a need to discriminate between a "constraint" and a "myth." An example of a constraint is: If you go beyond a certain limit, a distillation column will flood. An example of a "myth" is: Last time we increased this flow, the unit went unstable. In other words, there may be a real constraint behind the myth but it needs to be understood and explained in engineering terms.

In one of the audited refineries there was a feed maximization control scheme where one of the constraints was the "nominal" thruput of that unit. The control scheme brought the feed flow to its nominal thruput and then stopped. No equipment was operated to its limit, and while the scheme was kept "on" with 100% service factor, it was not making money for the refinery. It was only confusing people by creating that impression. This is what happens when people incorporate myths into the control schemes.

Even when a scheme's aim is stabilization of product quality and it is not expected to be constrained, there is a need to consider the influence of unusual circumstances or disturbances. For example, consider a scheme whose aim is to control sidestream quality in a fractionator. The sidestream is stripped, and quality control is done by manipulating sidestream yield downstream of the stripper. Occasionally, due to a heat duty disturbance or incorrect operation, there would be a shortage of liquid in the fractionator and the stripper level may start dropping quickly. If that disturbance is not recognized by the quality control scheme, a level alarm will soon sound and/or the sidestream pump will cavitate. Whereas if pump protection is built into the control logic, the scheme changes from mediocre to robust.

Having understood the constraints and disturbances, it is necessary to consider the side effects of control schemes. A control scheme may affect the unit in ways other than intended, and while helping operate one piece of equipment it may create problems in another. An example of a scheme which helps one unit at the expense of another is a fractionator pumparound heat duty control scheme where the pumparound is heat integrated with light ends reboilers. In considering such a scheme, one needs to verify that there are enough degrees of freedom, i.e., there is enough slack heat exchange which can be manipulated to hold the total pumparound duty to its target without affecting the light ends reboilers. Second, there is a need to install other schemes to control the reboilers so that when the pumparound heat duty changes the reboiler schemes will decouple the changes to keep the light ends undisturbed. If that is the case, then one of the conditions permitting the pumparound control to run must be that the reboiler control scheme is also working.

Analyzers in advanced control schemes should be known to provide reliable consistent readings. Before ordering an analyzer, verify that it is maintainable; that the sample point and sample loop are such that the deadtime will still be reasonable, and that it is the simplest analyzer to do the job. If the analyzer is already installed, study the repair record and how well the analyzer matches the laboratory.

Even more attention should be given to process models. Models which work well are very valuable, but given the relatively poor rate of success, models should be tested in open loop and compared against the laboratory, particularly those models which have not yet established a track record. If that is not possible at the time of functional design, unproven models should be treated as tentative control schemes and the level of effort of testing and tuning the model should be discussed. Procedures for updating the model in response to operational modifications should be documented.

Finally, one should consider the consequences of erroneous input measurements into the scheme and how to protect the unit from such consequences. The "control scheme input validation" section described a standard method of testing for erroneous measurements. During the operability review meeting decide which measurements are critical to the scheme, what tests are required for the specific measurements and test limits.

Operator interface. Much has been written about the excellent graphics capabilities of modern DCSs. Audited refineries were all using these graphics with P&ID-like displays and live updates, giving the operator a good overview of a section of a plant such as a crude furnace, crude light ends, FCC fractionator, etc. From the operational schematic, the operator can call related group displays or sometimes more detailed schematics and, from the group display, the operator can construct a trend. Operators are comfortable with that variety and they display the appropriate level of detail, using two or more adjacent screens to run the plant.

Where I found the operator interfaces lacking is in providing the operator with good information about the advanced controls. In three of the refineries audited the operator interface design was so poor that there was no way to tell by looking at the operations graphics whether any advanced control scheme was working at all. This is bad because operations graphics are most commonly on the screens, and even the most enthusiastic operator will not turn on functions he or she does not know about. In two of these refineries the advanced controls were implemented on a host computer, and the operator needed to know how to operate the complex keyboard of the host machine. To start an advanced control scheme the scheme display on the host computer screen and the related operational graphics on the DCS screen had to be called. These two displays are seemingly not related but the operator knows better, or does he? These two refineries had the worst service factors for the control schemes.

Fortunately, today's hardware no longer necessitates two different screens to operate advanced control schemes. It is possible to configure dummy DCS tags to serve as an operator interface. The advanced control scheme on the host machine would read the interface tag modes and setpoints, perform the control calculations and output to the DCS tags. Complete transparency is then obtained where the operator can start, stop and operate the advanced controls via DCS tags, and the host computer screen with its cumbersome keyboard is eliminated from the operator console.

The two refineries in question did not implement the more convenient operator interface because the new dummy tags would require additional DCS hardware. However, following the present investigation both refineries will reconsider the costs versus benefits, benefits being a better advanced control service factor.

The third refinery where the advanced control tags were not shown on the operational schematics did not have such an excuse. That refinery uses a DCS with no necessity for dummy tags. But there was a coordination problem. Operational schematics were the responsibility of one person whereas advanced controls displays were done by another. Again, this problem is now being corrected. Operational schematics will show the one or two most important tags of related schemes plus color coding to show when and how the schemes are working. By touching an area on the screen, the operator will be able to call more detailed displays for each scheme. Detailed scheme displays will show the tag structure in detail, plus color coding to provide a clear indication when any part of a scheme is not working properly.

