SUPPLY CHAIN OPTIMIZATION OF PETROLEUM REFINERY COMPLEXES

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Abstract

Production planning is a very important step in managing the operation of a single refinery and their complexes. Despite its significance, few contributions have been found in literature and the existing ones rely on linear models. Pinto *et al.* (*Comp. Chem. Engng.* 24, 2259-2276, 2000), presented a superstructure that represents a general topology and allows the implementation of nonlinear process models as well as blending relations. The main objective of the present work is to extend the single refinery model to a corporate planning model that contains multiple refineries, which can be connected by supply pipelines in common. Intermediate streams also interconnect refineries in order to take advantage of each plant infrastructure. The model is optimized along a planning horizon resulting in a large scale Mixed Integer Nonlinear Program (MINLP). The non-linearity arises from blending equations and physical properties. The objective function maximizes net present value under raw material and product inventory level constraints as well as mass balance and operating constraints in each refinery. Finally, detailed analysis for different crude oil types and product demand scenarios is incorporated in the model. A real-world application is developed for a refinery network composed of three refineries. Different petroleum types are supplied to the refineries from a single oil terminal. Results show that the optimization of the supply chain presents clear advantages of the corporate planning with respect to multiple one-site refinery production planning.

Keywords

Supply Chain Optimization, Petroleum Industry, Refinery Complex, MINLP.

Introduction

Companies have been forced to overstep their physical frontiers and to visualize the surrounding business environment before planning their activities. Range vision should cover all members that participate direct or indirectly in the work to satisfy a customer necessity. Coordination of this virtual corporation may result in benefits for all members of the chain individually. Beamon (1998) defines such virtual corporation as an integrated process wherein a number of business entities (suppliers, manufacturers, distributors and retailers) work together in an effort to acquire raw materials, convert them into specified final products and deliver these final products to retailers. Under another point of view, Tan (2001) states that there is a definition of supply chain management (SCM), which emerges from transportation and logistics literature of the wholesaling and retailing industry that emphasizes the importance of physical distribution and integrated logistics. According to Lamming (1996), this is probably where the term supply chain management was originally used.

According to Thomas and Griffin (1996), current research in the area of SCM can be classified in three categories: Buyer-Vendor, Production-Distribution and Inventory-Distribution coordination. The authors present an extensive literature review for each category. Vidal and Goetschalckx (1997) present a review of Mixed Integer Problems (MIP) that focuses on the identification of the

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relevant factors included in formulations of the chain or its subsystems. The same authors also highlight solution methodologies.

Bok *et al.* (2000) present an application to the optimization of continuous flexible process networks. Modeling considers intermittent deliveries, production shortfalls, delivery delays, inventory profiles and job changeovers. A bilevel solution methodology is proposed to reduce computational expense. Zhuo *et al.* (2000) introduce a supply chain model that involves conflicting decisions in the objective function. Goal programming is used to solve the multi-objective optimization problem. Perea *et al.* (2000) and Perea-López *et al.* (2001) present an approach that is capable of capturing the dynamic behavior of the supply chain by modeling flow of materials and information within the supply chain. Information is considered as perturbation of a system control whereas material flows are considered to be control variables. Therefore, this approach is able to react on time and to coordinate the whole supply chain for changing demand conditions. Song *et al.* (2002) present a design problem of multiproduct, multi-echelon supply chain. Transportation cost is treated as a continuous piecewise linear function of the distance and a discontinuous piecewise linear function of the transportation volume, whereas installation costs are expressed as a function of the capacity. Feord *et al.* (2002) propose a network model whose main objective is to decide which orders should be met, which delayed and which are no to be delivered.

The petroleum industry can be characterized as a typical supply chain. All levels of decisions arise in such a supply chain, namely, strategic, tactical and operational. In spite of the complexity involved in the decision making process at each level, much of their management is still based on heuristics or on simple linear models. Therefore, systematic methods for efficiently managing the petroleum supply chain must be exploited. In the next section, the petroleum supply chain scope is described as well as recent developments found in the literature concerning its subsystems.

Petroleum Supply Chain

The petroleum supply chain is illustrated in Figure 1. Petroleum exploration is at the highest level in the chain. Decisions at this level include design and planning of oil field infrastructure. Oil tankers transport petroleum to oil terminals, which are connected to refineries through a pipeline network. Decisions at this level incorporate transportation modes and supply planning and scheduling. Crude oil is converted to products at refineries, which can be connected to each other in order to take advantage of each refinery design within the complex. Products generated at the refineries are then sent to distribution centers. Crude oil and products up to this level are often transported through pipelines. From this level on, products can be transported either through pipelines or trucks, depending on consumer demands. In some cases, products are also transported through vessels or by train.

In general, production planning includes decisions such as individual production levels for each product as well as operating conditions for each refinery in the network, whereas product transportation focuses on scheduling and inventory management of the distribution network.

Products at the last level presented in Figure 1 are actually raw materials for a variety of processes. This fact indicates that the petroleum supply chain could be further extended. However, this work deals with the petroleum supply chain as shown in Figure 1.

Sear (1993) was probably the first to address the supply chain management in the context of an oil company. The author developed a linear programming network model for planning the logistics of a downstream oil company. The model involves crude oil purchase and transportation, processing of products and transportation, and depot operation. Escudero *et al.* (1999) proposed an LP model that handles the supply, transformation and distribution of an oil company that accounts for uncertainties in supply costs, demands and product prices. Dempster *et al.* (2000) applied a stochastic programming approach to planning problems for a consortium of oil companies. First, a deterministic multiperiod linear programming model is developed for supply, production and distribution. The deterministic model is then used as a basis for implementing a stochastic programming formulation with uncertainty in product demands and spot supply costs.

Important developments of subsystems of the petroleum supply chain can be found in literature. Iyer *et al.* (1998) developed a multiperiod MILP for planning and scheduling of offshore oil field infrastructure investments and operations. The nonlinear reservoir behavior is handled with piecewise linear approximation functions. A sequential decomposition technique is applied. Van den Heever and Grossmann (2000) presented a nonlinear model for oilfield infrastructure that involves design and planning decisions. The authors consider non-linear reservoir behavior. A logic-based model is proposed that is solved with a bilevel decomposition technique. This technique aggregates time periods for the design problem and subsequently disaggregates them for the planning sub-problem. Van den Heever *et al.* (2000) addressed the design and planning of offshore oilfield infrastructure focusing on business rules. A disjunctive model capable to deal with the increased order of magnitude due to the business rules is proposed. Ierapetritou *et al.* (1999) studied the optimal location of vertical wells for a given reservoir property map. The problem is formulated as a large scale MILP and solved by a decomposition technique that relies on quality cut constraints. Kosmidis *et al.* (2002) described an MILP formulation for the well allocation and operation of integrated gas-oil systems whereas Barnes *et al.* (2002) focused on the production design of offshore platforms.

At another level of the supply chain, Lee *et al.* (1996) concentrated on the short-term scheduling of crude oil supply for a single refinery. Más and Pinto (2002) developed a detailed MILP formulation for the optimal scheduling of oil supply comprised of tankers, piers, storage tanks, substations and refineries. Pinto *et al.* (2000) and Pinto and Moro (2000) focused on the refinery operations. The former work focuses on production scheduling for several specific areas in a refinery such as crude oil, fuel oil, asphalt and LPG whereas the latter addresses production planning. Ponnambalam *et al.* (1992) developed an approach that combines the simplex method for linear programming with an interior point method for solving a multiperiod planning model in the oil refinery industry. Still at the production planning level, Liu and Sahinidis (1997) presented a fuzzy programming approach for solving a petrochemical complex problem involving uncertainty in model parameters. Bok *et al.* (1998) addressed the problem of long-range capacity expansion planning for a petrochemical industry.

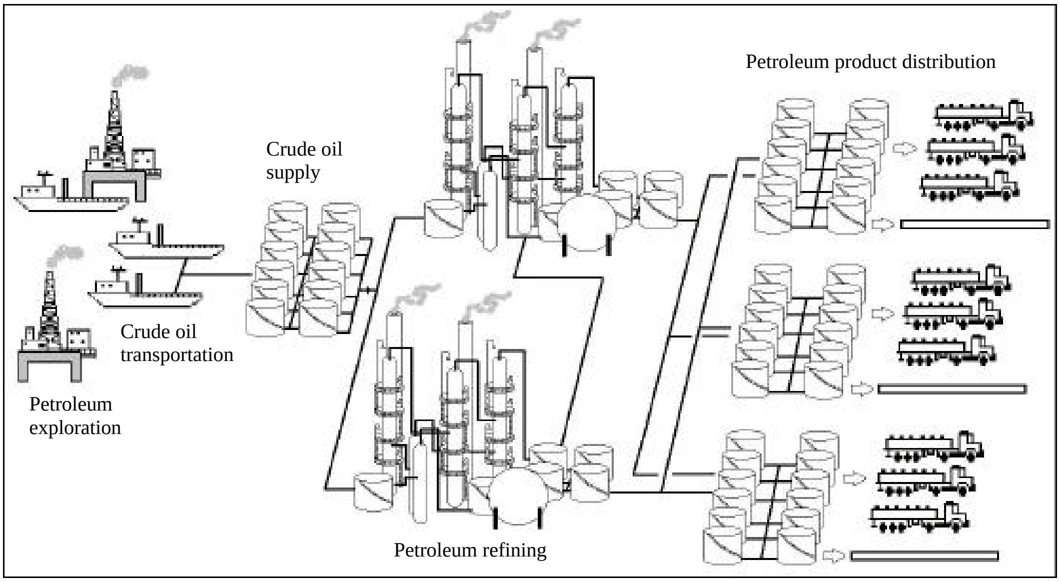

Figure 1. Petroleum Supply Chain.

