An approach to the optimization of an olefins plant

by S. REITER, D. J. SVETLIK and A. M. FAYON
Mobil Oil Corporation
New York, New York

INTRODUCTION

Mobil Chemical Company has operated an olefins plant in Beaumont, Texas since 1960. This paper discusses the approach taken by Mobil Chemical Company in the optimization of the operation of the olefins plant.

A linear programming (LP) model is used for long and short range planning purposes while open loop on line optimization via an 1800 process control computer is used by plant management for aid in the determination of day to day operating policies.

The heart of the olefins plant is the pyrolysis furnaces. The furnaces yield, from basic feedstocks, the desired olefin products. The processing units downstream of the furnaces separate the furnace effluent stream into the desired high purity olefin product streams.

The olefins plant is capable of processing a large number of feedstocks which range from ethane to naphtha and raffinate. The feeds are obtained primarily from the Mobil Oil refinery in Beaumont. Outside purchased stocks are also available for processing. Each feedstock has associated with it a yield pattern for ethylene and heavier olefins. Depending on market conditions and other factors, at any given time, feedstocks become relatively more or less desirable. Within the yield pattern associated with a given feedstock there is a specific furnace severity, defined by furnace temperature and pressure and hydrocarbon and steam flow, which defines the optimum processing conditions for a specific feedstock.

The two questions raised here, that is;

• what feedstock selection to make? and
• at what conditions selected feedstocks are to be processed?

led to the development and implementation of the systems in use today.

The overall structure of the linear programming model will be reviewed with a discussion of its application to planning decisions.

The hardware and software structure of the 1800 system will be discussed and specific functions associated with it reviewed. The nature of the optimization procedure employed and the structure of the online process model and its adaptive characteristics will also be discussed.

OLEFINS PLANT LINEAR PROGRAMMING MODEL

Structure

The olefins plant LP model is an economic and process model. It consists of linearized representations of the operations of process units with explicit representations of the key constraints associated with these units. Most importantly, it contains representations of the flexibility associated with feedstock purchasing and product sales. A simplified structure of the LP model is shown in Figure 1.

The first operation shown is purchasing. As many as six potential feedstocks may be available for purchase over a given time. Typically the furnace feedstocks available are: ethane, propane, n-butane, isobutane, naphtha, and raffinate. Each of these feedstocks has availabilities associated with it at various prices. In addition to furnace feeds, two additional feedstock availabilities are represented. These are refinery gas, a stream consisting of ethylene and lighter components, and a butadiene concentrate stream, consisting of butene and butadiene. Both these streams are introduced into the process downstream of the furnaces.

The next operation shown is the pyrolysis furnaces. Physically 22 furnaces are available. Due to decoking operations 19 furnaces are usually on-stream at any one time. As many as five yield slates may be associated with a furnace feedstock, as required to linearize the feed stocks yield pattern. Although severity is a multi variable function, a key control parameter is temperature, and in the LP, furnace yields are represented as functions of temperature only. Associated with each feedstock is a furnace stream factor. The stream factor defines the normal mass flow associated with a given feed-stock and is used in defining the maximum furnace availability constraint.

The remaining process operations shown are process gas compression and hot and cold train distillation. Linearized representations of these units are used to derive constraints associated with compressor horsepower requirements, and feed and overhead rate limits on various downstream distillation towers.

The last operation shown is sales. Four olefin product streams, ethylene, propylene, butadiene, and butylene, at
one or more pricing levels are represented as well as fuel gas and an aromatics concentrate stream.

Application

The range of applications associated with the LP are:

1. Profit plan and five year plan preparation.
   - Based on anticipated prices and feedstock availabilities a determination of the optimal way to operate the plant for a period of one to five years is developed for budget and profit plan preparation.
   - Based on the above, sales and purchase commitments for periods covering one to five years may be developed.

2. Short Range Planning.
   - Based on predefined feed and product prices optimal feed slates, and product mixes are determined and first pass estimates of furnace severities are established.

3. Equivalent pricing of alternate feedstocks.
   Due to factors, not controllable by Mobil Chemical, planned feedstocks may become unavailable and an alternate feedstock must be selected. The LP model will be used to determine:
   - Which alternate feedstock should be selected;
   - The required availability and price of the alternate feed stock to maintain a constant profit; and
   - The desirability of purchasing outside product streams to meet sales commitments.

4. Effect of debottlenecking on profit.
   The LP is used to assess the overall desirability of implementing recommendations developed from offline simulation studies aimed at debottlenecking.