Operator training is an important part of the operator interface. Each board operator needs to understand the objectives of advanced control schemes, constraints, overrides and logical structure. It is a good idea to spend about 30 minutes per shift per control scheme at commissioning time plus repeated sessions if a scheme is modified. This necessity was not well understood in the audited refineries. Engineers spoke down to operators, did not take the time to make sure each operator understood, did not solicit ideas from operators and by and large did not accept that their mission in life is to build tools in service of the operators.

Operators have the final responsibility for safety of the unit, and to that extent they must be permitted to turn control schemes off whenever they perceive a safety or operability problem with the scheme. On the other hand, when the operator turns off a scheme he or she should help by filling out a trouble report form explaining why the scheme was not performing and supporting the report with plots and other data. Once a trouble report is out, it should be treated seriously and some corrective action must be taken. That action may simply be a change of display, more operator training, additional constraint in the scheme or sometimes a more drastic change of logic before the scheme can be recommissioned.

Control engineers training. Training starts with hiring, and while there are misconceptions about the skill requirements for a process control engineer, it is known that people who choose to stay in the field for long-term careers tend to be chemical engineers with process engineering backgrounds. These are the ones who understand the operating variables and how they interact, and the mechanisms for advanced controls to make money. About 80% of advanced control incentives are related to what can be termed "on-line process engineering" or figuring out how the unit should be operated and driving it toward that objective without violating any constraints.

Once the process engineering skills are established, it is of value to the process control engineer to be versed in control theory and there ought to be at least one individual in each refinery who understands both process engineering and control theory.

In the audited refineries it was difficult to find a person with both process engineering and process dynamics skills. The process control groups generally were composed of instrument engineers without process engineering skills, process engineers without knowledge of control theory and chemical engineers who were hired from school or from other industries with an understanding of dynamics but insufficient training in refinery process engineering.

Suggested training starts with a basic course in dynamics and tuning of control loops. The course is very general and intended for new process control engineers as well as instrument technicians and operators. Then follows an intermediate level course to cover the most important advanced control techniques and their application to a process furnace. A third course is more advanced, covering the very difficult problem of controlling distillation columns. Given that this is the most common equipment in a refinery and also the most difficult to control, it is useful to introduce it as early as possible, but not before the participants had a chance to experience the control of simpler equipment. There are also courses for specific units, such as crude unit, FCC or reformer. These courses cover a mixture of process engineering considerations and objectives, plus the typical dynamics of the unit and how to control it. These courses do not have a wide audience and are, therefore, expensive. Still, a control engineer without experience on the unit for which he or she is responsible would benefit greatly from taking such a course.

Performance monitoring. One problem which has hampered progress in process control throughout the industry is lack of monitoring how the advanced control schemes perform once they are commissioned. In the audited refineries people did not know the service factors of schemes, did not have any measure of the performance of advanced control schemes when they are working, and did not know whether the advanced controls actually perform better than no controls. There were endless arguments about when a scheme has been turned off and why, and whether a scheme is really accomplishing its objectives.

It is only through objectively monitoring the performance of control schemes that one can learn how well a scheme is performing and when it is in need of maintenance, and we recommend the following specifics:

1. Service factor. This is the most important single criterion for whether a scheme is working. Monitoring service factor is done by historization of the scheme master tag. The *PV* of that tag is set to 100 when the scheme is working and to 0 when not working. Thus, the daily average of that *PV* is the daily service factor, and the monthly average is the monthly service factor.

2. Ability of the scheme to maintain a process variable at its target. This is an important performance criterion for schemes whose aim is to reduce variability around a given target, such as analyzer feedback control of product quality.

Variability monitoring is accomplished by historization of the value of:

ABS ($SP - PV$)

This value is averaged twice. Once for when the control scheme is working and once when it is not. When a scheme works well, it should exhibit a significant improvement in average deviation.

3. Constraint activity. Constraint controllers are normally designed with soft constraint targets such that when a constraint is active the desire is to keep the process variable at its limit, accepting small violations of that limit about 50% of the time. For that type of constraint controller, monitor the percentage of time the constraint is active. Also, during that time monitor the average approach to target as shown in item 2.

4. Constraint violation. In addition to the soft target of item 3, there is also a harder target we do not wish to exceed. For those constraints monitor the percentage of time there is a violation. Monitor it once for when the control scheme is working and also for when it is not. A well tuned scheme should be able to show a significant improvement in the frequency of violations.

5. Other parameters. It is advisable to historize important setpoints and other critical variables, at least temporarily, to monitor how well the operator handles the scheme and identify need for operator training.

6. Analyzer monitoring. It is a fact of life that analyzers rely on approximate procedures and they do not precisely duplicate official tests. While the analyzer, being a mechanical device, has in general better repeatability, the lab procedure is the legal procedure and therefore, by definition, more accurate. It is advisable to issue a monthly report showing how well the analyzer and lab measurements agree. The same report should also show analyzer validity status history.

7. Inferential model monitoring. Inferential models should be monitored in the same way as the onstream process analyzers. They should be compared against laboratory values and those comparisons should be reported. Reliable inferential calculations will then quickly gain acceptance and be used more often.

8. Given the monitoring of all constraints, important setpoints and control performance, it is of value to come up with formulas for computing the actual financial gain from advanced control. You will then know precisely where you stand and management will be in a better position to support the advanced control effort. ■

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