Ross (2000) formulated a planning supply network model on the petroleum distribution downstream segment. Resource allocation such as distribution centers (new and existing) and vehicles is managed in order to maximize profit. Delivery cost is determined depending on the geographic zone, trip cost, order frequency and travel distance for each customer. Iakovou (2001) proposed a model that focuses on the maritime transportation of petroleum products considering a set of transport modalities. One of the main objectives of this work was to take into account the risks of oil spill incidents. Magatão *et al.* (2002) propose an MILP approach to aid the decision-making process for schedule commodities on pipeline systems. On the product storage level, Stebel *et al.* (2002) present a model involving the decision making process on storage operations of liquefied petroleum gas (LPG).

As can be seen, only subsystems of the petroleum supply chain have been studied at a reasonable level of detail. The reason is the resulting complexity when parts of the chain are put together within the same model. Nevertheless, logic-based approaches have shown potential to efficiently model and solve large systems without reducing problem complexity (Türkay and Grossmann, 1996; Van den Heever and Grossmann, 1999; Vecchietti and Grossmann, 2000). This fact, allied to the development of new powerful computers and changing business necessities provide motivation to increase the scope of petroleum supply chain modeling.

The main objective of the present work is to extend a single refinery planning model to a corporate model that contains multiple refineries, which are connected by supply pipelines in common or interconnected by intermediate product streams. An example of a connection among refineries is the use of a unit to reduce sulfur levels of streams from other refineries that do not present this unit. The model is optimized along a planning horizon resulting in a large scale Mixed Integer Nonlinear Program (MINLP). The approach is demonstrated in a real-world application for a refinery network composed of three refineries.

The paper is organized as follows: the problem statement is given in the next section, followed by the mathematical model. An illustrative example of a single refinery is presented in order to clarify the model ideas and results of the real-world case study involving a petroleum complex are then presented and discussed. Finally, conclusions and future research are discussed in the last section.

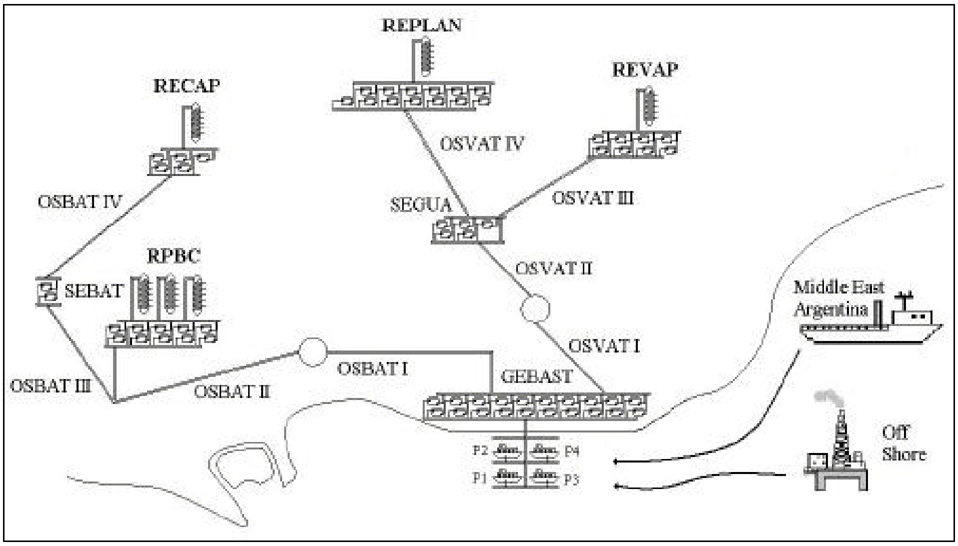

Figure 2. Supply Chain – Case Study.

Problem Statement

Petroleum supply usually involves a harbor, which contains a number of piers and each of them is able to receive vessels with different capacities. Petroleum types from different sources are discharged and stored in tanks at a terminal. Refineries are usually located near customer markets. Petroleum must be supplied to refineries through a pipeline network, since their demands are often high. Every element in such a system contains its own infrastructure, i. e., discharge, storage and pumping components.

Figure 2 illustrates a real-world petroleum supply chain such as the one considered in the present work. Petroleum is acquired from either offshore platforms or from foreign suppliers. Vessels discharge the crude oil at an oil terminal, namely GEBAST. Since petroleum properties depend on supplier origin, there exist dedicated tanks for petroleum types. Refinery supply is accomplished through two main pipeline branches: OSBAT and OSVAT. The OSBAT branch feeds RPBC and RECAP refineries. SEBAT is an intermediate oil terminal connecting OSBAT III and OSBAT IV segments. The OSVAT branch feeds REVAP and REPLAN refineries, which can be operated in three different ways: 1) crude oil is directly pumped from GEBAST to REVAP through OSVAT I, OSVAT II and OSVAT III pipelines; 2) crude oil is directly pumped from GEBAST to REPLAN through OSVAT I, OSVAT II and OSVAT IV pipelines. In this case maximum flow rate must respect OSVAT IV pumping capacity. 3) Crude oil is pumped from GEBAST to SEGUA at the maximum flow rate of the corresponding linking segment. The oil is temporarily stored at the SEGUA terminal and then transferred to REPLAN at its maximum flow rate capacity. In this case, SEGUA is used as a buffer. Appropriate mixing is performed at GEBAST before transportation in the case that refinery feed is composed by more than one crude oil type.

Mathematical Model

Network modeling is based on the representation proposed by Pinto and Moro (2000). Figure 3 illustrates the general representation of a unit and its corresponding variables.

As can be seen from figure 3, stream s_1 from unit u_1 is sent to unit u at flow rate $Q_{u1,s1,u}$. The same unit (u_1) can send a variety of its outlet streams to unit u, $\{s_1, s_2,\ldots, s_{NS1}\}$; moreover, a set of units $\{u_1, u_2,\ldots,u_{NU}\}$ may feed unit u. The resulting feed for unit u is denoted by QF_u. Mixture is always accomplished before feeding. Every stream is characterized by a set of properties $\{p_1, p_2,,\ldots,p_{NP}\}$. Variables $PF_{u,p}$ and $PS_{u,s,p}$ denote properties of the inlet and outlet stream, respectively. The unit feed is converted into a set of products $\{s_{1o}, s_{2o},\ldots, s_{No}\}$. Variable $QS_{u,s}$ represents the outlet flow rate of every stream s leaving unit u. Since a product stream can be sent to more than one unit for further processing or storage, a splitter is represented at every outlet stream. Different outlet streams are characterized by specific property sets.

In the present work the original variables in Pinto *et al.* (2000) are modified, by adding subscripts that denote time periods and scenarios. Inventory variables are also considered as well as connecting variables between refineries. The model relies on the following notation:

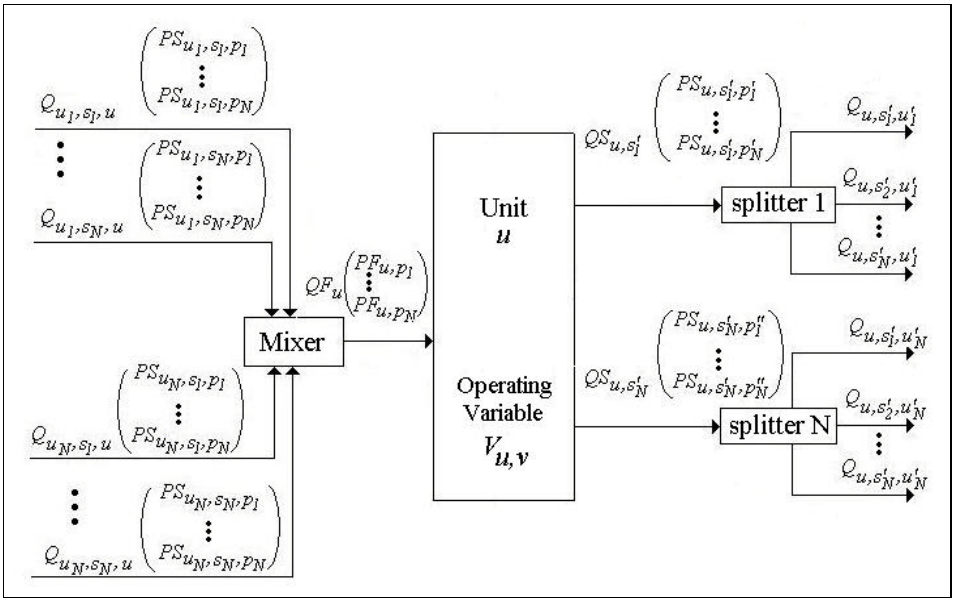

Figure 3. Unit Model Framework.

