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Large-Scale Adaptive Multivariable Controllers Eliminate Step Tests and Maximize Profit

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Title: Large-Scale Adaptive Multivariable Controllers Eliminate Step Tests and Maximize Plant Performance
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Abstract:

Deploying adaptive multivariable controllers at two light hydrocarbon plants improved operations and profitability significantly, generating project paybacks within a few months. Traditionally, implementing model-predictive controllers (MPC) in the process industries has required creation of a fixed, linear, dynamic model that relates changes in each input to all outputs. Further, the vast majority of past projects described in the literature have been executed using extensive step tests to develop a linearized control model, using process model identification techniques. Such deliberate tests can be quite costly, disruptive, invasive, and lengthy in duration – often lasting many weeks or months in a large unit. We should mention that both of the plants described herein would have been be very difficult, if not impossible, to subject to extensive step tests in the traditional manner. This is because they suffer very large measured and unmeasured disturbances, have very long settling times, and because the disruption caused by step tests would be quite unacceptable for reasons of safety and product quality.

We describe our experience in using economically optimal, plant-wide, adaptive multivariable controllers for stabilizing operations and improving profit without use of ANY step tests, and present two case histories: a lean oil gas absorption plant, and a refrigerated fractionation complex. These units obtain feed from pipelines and ship products directly to pipelines with no intermediate storage.

We discuss the manifold benefits observed from use of an adaptive control approach in both instances: (1) The adaptive nature of the controller enabled it to maintain excellent performance in the face of changing process dynamics, with very large and frequent feed rate disturbances exceeding 200% within 30 minutes in one of the units. In contrast, fixed model MPC controllers are generally incapable of handling such large disturbances. (2) The particular software package chosen in this effort required only that the user specify the steady-state gains for each manipulated and disturbance variable against each controlled variable. These were determined without use of traditional step tests. (3) The projects were executed with negligible impact on normal plant operations using analytical process simulation models, developed in an office environment, with reliance only on plant design /

historical data to provide the needed MPC controller inputs. (4) The adaptive MPC controllers showed excellent stability and their performance was deemed to be very good, no modeling or other tuning changes being required for over a year, despite huge intervening changes in product values and utility costs in one of the cases. Operator acceptance has been exceptional.

In this work, we developed and calibrated steady-state simulation models against plant data for the processes in question and used these to develop the required MPC controller gains. The chosen MPC controller synthesizes the process dynamics on-line and also makes model adjustments automatically as conditions change. The resulting quality of control was extremely satisfactory compared to pre-project conditions. In both cases cited, a single MPC application was used to achieve economically optimal control of the entire plant, thereby enabling the controllers to adjust operations dynamically to alleviate all constraints.

1. Introduction

This paper describes two kinds of nonlinear control problems that are frequently encountered in chemical processes:

- Dynamic nonlinearity
- Gain nonlinearity

What makes the project execution methodology adopted in this effort unique is that not a single step test was done. The Star multivariable controller, offered by Plant Automation Services, requires the user only to specify steady-state gains for each controlled variable with respect to each manipulated (MV) and disturbance (DV) variable. The dynamic relationship between each controlled variable (CV) and each manipulated (MV) or disturbance variable (DV) is synthesized online. The required steady-state gains were developed using a first principles, steady-state simulation model that was calibrated against plant data. This model was configured to match the control configuration of the plant. Starting from the same base case, each MV or DV in the model was varied systematically, one variable at a time, in both directions and the effect on each CV was captured. This procedure enabled calculating the process gains accurately and in a very straightforward manner. This procedure differs markedly from the traditional plant step testing methods used by others. Such plant step tests can be lengthy, very disruptive, costly, and inconvenient. They also risk product quality and plant safety violations.