As indicated above the results of the LP solution provide an indication of the optimal furnace severity to be associated with each feed stock. This severity is a first approximation. It is developed from a linearized economic model which provides adequate results for long and short range planning purposes but which leaves much to be desired in setting actual day-by-day operating conditions. Furnace yields represented in the LP are linearized for typical feedstocks. The effect of feed composition variation is not accounted for. Tower performance is represented by linearized or constant split factors which do not consider tower feed rate or composition variations. Process constraints are expressed as the expected or typical limitation associated with a given unit. Some constraints are implied rather than explicitly represented. As previously indicated furnace severity is equated to operating temperature with hydrocarbon and steam rate and pressure assumed constant.

These approximations are valid for the planning decisions to which LP results are applied. For operating decisions aimed at maximizing plant throughput on a daily or weekly basis a more rigorous representation of the process and its variability or non-linearity is required. It was with the objective of providing such a decision tool that an online control computer system was installed.

OLEFINS PLANT CONTROL COMPUTER SYSTEM

Overview

The justification for the installation of an online control computer was based on improved plant performance achievable through optimization. Since installation of the system additional benefits to operations have been achieved.

The functions performed by the control system can be placed in one of four categories, these are:
- Operations and Management Information Reporting;
- Supervisory Control;
- Operations Support; and
- Optimization.

Before reviewing each of these functions the hardware configuration and system software associated with the control computer will be discussed.

Hardware

Figure 2 is a simple schematic of the computer system hardware. The system consists of an 1800 computer with a 2-mu second core of 32,000-16 bit words. Secondary storage is provided for by a 2-disk storage unit. Operations and systems support input/output facilities are serviced by an IBM model 10 Process Operators Console (POC) and three 1053 typer and a card reader/punch.
Physically, two typers and the POC are in the Olefins plant control room while the main frame and remaining peripherals are in an adjacent computer room.

One typer and the card reader are used primarily by the control group for system maintenance functions. The POC and remaining two typers provide the interface facilities for the Operating Departments' personnel.

The system monitors 362 process input signals and exercises supervisory control over 113 set point stations. With feedbacks, a total of 585 process signals are interfaced with the control system. All input signals are multiplexed and are on a four minute scan cycle. Hardware comparator limit checking is continuously carried out for the detection of out of range signals.

Software

The system operates under TSX. Core is segmented into a fixed and variable core region. Fixed core is approximately 12,000 words and consists of core resident system and control programs and incore tables. Variable core is 20,000 words and is used for applications and control programs with relatively low utilization frequencies.

PROSPRO I is utilized for process I/O interface functions. This includes:
- Process variable scanning;
- Analog to digital conversions;
- Digital to engineering units conversion;
- Control of pulse out signals to set point stations; and,
- Out of limits and signal rate deviation alarm message generation.

POC software operates under TSX. The following functions are provided for:
- Display of selected process variables;
- Change set point and variable min/max limit values;
- Initiation of various log reports; and,
- Initiation of various control system functions, such as the optimization system.

TSX, PROSPRO, and the POC programs are the main system software programs.

Other major software systems, related to applications, include:
- The optimization and process model programs; and
- The furnace start-up/shut-down programs.

These will be discussed under the control system functions they are associated with.

Control System Functions

The four control system functions associated with the olefins plant are: (1) Operations and Management Information Reporting; (2) Supervisory Control; (3) Operations Support; and (4) Optimization.


Many control system installations have been justified on the basis of Management Information Reporting. Whatever the justification for a control system, the data collection function performed will always lead to the ability to selectively generate various process reports. Proper transformation of this raw data can provide more useful information to operating personnel.

A number of reports are generated for the operating department and management at the Olefins Plant. These include:
- Furnace log reports;
- Distillation Train reports;
- Compressor logs;
- Plant Material Balance Reports.

The furnace log report will identify for each furnace the feed associated with it, the feed rate, steam to oil ratio, the effluent temperature, the furnace box wall temperature and the quench boiler’s outlet temperature. The number of days the furnace has been online and an indication of the degree of coking associated with the furnace are also provide.

The material balance report displays the rate and composition of all feed and product streams associated with the plant and an indication of the percent of material balance closure. This report can provide a four or twenty four hour plant material balance.

The other reports contain raw data as well as computed values of key non-measurable parameters associated with the plant sections they are related to. All reports are either generated on a fixed time cycle or on demand via function keys available on the Process Operator’s Console.

2. Supervisory Control

Supervisory control functions are currently carried for furnace mass flow compensation and internal reflux control.
Furnace flow consists of both hydrocarbon and steam. The steam flow is ratio controlled to the hydrocarbon furnace feed. The flow orifice coefficient on each furnace stream is periodically updated. This assures that measured flows are accurate, and set point mass flows, which are determined via the control system optimization programs, are correctly maintained.