Indices:

c	scenario
p	property
s	stream
t	time period
u, u'	unit
v	operating variable

Sets:

$\mathbf{C}$	scenarios $\{ c \mid c$ the $1,\ldots,NC \}$
$\mathbf{PI}_u$	properties of the inlet stream of unit u
$\mathbf{PO}_{u,s}$	properties of outlet stream s of unit u
$\mathbf{SO}_u$	outlet streams of unit u
$\mathbf{T}$	time periods $\{ t \mid t$ the $1,\ldots,NT \}$
$\mathbf{U}$	units of the refinery complex
$\mathbf{U}_f$	petroleum tanks
$\mathbf{U}_{feed}$	units that process petroleum
$\mathbf{UI}_u$	units whose outlet streams feed unit u
$\mathbf{UO}_{u,s}$	units that are fed by stream s of unit u
$\mathbf{U}_p$	product tanks
$\mathbf{US}_u$	ordered pairs (u',s) that feeds unit u
$\mathbf{VO}_u$	operating variables of unit u

Parameters:

Cb_u	pumping cost for unit u
$Cf_{u,t,c}$	price of petroleum u ($u \in \mathbf{U}_f$) in time period t under the scenario c
$Cinv_{u,t}$	inventory cost of product u ($u \in \mathbf{U}_p$) in time period t
$Cp_{u,t}$	price of product u ($u \in \mathbf{U}_p$) in time period t
Cr_u	fixed operating cost coefficient of unit u
$Cv_{u,v}$	variable coefficient cost of the operating variable v of unit u
$Dem_{u,t}$	demand of prod. u in time period t ($u \in \mathbf{U}_p$)
$PF^L_{u,p,t}$	lower bound of inlet property p of unit u in time period t
$PF^U_{u,p,t}$	upper bound of inlet property p of unit u in time period t
$prob_{t,c}$	probability of scenario c in time period t
$Prop_{u,s,p}$	standard property value p of the outlet stream s from unit u
QF^L_u	lower bound for feed flow rate of unit u
QF^U_u	upper bound for feed flow rate of unit u
$QGain_{u,s,v}$	flow rate gain of outlet stream s of unit u due to operating variable u
QS^L_u	lower bound for outlet flow rate ofunit u
QS^U_u	upper bound for outlet flow rate of unit u
$V^L_{u,v}$	lower bound for operating variable v of unit u
$V^U_{u,v}$	upper bound for operating variable v of unit u

Variables:

$Est_{u,t}$	inventory of product u in time t ($u \in \mathbf{U}_p$)
$I_{u,s,p,t}$	mixture indices of property p of stream s of unit u in time period t
$PF_{u,p,t}$	property p of the feed stream at unit u in time period t
$PS_{u,s,p,t}$	property p of the outlet stream s at unit u in time period t
$QF_{u,t}$	feed flow rate of unit u in time t
$QS_{u,s,t}$	outlet flow rate of stream s at unit u in time period t
$Q_{u,s,u',c,t}$	flow rate of stream s between units u' and u in time period t for scenario c
$Q_{u',s,u,t}$	flow rate of stream s between units u' and u in time period t
$V_{u,v,t}$	operating variable v of unit u in time t
$y_{u,t}$	binary variable which assumes 1 if petroleum type u is chosen to compose refinery feed ($u \in \mathbf{U}_f$); 0 otherwise.

As a basis for the multiperiod network model, **MR** is defined for each refinery.

$$\begin{aligned}\text{Max } Z = & \sum_{u \in \mathbf{U}_p} \sum_{t \in \mathbf{T}} Cp_{u,t} QF_{u,t} - \\ & \sum_{u \in \mathbf{U}_f} \sum_{s \in \mathbf{SO}_{\mathbf{U}_f}} \sum_{t \in \mathbf{T}} \sum_{c \in \mathbf{C}} prob_{t,c} Cf_{u,t,c} Q_{u,s,u',t,c} - \\ & \sum_{u \in \mathbf{U}} \sum_{t \in \mathbf{T}} [Cr_u + \sum_{v \in \mathbf{VO}_u} (Cv_{u,v} . V_{u,v,t})] \, QF_{u,t} - \\ & \sum_{u \in \mathbf{U}_p} \sum_{t \in \mathbf{T}} Cinv_{u,t} Est_{u,t} - \sum_{u \in \mathbf{U}_f} \sum_{t \in \mathbf{T}} Cb_u y_{u,t}\end{aligned} \tag{1}$$

subject to:

$$QF_{u,t} = \sum_{(u',s) \in \mathbf{US}_u} Q_{u',s,u,t} \qquad \forall\, u \in \mathbf{U} \setminus \{ \mathbf{U}_f, \mathbf{U}_{feed} \}, t \in \mathbf{T} \tag{2}$$

$$QF_{u,t} = \sum_{(u',s) \in \mathbf{US}_u} \sum_{c \in \mathbf{C}} Q_{u',s,u,c,t} \qquad \forall\, u \in \mathbf{U}_{feed}, t \in \mathbf{T} \tag{3}$$

$$\begin{aligned} QS_{u,s,t} = QF_{u,t} . f_{u,s}\left(PF_{u,p,t} \right) + \sum_{u \in \mathbf{VO}_u} QGain_{u,s,v} . V_{u,v,t} \\ \forall\, u \in \mathbf{U} \setminus \{ \mathbf{U}_p, \mathbf{U}_f \}, s \in \mathbf{SO}_u, p \in \mathbf{PI}_u, t \in \mathbf{T}\end{aligned} \tag{4}$$

$$QS_{u,s,t} = \sum_{u' \in \mathbf{UO}_{u,s}} Q_{u,s,u',t} \quad \forall\, u \in \mathbf{U} \setminus \{ \mathbf{U}_p, \mathbf{U}_f \},\ s \in \mathbf{SO}_u, t \in \mathbf{T} \tag{5}$$

$$QS_{u,s,t} = \sum_{u' \in \mathbf{UO}_{u,s}} \sum_{c \in \mathbf{C}} Q_{u,s,u',c,t} \qquad \forall u \in \mathbf{U}_f, s \in \mathbf{SO}_u, t \in \mathbf{T} \tag{6}$$

$$PF_{u,p,t} = \frac{\sum_{(u',s) \in \mathbf{US}_u} Q_{u',s,u,t} . PS_{u',s,p,t}}{\sum_{(u',s) \in \mathbf{US}_u} Q_{u',s,u,t}} \quad \forall u \in \mathbf{U} \setminus \mathbf{U}_f, p \in \mathbf{PI}_u, t \in \mathbf{T} \tag{7}$$

$$\begin{aligned} PS_{u,s,p,t} = f_{u,s,p}(PF_{u,p,t} \mid p \in \mathbf{PI}_u, V_{u,v,t} \mid v \in \mathbf{VO}_u) \\ \forall u \in \mathbf{U} \setminus \mathbf{U}_p, s \in \mathbf{SO}_u, p \in \mathbf{PO}_{u,s}, t \in \mathbf{T}\end{aligned} \tag{8}$$

$$Est_{u,t} = Est_{u,t-1} + QF_{u,t} - Dem_{u,t} \qquad \forall u \in \mathbf{U}_p, t \in \mathbf{T} \tag{9}$$

$$QF_u^L \le QF_{u,t} \le QF_u^U \qquad \forall u \in \mathbf{U} \setminus \mathbf{U}_f, t \in \mathbf{T} \tag{10}$$

$$y_{u,t} . QS_u^L \le QS_{u,s,t} \le QS_u^U . y_{u,t} \qquad \forall u \in \mathbf{U}_f, s \in SO_u, t \in \mathbf{T} \tag{11}$$

$$PF_{u,p,t}^L \le PF_{u,p,t} \le PF_{u,p,t}^U \qquad \forall u \in \mathbf{U} \setminus \mathbf{U}_f, p \in \mathbf{PI}_u, t \in \mathbf{T} \tag{12}$$

$$V_{u,v}^L \le V_{u,v,t} \le V_{u,v}^U \qquad \forall u \in \mathbf{U} \setminus \{ \mathbf{U}_f, \mathbf{U}_p \}, v \in \mathbf{VO}_u, t \in \mathbf{T} \tag{13}$$

$$QF, QS, Q, Est \in \Re^+ \quad PF, PS, V \in \Re \quad y \in \{0,1\} \tag{14}$$