As far as we are aware, the projects described in this paper represent the first time that large-scale multivariable controllers have been implemented in the process industries with absolutely no reliance on step tests. The great merit of this approach is that it enables implementing large-scale model predictive, multivariable controllers, starting only with a calibrated plant-wide steady-state simulation model, and achieving all desired project objectives with minimal impact on plant operations. The adaptive nature of the Star controller is unique and it enables handling nonlinear process dynamics in a very straightforward fashion; for example, the effects of large fluctuations in feed flow rates are handled automatically.

2. Dynamic Nonlinearity

This is a very common problem in those processes that are subject to strong fluctuations in process flow rate. As the process flow increases, the process dead time and other dynamic parameters all change proportionately so that the process comes to a steady state sooner. For example, a 200% increase in flow would reduce the dead time to one third of the original value. If a dynamic controller is deployed that uses a fixed dynamic model, such a drastic reduction in process dead time would create a severe mismatch between the process model and the real world. Accordingly, the performance of the controller would be severely impacted owing to excessive reliance on feedback correction to handle the prediction error. Unfortunately, all major currently available commercial software for model-based control, with the exception of the Star controller, uses a fixed dynamic model for describing the process. The Star controller, on the other hand, has proven to be extremely reliable when handling very large changes in process flow rate both when the flow is increasing and when the flow is decreasing.

3. Gain Nonlinearity

Nonlinearity in process gains is far more commonly understood as a control problem in traditional practice. When using linear control models, that is a controller with fixed gains, an attempt is usually made to find some nonlinear transformation that will linearize the gain. For high purity distillation columns, it is common to see use of the logarithm of the concentration of the impurity, rather than the concentration itself. Such a transformation is successful because the logarithm of the concentration of the impurity is fairly linearly related to the reflux flow or the boiler duty. Other nonlinear transforms have been used, for example, for valve positions.

4. Lean Oil Absorption Plant

This is a conventional lean oil absorption plant consisting all the following sections:

1. Inlet separator
2. Dehydration unit
3. Inlet scrubber
4. Refrigerated absorber and sponge absorber (two parallel trains)
5. Rich oil partial demethanizer
6. Demethanizer column
7. Lean oil still
8. Sponge oil/lean oil recovery system
9. Three-stage refrigeration system
10. High, medium, and low pressure steam system

The feed gas to this unit comes from many wells located in the Gulf of Mexico. It is subject to huge fluctuations in both flow rate and composition. To illustrate the magnitude of this problem, the feed flow rates was observed to vary from a low of 180 million standard cubic feet per day (SCFD) to over 550 million SCFD. In addition, disturbances of over 150 million SCFD (up or down) were observed over a period of less than 30 minutes. The plant design capacity was nominally 500 million SCFD. Besides these huge fluctuations in feed rates, the feed composition was also subject to very large disturbances. Since the feed gas analyzer executed at a frequency of about once every 30 minutes, it was virtually impossible to expect that such feed composition fluctuations could be measured in a timely manner. Therefore, feed gas composition could not be made an independent disturbance variable (DV) in this controller.

The lean oil circulation rate has a marked impact on the operation of the unit (Items 4 through 10 above are all affected). This is flow rate and temperature of this huge recycle stream affects the absorption of ethane and propane from the feed gas directly. In addition, changing the lean oil flow affects dramatically the operation of the demethanizer column and lean oil still. Finally, the chilling load on the refrigeration system is affected proportionately as is the demand for high-pressure steam in the refrigeration compressor driver turbine. The composition transients resulting from changes in lean oil flow rate have a long time to steady-state, and these settling times are inversely proportional to the flow rate.

This plant, built in the 1960s, had many problems at the basic regulatory control level. This included improper or incorrect regulatory control loop configuration, missing or inadequate instrumentation, improper analyzer calibration, and poor control loop tuning. Significant attention and effort was devoted to correcting all these problems as a prerequisite, prior to commencing work on the advanced control project. Some details are provided in Appendix A.