Internal reflux control keeps key towers operating near flood conditions to maximize product recovery.

3. Operations Support

Several programs have been developed to aid the operating department in repetitive tasks. One of these is the furnace start-up/shut-down program.

The Olefins plant consists of 22 cracking furnaces. On the average 19 furnaces will be online at any time while the remaining three are being decoked. The length of online service for any furnace is a function of the feedstock it is associated with and its operating temperature and steam to oil ratio. Typically, within a given week at least one furnace will be taken out of service while another is brought into service as a replacement. The task of smoothly shutting down a coked furnace and simultaneously bringing online a decoked furnace has been automated via a furnace start-up/shut-down program.

Several benefits have been associated with this program, these are:

- Reduced downstream disturbances in process flows during furnace start-up operations;
- Relieving operators from the task of continuously monitoring furnace start-up operations;
- Elimination of rapid coking during start-ups with resulting short furnace runs; and
- Aid in gaining operator acceptance of the control system by demonstrating the system’s ability to free them from repetitive time consuming tasks.

4. Optimization

a. The System

The optimization system in use consists of two major programs. A Control Optimization Program, COP, which is an IBM developed general purpose nonlinear optimizer and a Fortran written process model developed by Mobil.

COP is an assembly language program which employs the method of Sectional Linear Programming to solve a non-linear optimization problem. To understand the structure of the overall optimization system employed it is important to briefly review the working of the optimization method employed by COP and its relation to any non-linear system model which is to be operated upon. A more detailed discussion of COP can be found in Reference 1.

Given a set of independent and dependent variables which represent a given state of some system model, the optimization technique employed linearizes the constraints and objective function about this starting point. This is accomplished by developing a differential relation between each dependent or constraint variable and the objective function with each independent or controllable variable. The differentials are developed by making an incremental change in one independent variable and using the model to determine the effect on each dependent variable and the objective function.

These dependent variable and objective function changes are determined for each independent variable in the system.

The differential set of linear relationships that are developed are solved using linear programming techniques to determine the factor by which each independent variable should be incremented or decremented to improve the value of the objective function.

The overall optimization is accomplished in successive steps which move the model from an initial position to an optimal position.

Each step entails the following:

- Linearization — i.e. The development of the differential relationships between independent and dependent variables;
- Formulation and solution of the differential linearized LP problem;
- Calculation of the moves to be associated with each independent variable based on the LP solution; and
- Linearity error correction to determine the actual model position.

In practice, with the olefins plant process model, it has been found that 12 to 15 steps are typically required to move the model from an initial position to a predicted optimum solution. Approximately eight to twelve minutes are required for the completion of each step calculation with 50 percent CPU time available for optimization.

b. The Process Model

The olefins plant process model is a Fortran program consisting of a number of subroutines. The overall structure of the model is shown in Figure 3. There are one or more subroutines provided to stimulate the operation of each processing unit shown in Figure 3.

The first unit shown is the pyrolysis furnaces. The calculation of the yield for any feed stock is based on the conversion of that feed or an indication of conversion. For light feedstocks, Ethane through Butane, volumetric conversion is developed to calculate component yields. For Naphtha and Raffinate, methane yield is used as a measure of conversion to calculate the yield of the remaining furnace effluent components. The feedstock conversion is a correlated function of furnace severity. The severity in turn is a function of hydrocarbon and steam flow, and the furnace exit temperature and pressure. Due to volumetric changes in the pyrolysis furnace the conversion calculation is iterative. The yield curves associated with each feedstock are determined by furnace effluent analysis. Originally, a furnace effluent
sampling system was installed, however, due to continuing maintenance difficulties, online yield analysis never proved to be a workable system for updating yield relationships.

In summary the procedure used to compute feedstock yields involves:

First—the calculation of a furnace severity;
Second—the calculation of feedstock conversion;
Third—updating severity with calculated conversion to account for volumetric changes; and,
Finally—the calculation of yields based on conversion for light feedstocks and methane yield for naphtha and raffinate.

In addition to yields, the furnace subroutine calculates fuel requirements and furnace box wall temperature. Box wall temperature is a constraint to the overall optimization.

The next unit shown is the quench boiler. The quench boiler calculation will determine the suction pressure to the first stage process compressor and the steam generation resulting from the furnace effluent. There is a maximum limit on the quench boiler's outlet temperature due to material considerations associated with downstream piping. This outlet temperature limit is another constraint to the optimization program.

The next unit shown is the four stage process gas compressor. Design head capacity curves are utilized to calculate power requirements and discharge pressures for each stage. Second stage suction pressure and overall compressor power requirements are developed as constraints to the optimization.