The objective function (1) is defined as the maximization of the revenue obtained by the product sales minus costs related to raw material, operation, inventory and pumping. Raw material is purchased considering different discrete scenarios of the petroleum market. Each scenario is weighted by its discrete probability of occurrence, $prob_{c,t}$. The operating cost is a non-linear term that depends on the unit operating mode. If the unit is operated at its design condition, a fixed cost is incurred. Otherwise, a proportional cost is incurred, which depends on the deviation variable. A constant transportation cost (Cb_u) that depends on crude selection is incurred for crude oil transfer through pipelines. Equation (2) describes mass balances at the mixers for general units whereas Eq. (3) describes mass balances at the mixers for units that process crude oil. These units can receive petroleum according to the different scenarios. Equation (4) denotes the relation of the product flow rates with the feed flow rate ($QF_{u,t}$), feed properties ($f_{u,s}$ is a linear function of $PF_{u,p,t}$ and depends on the unit and outlet stream) and operating variables ($V_{u,v,t}$). Equation (4) is valid for units whose product yields closely depend on petroleum type, such as atmospheric and vacuum distillation columns. The other units operate at constant yields, which means that the function $f_{u,s}(PF_{u,p,t})$ is replaced by a constant parameter. Therefore, Eq. (4) becomes linear for these cases. Equations (5) and (6) represent mass balances at the splitters of general units and petroleum tanks, respectively. Equation (7) represents a weighted average that relates properties of the unit feed stream with properties of the inlet streams. There are some cases for which property must be replaced by mixture indices in order to apply Eq. (7) and some properties must be weighted on a mass basis. In the latter case, the density of the corresponding stream must multiply every term in the numerator and denominator of Eq. (7). Equation (8) shows the general relationship among outlet properties, feed properties and operating variables. The functional form of Eq. (8) depends on the unit, stream and property under consideration. In the present work, most of the outlet properties are considered to be measured values, and therefore only a few are estimated. Those are usually properties that depend on petroleum types, such as sulfur content. Equation (9) represents the production balance over each time period. Equations (10-13) are operating and quality constraints for decision variables. Equation (11) also forbids low petroleum requirements. Eq. (14) defines the domain of the decision variables.

The network model is, therefore, composed of Eq. (1) as objective function whose terms result of the sum of sub-terms corresponding to each refinery. The objective function is subject to Eqs. (2-14) applied to every refinery as constraints. Elements that compose the set of existing units in the network are defined so that they identify the type of unit and of the refinery site to which they belong. This approach prevents the addition of another subscript to identify refineries. This is also convenient because despite the fact that refineries may contain the same type of units, these usually present different designs. The proposed approach is detailed in the “Case Study” section.

Illustrative Example

Consider a simplified refinery example in Figure 4 to illustrate the application of model **MR**. The refinery is composed of an atmospheric distillation column (CD1), a vacuum distillation unit (VD1), a fluid catalytic cracking unit (FCC), a propane deasphalting unit (PDA) and a hydrotreating unit (HT3). Atmospheric distillation fractionates crude oil into the following hydrocarbon streams: compounds with 3 and 4 carbon atoms (C3C4), light naphtha (LN), heavy naphtha (HN), kerosene (K), light diesel (LD), heavy diesel (HD) and residue (ATR). The vacuum distillation column fractionates the ATR residue from CD1 in two streams: vacuum gasoil (VGO) and vacuum residue (VR). The FCC unit produces light cycle oil (LCO), decanted oil (DO), cracked naphtha (CRAN) and a light hydrocarbon mixture (C3C4). PDA produces deasphalted oil (DAO) and asphaltic residue

(ASFR), and HT3 improves product quality by reducing sulfur content (HTD). Products are identified by their pool names: liquefied petroleum gas (PGLP), interior diesel (PDIN), gasoline (PGLN), petrochemical naphtha (PPQN) and fuel oil (PFO1A). Three crude oil types are available for feeding the refinery: BONITO, MARLIN and RGN. Figure 4 shows the topology of the refinery based on that of Pinto and Moro (2000).

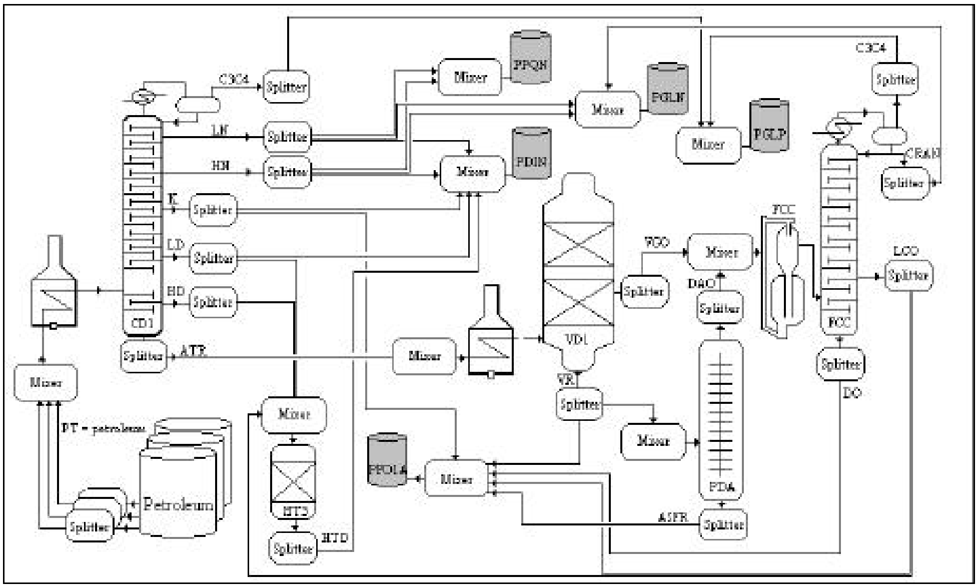

Figure 4. A Simplified Refinery.

The planning horizon spans two time periods and there are two possible scenarios for petroleum market prices. Therefore, the refinery multiperiod planning model can be described as follows[1]:

Petroleum tank model

Equations (6) and (11) model the outlet streams of petroleum tanks[2]. Since these are simply mass balances and bound constraints, it is only necessary to define the valid sets that are as follows: $\mathbf{U}_f$ = {BONITO, MARLIN, RGN} and $\mathbf{T}$ = {1,2}.

$$QS_{u,PT,t} = \sum_{c=1}^{2} Q_{u,PT,CD1,c,t} \qquad \forall u \in \mathbf{U}_f, t \in \mathbf{T} \quad (15)$$

$$y_{u,t}.500 \le QS_{u,PT,t} \le 15000.y_{u,t} \qquad \forall u \in \mathbf{U}_f, t \in \mathbf{T} \quad (16)$$

CD1 Model

Equations (3-5), (7-8), (10) and (13) model the atmospheric distillation column. *CD1* feed is composed of a petroleum mixture that results from all petroleum types available in market to be purchased by the refinery ($\mathbf{UI}_{CD1} = \mathbf{U}_f$) as stated in Eq. (17):

$$QF_{CD1,t} = \sum_{u' \in \mathbf{UI}_{CD1}} \sum_{c=1}^{2} Q_{u',PT,CD1,c,t} \qquad \forall\ t \in \mathbf{T} \quad (17)$$

Moreover, feed flow rate must satisfy *CD1* operating capacity:

$$14000 \le QF_{CD1,t} \le 36000 \qquad \forall t \in \mathbf{T} \quad (18)$$

Production level depends on the feed flow rate, feed properties and on a single operating variable:

$$QS_{CD1,s,t} = QF_{CD1,t}.PF_{CD1,p,t} + QGain_{CD1,s}.V_{CD1,V1,t} \qquad \forall\ s \in \mathbf{SO}_{CD1}, p \in \mathbf{PI}_{CD1}, t \in \mathbf{T} \quad (19)$$

where $\mathbf{SO}_{CD1}$ = {*C3C4, LN, HN, K, LD, HD, ATR*} and $\mathbf{PI}_{CD1}$ = {Y_{C3C4}, Y_{LN}, Y_{HN}, Y_K, Y_{LD}, Y_{HD}, Y_{ATR}}. The elements of the set $\mathbf{PI}_{CD1}$ denote yields of the outlet streams and depend on the petroleum type. The operating variable $V_{CD1,V1,t}$ is the feed temperature deviation (*V1*). Distillation column is fed at the design temperature value when $V_{CD1,V1,t} = 0$ and it yields the distance from the design temperature when $V_{CD1,V1,t} \neq 0$. Temperature deviation of the column feed must also satisfy the following design constraint:

$$-10 \le V_{CD1,V1,t} \le 10 \qquad \forall t \in \mathbf{T} \quad (20)$$

Feed properties that appear in Eq. (19) are weighted according to each petroleum type selected to compose the refinery feed:

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[1] Mass balances at splitters are not shown. Figure 4 clearly shows connections among units.