A single multivariable controller was built for controlling the entire plant within a single application. The key to improving the profitability of this plant lies in finding the optimal recovery of ethane and propane: too low a recovery is obviously uneconomic, while too high a recovery is not justifiable because of high utility costs. Therefore, the plant-wide economic optimization function in the Star controller seeks to determine the optimal recovery of ethane as a way of maximizing overall plant profit, defined as the value of products minus the cost of feeds and utilities. The Star controller proved capable of handling widely varying economic scenarios, going seamlessly from maximum liquids recovery to maximum rejection, depending only on economics. The same controller gains were used at both ends of the spectrum and the adaptive nature of the Star controller was able to handle very different process dynamics and economic conditions with no operator interference.

The performance of the controller was judged to be exceptional by plant management, and project payout was calculated to be within a few months. The project team was nominated to receive BP's Helios award, in recognition all significant utility consumption reductions, and attained the "commended" stage.

5. Fractionation Plant

This plant receives a liquid feed mixture composed of ethane, propane, butanes and heavier components via pipeline. There are two distillation columns and an electrically driven refrigeration machine:

1. Refrigerated deethanizer
2. Depropanizer
3. Multi-stage refrigeration compressor

This is a fairly modern and well-instrumented plant. Prior to commencing the advanced control project, some changes were required in the basic regulatory control configuration of this plant. For example, the boiler outlet temperature control cascade to the fuel gas was decommissioned and it was decided to control fuel gas flow directly as an MV. Tower pressure control for both columns was tightened significantly. Only minor changes were required in most of the tuning constants for the remaining basic regulatory control loops.

The composition of the feed stream entering the plant is subject to significant fluctuations. Maximizing plant profit generally requires maximizing feed rate subject to existing equipment, process, safety, and product quality constraints. As a result of this requirement, the deethanizer column is often run against its flooding constraint. To enable this to be done reliably, we implemented an online calculation for a new parameter called the "flood factor". This factor accounts for the effects of all major operating variables on tower flooding. The concentration of the impurities at both ends of these two columns is affected strongly by the feed rate, the reflux rate, the reboiler duty, and tower pressure. As might be expected, these steady-state gains are highly nonlinear. Accordingly, we used logarithmic impurity composition transformations to help improve the performance of the controller.

A single multivariable controller was used for this entire plant. Once again, absolutely no plant step tests were performed in order to execute this project. This Star controller application also uses a global profit maximization objective function, defined as the value of products minus the cost of feeds and utilities.

This multivariable control project was also considered by plant management to be highly successful. The controller was able to maximize productivity by pushing feed rates against equipment limits in a safe and reliable manner, while ensuring that product quality was not violated. Here again, project payout was achieved within a few months.

6. Conclusions

These two projects have shown conclusively that following benefits of using a Star adaptive multivariable control technology in highly non-linear processes:

1. Such projects can be executed very successfully without any step tests in the plant whatsoever.
2. PAS' Star adaptive controller requires the user to provide only steady-state gains that can be obtained very reliably from plant-wide steady-state simulation models, taking care to match the control configuration of the plant.
3. The adaptive nature of Star is extremely helpful in enabling it to handle large dynamic nonlinearities caused, for example, by large changes in feed flow rates.
4. Using a single Star controller application for the entire plant is extremely helpful in ensuring that all relevant constraints are incorporated correctly.
5. Maximizing profit requires using a plant-wide, global objective function defined as the value of all products minus cost of all feeds and utilities.
6. Nonlinear transformations should be used as a way of enhancing the range of applicability of the control model gains.
7. Creating online calculations to help predict the onset of tower flooding can be useful when maximizing feed rates.
8. Great attention should be paid to the regulatory control configuration of the plant, and instances of improper configuration should be corrected up front.
9. Plant regulatory controllers must be tuned reliably and robustly to handle a full range of operating conditions likely to be encountered.
10. Effective maintenance of instrumentation, analyzers, and control valves is crucial at all times.

