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Multisite Refinery and Petrochemical Network Design: Optimal Integration and Coordination

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This paper addresses the design of optimal integration and coordination of a refinery and petrochemical network to satisfy given chemical products demand. The refinery and petrochemical systems were modeled as a mixed-integer problem with the objective of minimizing the annualized cost over a given time horizon among the refineries and maximizing the added value of the petrochemical network. The main feature of the paper is the development of a methodology for simultaneous analysis of process network integration within a multisite refinery and petrochemical system. This approach provides appropriate planning across the petroleum refining and petrochemical industry and achieves an optimal production strategy by allowing appropriate trade-offs between the refinery and the downstream petrochemical markets. The performance of the proposed model was tested on industrial-scale examples of multiple refineries and a poly(vinyl chloride) (PVC) complex to illustrate the economic potential and trade-offs involved in the optimization of the network. The proposed methodology not only devises the integration network in the refineries and synthesizes the petrochemical industry, but also provides refinery expansion requirements, production levels, and blending levels. The use of mathematical programming on an enterprise-wide scale to address strategic decisions considering various process integration alternatives yields substantial benefits. These benefits of process integration materialize in terms of economic considerations and improvements in the understanding of the process interactions and systems limitations.

1. Introduction

In view of the current situation of high oil prices and the increasing consciousness and implementation of strict environmental regulations, petroleum refiners and petrochemical companies have started seeking opportunities for mergers and integration. This is evidently seen from the current projects around the world for building integrated refineries and the development of complex petrochemical industries that are aligned through advanced integration platforms. The realization of coordination and objective alignment benefits across the range of enterprises has been the main driver of such efforts.[1,2]

Despite the fact that petroleum refining and petrochemical companies have recently engaged in more integration projects, relatively little research in the open literature has been reported, mostly due to confidentiality reasons. Such concerns render the development of a systematic framework of network integration and coordination difficult. The studies published in the open literature were mainly developed by consulting and design firms as well as the operating companies and generally lack a structured methodology for evaluating the feasibility of the projects. Just to mention a few of the studies published, Swaty[3] studied the possibility of integrating a refinery and an ethylene plant through the exchange of process intermediate streams. The analysis was based on a linear programming model for each plant and profit marginal analysis of possible intermediate plant exchange. The study was implemented on a real-life application in western Japan. Another study that highlighted the benefits of refining and petrochemical integration was based on a project of installing a hydrocracker in Repsol's Refinery in Spain.[4] The authors discussed the benefits of the overall synergy between the refinery and a stream cracker plant.

In the academic arena, Sadhukhan[5] developed an analytical flow sheet optimization method for applications in the petroleum refining and petrochemical industry. The proposed methodology consisted of three main steps: market integration, facility network optimization through economic margin analysis and load shifting, and elimination of less profitable processes. They demonstrated their method on two case studies: a single site refinery and a petrochemical complex. In another effort, a linear programming (LP) model for the integration of a refinery and an ethylene cracker was proposed.[6] The evaluation for different crude types to optimize the refinery and ethylene cracker operations was implemented iteratively. The best scheme was selected based on the highest profit from the cases studied. More recently, a short-term multiperiod planning model for a benzene−toluene−xylene (BTX) complex was studied.[7] The system was modeled as a mixed-integer linear programming (MILP) model with binary variables mainly for mode switching and inventory control (backlog and surplus indication). They divided the processing units into two sets: (1) reaction processes, including reforming, isomerization units, and tatory units; (2) separation processes, including aromatics extraction units, xylene fractionation units, and parex units. Optimization of the other refinery units, blending levels, and olefin cracking process were not considered. The decisions in their study mainly included the optimal throughput and operation mode of each production unit, inventory levels, and feedstock supplies. For more details on strategic multisite planning studies, multisite refinery optimization, and petrochemical industry planning the interested reader is referred to our recent work[8] and the references given herein.

Petroleum refining and petrochemical plant integration are gaining a great deal of interest as the realization of such coordination and vertical integration benefits grows. Previous research in the field either assumed no limitations on refinery feedstock availability for the petrochemical planning problem or fixed the refinery production levels assuming an optimal operation. However, in this paper we present a mathematical