[2] Outlet streams from petroleum tanks are referred to as *PT* to denote *Petroleum.*

$$PF_{CD1,p,t} = \frac{\sum_{u' \in \mathbf{UI}_{CD1}} Q_{u',PT,CD1,t} . Prop_{u',PT,p}}{\sum_{u' \in \mathbf{UI}_{CD1}} Q_{u',PT,CD1,t}} \quad \forall p \in \mathbf{PI}_{CD1}, t \in \mathbf{T} \quad (21)$$

where $Prop_{u',PT,p}$ is a parameter that denotes the property p assigned by the petroleum type u'. Properties of the outlet streams can be modified by the operating variable as in Eq. (22):

$$PS_{CD1,s,p,t} = Prop_{CD1,s,p} + PGain_{CD1,s,p} . V_{CD1,V1,t} \quad \forall, s \in \mathbf{SO}_{CD1}, p \in \mathbf{PO}_{CD1,s}, t \in \mathbf{T} \quad (22)$$

Analogously, $Prop_{CD1,s,p}$ is a parameter that denotes the property p of the product stream s, and $\mathbf{SO}_{CD1}$ and $\mathbf{PO}_{CD1,s}$ are defined according to the refinery topology and product stream, respectively. The elements of these sets are not shown for the sake of simplicity, since every stream $s \in \mathbf{SO}_{CD1}$ defines a set $\mathbf{PO}_{CD1,s}$.

Petroleum types characterize both production yields for every product stream of *CD1* and the sulfur content carried by each of these product streams. Consequently, sulfur amount strongly depends on the petroleum types fed into *CD1* and must be estimated through a relation that is similar to Eq. (21):

$$PF_{CD1,S,t} = \frac{\sum_{u' \in \mathbf{UI}_{CD1}} Q_{u',PT,CD1,t} . Prop_{u',PT,S}}{\sum_{u' \in \mathbf{UI}_{CD1}} Q_{u',PT,CD1,t}} \quad \forall t \in \mathbf{T} \quad (23)$$

where S denotes sulfur and $Prop_{u',PT,S}$ is the sulfur content present in the petroleum type u'.

VD1 Model

Equations (2), (4-5), (7-8) and (10) model the vacuum distillation column. Since *VD1* is fed only with atmospheric residue from *CD1*, inlet variables are equal to outlet variables of *ATR* stream:

$$\begin{aligned} QF_{VD1,t} &= Q_{CD1,ATR,VD1,t} && \forall t \in \mathbf{T} \\ PF_{VD1,p,t} &= PS_{CD1,ATR,p,t} && p \in \mathbf{PI}_{VD1}, t \in \mathbf{T} \end{aligned} \quad (24)$$

Production yields of the *VD1* outlet streams depend on the petroleum types supplied as well as on the sulfur content of the inlet stream. Therefore, the calculation procedure is identical to that of unit *CD1*. Since there is no relevant operating variable for *VD1*, Eqs. (4) and (8) simplify respectively to Eqs. (25) and (26).

$$\begin{aligned} QS_{VD1,VGO,t} &= QF_{VD1,t} . PF_{VD1,Y_{VGO},t} && \forall t \in \mathbf{T} \\ QS_{VD1,VR,t} &= QF_{VD1,t} . PF_{VD1,Y_{VR},t} && \forall t \in \mathbf{T} \end{aligned} \quad (25)$$

$$\begin{aligned} PS_{VD1,VGO,p,t} &= Prop_{VD1,VGO,p} && \forall p \in \mathbf{PO}_{VD1,VGO}, t \in \mathbf{T} \\ PS_{VD1,VR,p,t} &= Prop_{VD1,VR,p} && \forall p \in \mathbf{PO}_{VD1,VR}, t \in \mathbf{T} \end{aligned} \quad (26)$$

where $Prop_{VD1,VGO,p}$ and $Prop_{VD1,VR,p}$ are property values of the outlet streams *VGO* and *VR*, respectively. Unit *VD1* operates within the following range:

$$10000 \le QF_{VD1,t} \le 24000 \quad \forall t \in \mathbf{T} \quad (27)$$

PDA Model

The set of equations used for *VD1* also applies to *PDA*. Since *PDA* is exclusively fed by *VR* from *VD1* and no operating variable is considered, the following equations hold:

$$\begin{aligned} QF_{PDA,t} &= Q_{VD1,VR,PDA,t} && \forall t \in \mathbf{T} \\ 4000 \le QF_{PDA,t} &\le 7200 && \forall t \in \mathbf{T} \end{aligned} \quad (28)$$

$$PF_{PDA,p,t} = PS_{PDA,VR,p,t} \quad p \in \mathbf{PI}_{PDA}, t \in \mathbf{T}$$

$$\begin{aligned} QS_{PDA,DAO,t} &= Yield_{PDA,DAO} . QF_{PDA,t} && \forall t \in \mathbf{T} \\ QS_{PDA,ASFR,t} &= Yield_{PDA,ASFR} . QF_{PDA,t} && \forall t \in \mathbf{T} \end{aligned} \quad (29)$$

$$\begin{aligned} PS_{PDA,DAO,p,t} &= Prop_{PDA,DAO,p} && \forall p \in \mathbf{PO}_{PDA,DAO}, t \in \mathbf{T} \\ PS_{PDA,ASFR,p,t} &= Prop_{PDA,ASFR,p} && \forall p \in \mathbf{PO}_{PDA,ASFR}, t \in \mathbf{T} \end{aligned} \quad (30)$$

Product flow rates are calculated from constant yield values as shown in Eq. (29). Note that $Yield_{PDA,DAO}$ and $Yield_{PDA,ASFR}$ denote fixed parameters differently from *PF* used for *CD1* and *VD1* units. Equation (30) holds in the case of properties that do not depend on the properties of the inlet stream. Equation (31) evaluates sulfur content of the product streams of *PDA*, which depends on the inlet conditions.

$$\begin{aligned} PS_{PDA,DAO,S,t} &= sulfur_{PDA,DAO} . PF_{PDA,S,t} && \forall t \in \mathbf{T} \\ PS_{PDA,ASFR,S,t} &= sulfur_{PDA,ASFR} . PF_{PDA,S,t} && \forall t \in \mathbf{T} \end{aligned} \quad (31)$$

where $sulfur_{PDA,DAO}$ and $sulfur_{PDA,ASFR}$ are constant parameters.

FCC Model

Equations (2), (4-5), (7-8) and (10) model the fluid catalytic cracking unit. Its feed is composed by the combination of *DAO* from *PDA* and *VGO* from *VD1* so that feed flow rate is determined by Eq. (32) and its operating range is expressed by Eq. (33).

$$QF_{VD1,t} = Q_{VD1,VGO,FCC,t} + Q_{PDA,DAO,FCC,t} \quad \forall t \in \mathbf{T} \quad (32)$$

$$7000 \le QF_{FCC,t} \le 12500 \quad \forall t \in \mathbf{T} \quad (33)$$

Properties of the inlet stream of *FCC* are calculated through the weighted average of properties of the two streams that compose the *FCC* feed:

$$PF_{FCC,p,t} = \frac{\sum_{(u',s) \in \mathbf{UI}_{FCC}} Q_{u',s,FCC,t} . PS_{u',s,D20,t} . PS_{u',s,p,t}}{\sum_{(u',s) \in \mathbf{UI}_{FCC}} Q_{u',s,FCC,t} . PS_{u',s,D20,t}} \quad \forall p \in \mathbf{PI}_{FCC}, t \in \mathbf{T} \quad (34)$$

where $\mathbf{UI}_{FCC}$ = {(*VD1*, *VGO*), (*PDA*, *DAO*)} and density $PS_{u',s,D20,t}$ is used to estimate properties in a mass basis. Product flow rates are not influenced by any operating variable, but depend on carbon residue (*RCR*) fed to *FCC* resulting in a special form of Eq (04) :

$$QS_{FCC,s,t} = QF_{FCC,t}.[Yield_{FCC,s} + YGain_{FCC,s}\left(PF_{FCC,RCR,t} - RC_{FCC}\right)] \quad \forall s \in \mathbf{SO}_{FCC}, t \in \mathbf{T} \quad (35)$$

In Eq. (35), $YGain_{FCC,s}$ is a flow rate gain parameter related to the carbon residue property ($PF_{FCC,RCR,t}$), RC_{FCC} is a constant parameter and $\mathbf{SO}_{FCC}$ = {*C3C4*, *CRAN*, *LCO*, *ATR*}. The parameter $YGain_{FCC,s}$ may assume either positive or negative values. Properties of the outlet streams are standard values (Eq. 36), with exception of sulfur content that is defined according to sulfur content at the inlet of *FCC* (Eq. 37).