APPENDIX A

It was only after the regulatory control changes described below had been implemented and commissioned successfully that we were able to proceed with the model-predictive control project.

A.1 Instrumentation and Analyzer issues

Since model-predictive controllers (MPC) write to the set-points of the underlying regulatory controllers, it is imperative that the regulatory controls function flawlessly. We found that many required stream flows were not being measured at the plant. As a result, we added almost two dozen new flow meters. Prior to commissioning the MPC applications described in this paper, we spent considerable effort on reconfiguring and tuning the regulatory control loops throughout the plant.

The lean oil plant had a Fisher-Rosemount DCS system that was over eight years old. The first priority was to decommission nonfunctional or obsolete control strategies in order to reclaim “links” needed to commission the new control strategies. New and improved regulatory control strategies were then added that improved the normal functioning of the plant and laid the groundwork for adding MPC applications.

The cycle time of the existing analyzers at the plant was quite lengthy, leading to cycle times of almost 30 minutes. The most critical analyzer measurement is for the methane content in the demethanizer bottoms. Based on analysis of current column performance, it was decided to add a new analyzer with a much faster cycle time for this stream.

A.2 Column Pressure Control

Proper control of tower pressure is an essential requirement for all distillation columns (1). We noticed that several of the distillation columns in this unit did not have any pressure controls at all. Therefore, we spent considerable effort on reconfiguring and tuning tower pressure control strategies. As a result, major diurnal composition disturbances caused by fluctuating tower pressure were eliminated and column operation was improved considerably.

A.3 Process Equipment Problems

We observed that the demethanizer tower bottoms methane content was subject to huge fluctuations and disturbances. These occurred even when feed to the unit was steady. Therefore, we decided to perform a detailed tray hydraulic analysis for all trays in the column. It was found that even at 100% of design capacity, the trays in the bottom section of this column were subject to excessive weeping, with over 70% of the liquid falling through the holes on the sieve trays. This problem was attributed to the use of a half-inch hole diameter, which was considered excessive. It should be pointed out that this tower was designed and built in the 1960s, a time when much of the currently available information concerning tray weeping was not available.

The only option to correct this problem was to re-tray the entire column. Therefore, new trays were installed and column performance improved dramatically. We also decommissioned a reboiler outlet vapor temperature control that adjusted the steam to the reboiler and replaced it with a simple steam flow controller. This change enabled much steadier tower operation.

A.4 Compressor Anti-surge Controls

The multistage compressor for the refrigeration machine in the lean oil plant provides feet chilling and demethanizer condenser refrigeration. Owing to a regulatory control configuration error, it was found that the controller action for the speed control was the opposite of what is required. As a result, the speed controller for the refrigeration machine was being run in manual mode. When the refrigeration load diminished as feed rates fell, it was necessary for an operator to reduce machine speed very quickly, for fear of driving the machine into surge. Correcting the speed controller configuration error, and retuning this controller eliminated these problems dramatically.

A.5 Regulatory Control Loop Configuration and Tuning

The ability to run the process smoothly depends heavily on the proper configuration of all regulatory control loops for the manipulated variables (MVs) in the model-predictive controller. This challenge was quite acute at the lean oil absorption plant, in particular, as the unit was built over 35 years ago, had not seen much attention being paid to the control hardware and distributed control system, and was suffering from the lack of many essential measurements.

After these issues were corrected, there still remained the problem of improper regulatory control loop configuration. Identifying such problem areas and devising the proper solutions was a significant effort, especially since the DCS was overloaded with a number of obsolete and non-functional strategies. These were first decommissioned and eliminated so as to free up “links” that were needed to add the many new on-line calculations required to create additional controlled variables (CVs).

Additionally, we tuned all the control loops carefully for stability during widely varying loads and operating conditions. In many instances, entirely new control strategies were devised to enable a more logical and consistent approach to plant operation than had been possible in the past.