The fourth stage flash and prefractionator are the next downstream processing units. These units are extremely important to the operation of the hot and cold train distillation towers which follow. The split of C2 and C3 Components and the load placed on downstream towers is determined from the furnace effluent. There is a maximum limit on the quench boiler's outlet temperature due to material considerations associated with downstream piping. This outlet temperature limit is another constraint to the optimization program.

An optimization calculation is initiated on an eight hour cycle or on request by operations. When an optimization is initiated, the first step required is the calculation of the plant's present position. This calculation serves two functions:

First—It develops the initial set of independent and dependent variables required by COP; and,
Second—It develops various tuning factors based on measured plant inputs to provide for adaptive tuning in the process model.

The general nature of the tuning carried out is the form:

\[ Y = F(X) \cdot \text{Constant} \]

i.e.,

The parameter to be calculated = (a function of known variables) times (a constant)

Since the control system in many cases is monitoring actual plant data associated with both model dependent and independent variables, the actual value of the parameter to be calculated and its associated independent variable
values are known. The constant term can then be determined as the ratio of the measured parameter to its calculated value. This constant term is a tuning factor which accounts for process variability with time. This type of tuning is carried out for the following process units:

Furnaces: Used to determine the degree of coking based on heat transfer calculations and is applied to the determination of the furnace box wall temperature.

Quench Boilers: Adjusts clean surface heat transfer coefficients to take into account fouling for steam generation and discharge temperature calculations. In addition, the effect of restricted flow area on the pressure drop calculation is assessed.

Compressor: Develops factors to account for errors in simplified compressor head-flow relations and to take into account degradation in compressor performance due to fouling.

These adaptive coefficients are held constant throughout the optimization. This is an approximation of course, since plant conditions will change as the process is moved from one point of operation to another. In practice, this has not been found to be a difficulty since at any time only small process moves are made and optimizations are carried out on a repetitive cycle. Although analysis of tower product streams are made in the plant, the compositions are not directly available to the computer. This precludes the possibility of carrying out adaptive online tuning of tower performance. Data files containing parameters associated with tower calculations are accessed by operating personnel as required to update these parameters from laboratory analysis of tower product streams.

The optimization system employed provides directional guidance in the setting of controllable variables to improve overall profit. In practice there was found to be considerable hesitancy on the part of the operating department's personnel to except optimizer results when the absolute values of model and optimizer flow predictions differed with measured plant flows. This problem was brought about by material balance errors associated with the plant metering system and unaccounted for material losses.

A real problem presented in the discrepancy between model predicted flows and measured flows was in knowing the correct value to set flow constraints.

The difficulty was rectified by introducing into the model a furnace effluent biasing stream. This stream was calculated by determining the difference in the furnace effluent as determined by the plant material balance metering system and the effluent predicted by the process model. This biasing stream was then held constant throughout the optimization. It achieved the objective of biasing the flow of each component leaving the furnace to the levels measured by the plant metering system without affecting the slopes of the non-linear correlations used by the optimization program to develop the required differentials.

As mentioned, some inputs required for model tuning are available via the control system while others must be provided by plant personnel. A large number of parameters are associated with the Olefins plant control system that require frequent updating. Only some of these parameters are related to the model tuning function. In order to place the responsibility of maintaining data integrity with the Operating Department a Control Interface System was developed. This system allows key parameters to be stored in data files and provides for easy access to these files by operating department personnel for either inspection or change functions. These parameters are utilized by various control system programs.

Briefly the information stored relates to:

- Feed compositions for use in T-P-C compensation of furnace feeds;
- Product stream compositions for use in material balance reporting;
- Limits on the ranges of independent and dependent variables used in the optimization;
- Feed, product and utility costs required for the optimization; and
- Furnace feed identifications for use in various plant reports.

Figure 4 presents an overview of all applications software related to the Olefins plant control system. In summary this entails:

- Management and Operations Information Reporting;
- Supervisory Control;
- Operations Support; and
- Optimization.

CONCLUSION

In conclusion, Mobil Chemical Company has developed and implemented a dual approach toward the optimization of its olefins plant.
A linear programming model is used for long and short range planning studies. This is an economic model which represents the flexibility available in feed stock selection and in establishing product mixes. Linearized representations of process units are used.

Benefits related to the use of this planning tool are associated with the quantitative procedures it affords in evaluating alternate choices related to long and short range planning decisions.

An online Control System provides for open loop optimization; the results of the optimization provide operating department personnel with day to day guidance in establishing optimum processing conditions. Additional benefits attributed to the control system are associated with management and operations reporting functions, stabilization of plant flows and various operations support functions.

REFERENCE