$$PS_{FCC,s,p,t} = Prop_{FCC,s,p} \quad \forall p \in \mathbf{PO}_{FCC,s}, t \in \mathbf{T} \quad (36)$$

$$PS_{FCC,s,S,t} = sulfur_{FCC,s}.PF_{FCC,S,t} \quad \forall t \in \mathbf{T} \quad (37)$$

HT3 Model

Equations (2), (4-5), (7-8), (10) and (13) model the hydrotreating unit. *HT3* is fed by three streams (*LD*, *HD*, *LCO*) that leave two units (*CD1*, *FCC*), so that feed flow rate is given by Eq. (37) whereas the operating range is limited by Eq. (39).

$$QF_{HT3,t} = Q_{CD1,LD,HT3,t} + Q_{CD1,HD,HT3,t} + Q_{FCC,LCO,HT3,t} \quad \forall t \in \mathbf{T} \quad (38)$$

$$3200 \le QF_{HT3,t} \le 7500 \quad \forall t \in \mathbf{T} \quad (39)$$

Some properties at the inlet of *HT3* must be converted to index form in order to be additive. Three properties are subject to such procedure: viscosity (*VISCO*), flash point (*FP*) and DASTM 85% (*A85*), which limits the content of heavy fractions that are related to large carbon residue and poor color. Their mixture indices are calculated from Eqs. (40-42).

$$I_{u',s,VISCO,t} = \frac{\log_{10} P_{u',s,VISCO,t}}{\log_{10} 1000 P_{u',s,VISCO,t}} \quad (40)$$
$$\forall (u',s) \in \mathbf{US}_{HT3}, t \in \mathbf{T}$$

$$I_{u',s,FP,t} = \exp\left[\frac{10006.1}{1.8P_{u',s,FP,t} + 415} - 14.0922\right] \quad (41)$$
$$\forall (u',s) \in \mathbf{US}_{HT3}, t \in \mathbf{T}$$

$$I_{u',s,A85,t} = \left(\frac{1.8P_{u',s,A85,t} + 32}{549}\right)^{7.8} \quad (42)$$
$$\forall (u',s) \in \mathbf{US}_{HT3}, t \in \mathbf{T}$$

Once these mixture indices have been determined, properties of the inlet stream of *HT3* can be evaluated through Eq. (7). The exception is property *A85* that must be calculated by Eq. (43).

$$PF_{FCC,A85,t} = 305\left(\frac{\sum_{(u',s)\in \mathbf{UI}_{FCC}} Q_{u',s,FCC,t}.I_{u',s,A85,t}}{\sum_{(u',s)\in \mathbf{UI}_{FCC}} Q_{u',s,FCC,t}}\right)^{0.128} \text{-17.78} \quad (43)$$
$$\forall p \in \mathbf{PI}_{FCC}, t \in \mathbf{T}$$

Outlet flow rate equals inlet flow rate as well as some properties at the outlet stream. On the other hand, sulfur content (*S*) and cetane number (*CN*) depend on the operating variable and are calculated through Eqs. (44) and (45), where $VR_{HT3,S}$ and $VR_{HT3,CN}$ are constant parameters. Operating variable range must assume values within the interval defined through Eq. (46).

$$PS_{HT3,HTD,S,t} = PF_{HT3,S,t}.\left(VR_{HT3,S} \text{-} V_{HT3,V1,t}\right) \quad \forall t \in \mathbf{T} \quad (44)$$

$$PS_{HT3,HTD,CN,t} = PF_{HT3,CN,t} - VR_{HT3,CN}.V_{HT3,V1,t} \quad \forall t \in \mathbf{T} \quad (45)$$

$$50 \le V_{HT3,V1,t} \le 90 \quad \forall t \in \mathbf{T} \quad (46)$$

Product Tank Model

In tanks dedicated to product storage, only the inlet stream is modeled (opposite to petroleum tanks). Equations (2), (7), (9) and (12) model product tanks and are analogous to previous equations. Equation (12) plays an important role since it defines product qualities.

The refinery model corresponds to a mixed-integer nonlinear programming (MINLP) problem, which contains 383 variables and 349 equations. Six binary variables were necessary in the decision making process of purchasing crude oil (three for each period). The model was implemented in the modeling system GAMS (Brooke *et al.*, 1998) and solved with DICOPT (Visvanathan and Grossmann, 1990). The NLP subproblems were solved with CONOPT2 (Drud, 1994), whereas the MILP master problems were solved with OSL (IBM, 1991). Overall, 1.98 CPU seconds were consumed to iterate 906 times between NLP and MIP sub-problems. NLP sub-problems represent nearly 75% of total solution time. Table 1 presents results of the amount purchased of each petroleum type considering the two scenarios. The first scenario was 40% likely to occur in the first period and 35% in the second one. Table 2 shows production and inventory levels for each period and Table 3 presents the optimal product properties calculated and their specifications.

Table 1. Volume of Petroleum Purchased (m^3).

Petroleum	First period		Second period	
	scenario		scenario	
	1	2	1	2
BONITO	0	0	15000	0
MARLIN	0	9649	0	0
RGN	0	15000	13118	1241

Table 2. Production and Inventory Levels (m^3).

Product	Production (*QF*)		Inventory (*Est*)	
Tank	Time period		Time period	
	1	2	1	2
PLPG	2689	3040	689	729
PPQN	200	250	0	0
PGLN	6152	7703	1152	4055
PDIN	11115	12768	615	1384
PFO1A	5232	6384	1632	3716

Table 3. Product Properties and Bounds.

Product tank	*Property*	Lower bound	Period 1	Period 2	Upper bound
PLPG	*MON*		83	83	
	PVR		4.96	4.94	15
PGLN	*MON*	81	82	81.4	
	PVR	0.3	0.55	0.57	0.7
PDIN	*FP*		0	2.4	
	A50	245	279	279	313
	A85	300	357	355	370
	S		0.14	0.11	0.5
	NC	40	43	43	
	D20	0.82	0.82	0.82	0.88
PFO1A	*S*		0.80	0.83	2.5
	VISCO		0.48	0.43	0.48

Equations 40-43 are critical constraints for the solution procedure of Model **MR** because of the high non-linearity and the wide domain of variables. This fact requires the problem to be carefully scaled and bounded. Another important aspect is that concerning starting point. Since Model **MR** leads to a non-convex NLP problem, different starting points can lead to different local optimums. In some cases search can be directed to infeasible solutions.

Case Study – A Refinery Network

Three of the refineries shown in Fig 2 are considered in the case study: REVAP, RPBC and RECAP. Connections between refineries and the oil terminal (GEBAST) are shown in Fig. 5. GEBAST is considered as a refinery that contains only petroleum tanks. It is assumed that refineries do not hold petroleum storage. Units *VD1* and *VD2* from RPBC can either send *VGO* to be processed at its own *FCC* unit or send it to be processed at the *FCC* from REVAP. Moreover, *CD1* from RECAP can either send *ATR* to be processed at *FCC* from its site or send it to *FCC* from REVAP. Another possibility is to use *DO* and *LCO* produced at RECAP to compose products at its site or send those streams to compose fuel oil products at REVAP.

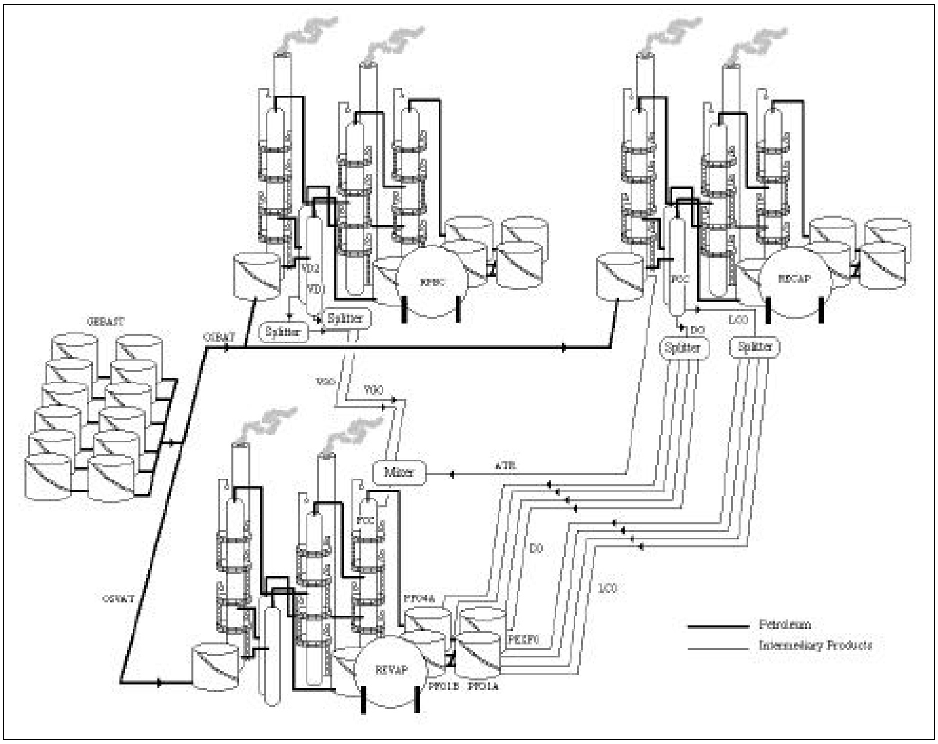

Figure 5. Case Study Complex.

Ten petroleum types are available to supply REVAP and RPBC. Since *CD1* from RECAP is operated differently, ten other petroleum types are dedicated to possibly supply RECAP. Besides, two scenarios represent market environment for the petroleum prices with the same probabilities as those of the illustrative example. Figures 6-8 show flowsheet of REVAP, RECAP and RPBC respectively detailing streams, units and their topologies. REVAP is composed of 8 units and has a processing capacity of 36,000 m^3/d of crude oil that is converted in 14 products RECAP is composed of 4 units and has a processing capacity of 8,500 m^3/d of crude oil that is converted in 10 products. RPBC is composed of 13 units and has a processing capacity of 27,000 m^3/d of crude oil that is converted in 15 products. Units will be correlated to the refineries to which they belong through the following notation: *RV* refers to REVAP, *RP* refers to RPBC and *RC* refers to RECAP.

The network optimization problem was built by applying model **MR** for every refinery and including connections shown in Figure5. The resulting model was solved using the modeling system GAMS (Brooke *et al.*, 1998) on a Pentium III / 700 MHz PC platform. DICOPT++ (Viswanathan and Grossmann, 1990) was chosen to solve the model since it corresponds to an (MINLP). NLP sub-problems were solved using CONOPT2 (Drud, 1994) whereas MIP master problems were solved with OSL (IBM, 1991).

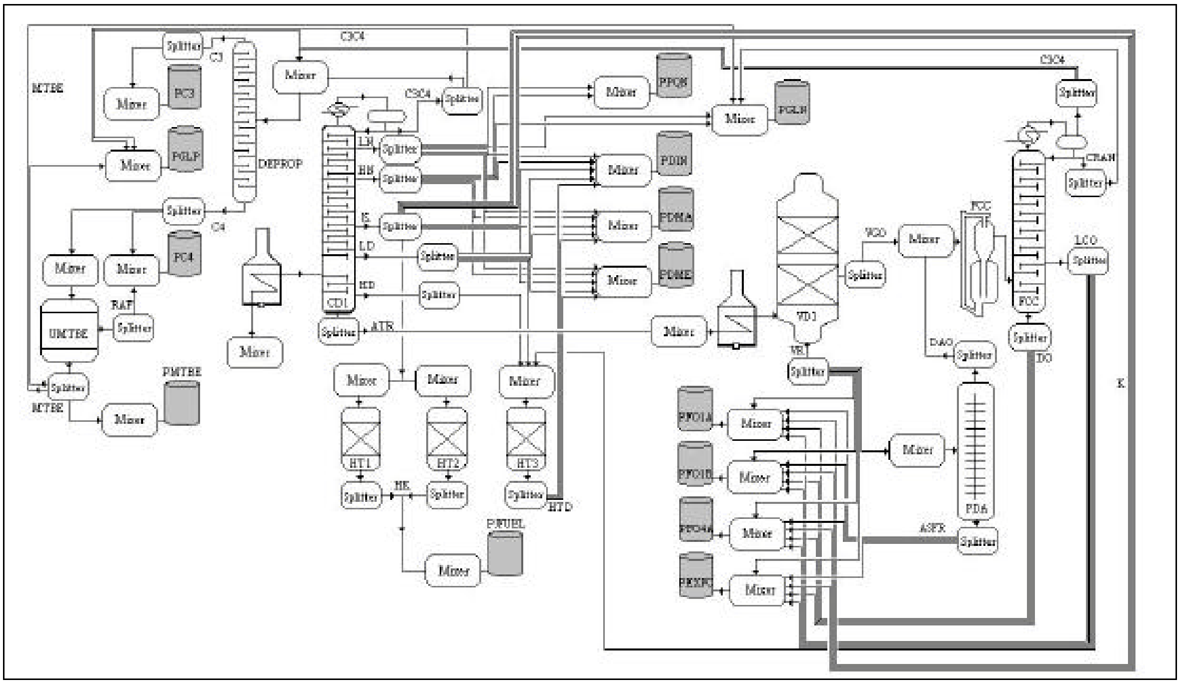

Figure 6. REVAP Flowsheet.

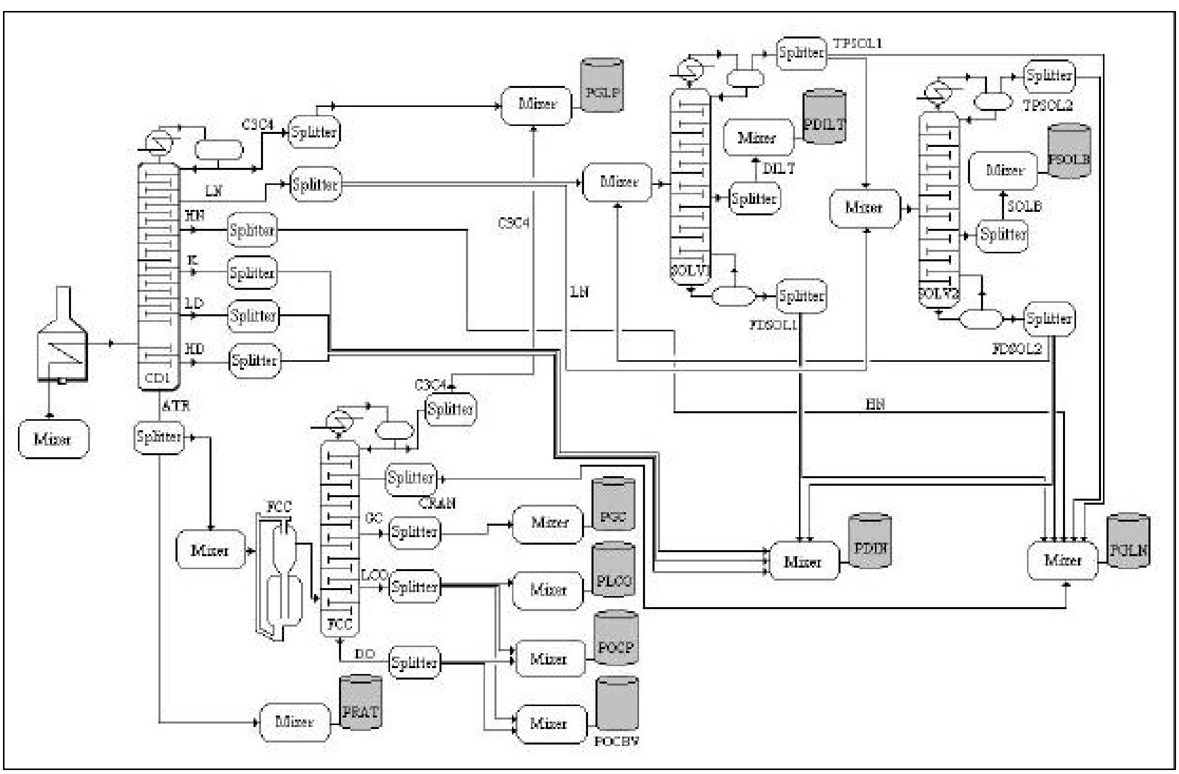

Figure 7. RECAP Flowsheet.

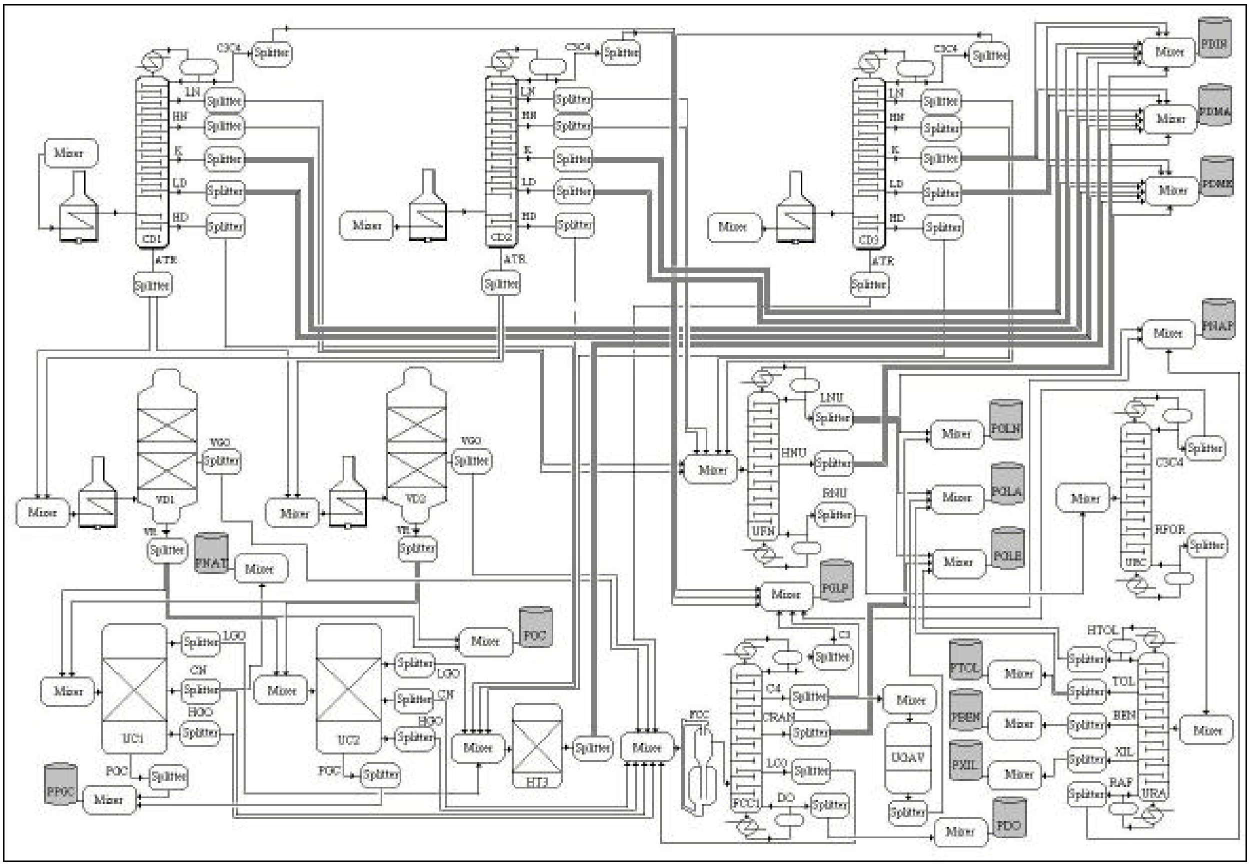

Figure 8. RPBC Flowsheet

Problems with up to 3 time periods were solved for the network complex and results were then compared to the production planning that considers refineries independently, namely Single-Site. Table 4 presents the resulting flowrates established among refineries. Tables 5, 6 and 7 present demand and product prices as input data and feed flowrate and inventory level as results obtained for REVAP, RPBC and RECAP, respectively. Feed flowrate is presented only for the Network case whereas inventory results are presented for both Single-Site and Network cases. Figure 9 shows the feed flowrates obtained for the Network case minus the feed flowrates obtained for the Single-Site case. Only units that present feed flowrate variation are presented in Figure 9.

As can be seen from Figure 9, *RV-CD1*, *RV-VD1* and *RV-PDA* undergo reduction in their feed flowrates as a result of the *VGO* and *ATR* amount sent respectively from *RP-VD1* and *RC-FCC* to *RV-FCC*. Moreover, the other connections shown in Table 4 cause a decrease to the *LCO* amount sent from *RV-FCC* to fuel oil tanks promoting an increase of the feed flowrate of *RV-HT3*. Figure 10 shows the amount of each type of petroleum purchased according to each case study whereas Table 8 reveals the gains obtained with introduction of the connections. Gains up to 540,000 $/d are possible.

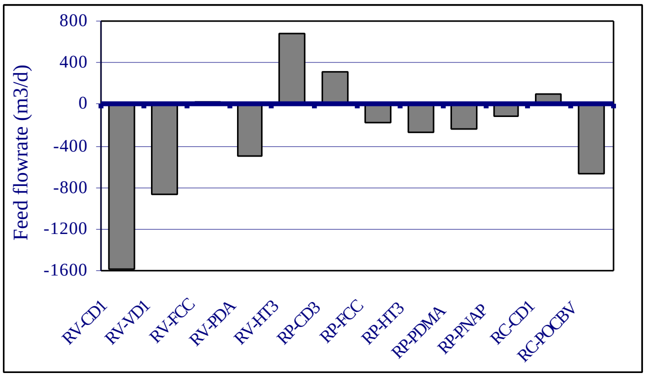

Figure 9. Feed flowrate comparison between Single-Site and Network cases.

Table 4. Connection flowrate among refineries

Origin Unit	Stream	Target Unit	Flowrate (m^3/d)
RP-VD1	VGO	RV-FCC	372
RP-FCC	LCO	RV-PFO1A	228
RP-FCC	LCO	RV-PFO1B	39
RC-CD1	ATR	RV-FCC	227
RC-FCC	LCO	RV-PFO1A	19
RC-FCC	DO	RV-PFO4A	701

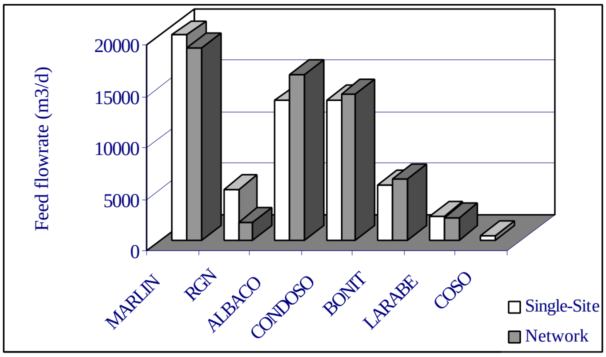

Figure 10. Petroleum purchase according to each case study.

Table 5. REVAP data and results

Product tanks	*Dem* (m^3)	*Cp* ($\$/m^3$)	*QF* (m^3/d)	*Est* (m^3) One-Site	*Est* (m^3) Network
PGLP	630	115	630	0	0
PJFUEL	0	130	0	0	0
PPQN	600	148	600	0	0
PGLN	9000	149	9000	0	0
PDIN	4000	210	4000	0	0
PDME	5000	230	5000	0	0
PDMA	1000	206	1000	0	0
PFO1A	800	139	800	0	0
PFO1B	50	142	50	0	0
PFO4A	4000	151	4000	0	0
PEXFO	300	146	300	0	0
PC3	1000	180	1000	0	0
PC4	100	100	292	192	192
PMTBE	100	100	100	0	0

Table 6. RPBC data and results

Product tanks	*Dem* (m^3)	*Cp* ($\$/m^3$)	*QF* (m^3/d)	*Est* (m^3) One-Site	*Est* (m^3) Network
PGLP	1100	118	1100	0	0
PDIN	4000	230	4000	0	0
PDME	3500	242	3500	0	0
PDMA	4500	213	6627	2371	2127
PNAP	900	146	1510	730	610
PNAT	158	160	158	0	0
PPGC	600	152	600	0	0
POC	1500	180	1500	0	0
PDO	700	159	700	0	0
PGLN	2100	270	2100	0	0
PGLA	1100	290	1100	0	0
PGLE	2100	298	2100	0	0
PXIL	100	160	115	15	15
PTOL	45	167	45	0	0
PBEN	210	280	210	0	0

Table 7. RECAP data and results

Product tanks	*Dem* (m^3)	*Cp* ($\$/m^3$)	*QF* (m^3/d)	*Est* (m^3) One-Site	*Est* (m^3) Network
PDIN	3600	203	3600	0	0
PRAT	200	66	200	0	0
PGC	500	100	500	0	0
POCP	90	144	90	0	0
PLCO	400	0	400	59	0
PGLP	650	141	650	0	0
PGLN	2300	232	2300	0	0
PSOLB	200	236	268	68	68
PDILT	90	257	112	22	22
POCBV	0	133	0	667	0

Table 8. Objective Function Values.

Time periods	One-Site	Network
1	$8.3\text{x}10^6$	$8.37\text{x}10^6$
2	$16.37\text{x}10^6$	$16.62\text{x}10^6$
3	$23.65\text{x}10^6$	$24.19\text{x}10^6$

An evident problem that arises when solving multiperiod production planning is the dramatic increase in computational effort with the number of time periods. Table 9 shows the increase in model size as a function of the number of periods.

Table 9. Computational Results –Network Case

Time periods	1	2	3
Constraints	1,023	2,045	3,067
Variables	1,176	2,351	3,526
Iterations	9567	57,533	2,966,973
CPU time (s)	3.23	603.12	59,513.90

Conclusions

The proposed approach clearly reveals the advantages of solving the entire refinery network complex in comparison to planning refineries independently. Nevertheless, it should be pointed out that significant effort should be directed for increasing the computational performance of simultaneous models such as the use of aggregation/ disaggregation and decomposition techniques. Moreover, pipeline transportation constraints both for supply and refinery interconnection should be considered. Coordination between production and transportation planning is the next step in modeling the petroleum supply chain.

Acknowledgments

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