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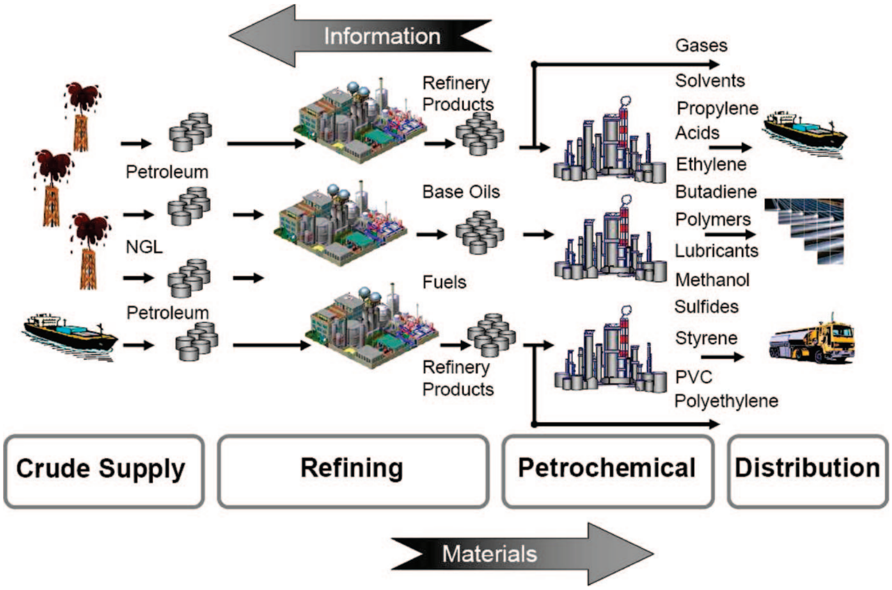

Figure 1. Refinery and petrochemical industry supply chain.

Table 1. Petrochemical Alternative Use of Refinery Streams[16]

refinery stream	petrochemical stream	alternative refinery use
FCC off-gas	ethylene	fuel gas
refinery propylene (FCC)	propylene	alkylation/polygasoline
reformate	benzene, toluene, xylenes	gasoline blending
naphtha and LPG	ethylene	gasoline blending
dilute ethylene (FCC and delayed coker off-gas)	ethylbenzene	fuel gas
refinery propylene (FCC product)	polypropylene, cumene, isopropyl alcohol, oligomers	alkylation
butylenes (FCC and delayed coker)	MEK (methyl ethyl ketone)	alkylation, MTBE
butylenes (FCC and delayed coker)	MTBE	alkylation, MTBE
refinery benzene and hydrogen	cyclohexane	gasoline blending
reformate	*o*-xylene	gasoline blending
reformate	*p*-xylene	gasoline blending
kerosene	*n*-paraffins	refinery product
FCC light cycle oil	naphthalene	diesel blending

model for the determination of the optimal integration and coordination strategy for a refinery network and synthesize the optimal petrochemical network required to satisfy a given demand from any set of available technologies. Therefore, we provide a seamless alignment across the petroleum refining and petrochemical industry supply chain (see Figure 1) and achieve a global optimal production strategy by allowing optimal trade-offs between the refining production system and the downstream petrochemical industry. The refinery and petrochemical systems were modeled as MILP problems that will also lead to an overall refinery and petrochemical process production levels and detailed blending levels at each refinery site. The objective function is a minimization of the annualized cost over a given time horizon among the refineries by improving the coordination and utilization of excess capacities in each facility and maximization of the added value of each product in the petrochemical system. Expansion requirements to improve production flexibility and reliability in the refineries are also considered.

The remainder of this paper is organized as follows. In section 2 we provide an overview of the essential petrochemical feedstocks that have direct interfaces with the petroleum refining industry. In section 3, we discuss some of the benefits and refinery and petrochemical potential synergy alternatives by examination of various aspects. We then explain the problem statement and proposed model formulation for the refinery and petrochemical integration in sections 4 and 5, respectively. In section 6, we illustrate the performance of the model through an industrial-scale case study. The paper ends with some concluding remarks in section 7.

2. Petrochemical Feedstocks

Petroleum feedstock, natural gas, and tar represent the main production chain drivers for the petrochemical industry.[9] From these basic components, many important petrochemical intermediates are produced including ethylene, propylene, butylenes, butadiene, benzene, toluene, and xylene. These essential intermediates are in turn converted into many other intermediates and final petrochemical products constructing a complex petrochemical network.

The preparation of intermediate petrochemical streams requires different processing alternatives depending on the feedstock quality. In our classification of petrochemical feedstocks we closely follow the one by Gary and Handwerk[10] consisting of aromatics, olefins, and paraffin/cycloparaffin compounds. The classification of petrochemical feedstocks into these clusters helps in identifying the different sources in the refinery that provide suitable feedstocks and therefore to better recognize areas of synergy between the refinery and petrochemical systems.

2.1. Aromatics. Aromatics are hydrocarbons containing a benzene ring, which is a stable and saturated compound. Aromatics used by the petrochemical industry are mainly benzene, toluene, and xylene (BTX) as well as ethylbenzene and are produced by catalytic reforming where their yield would

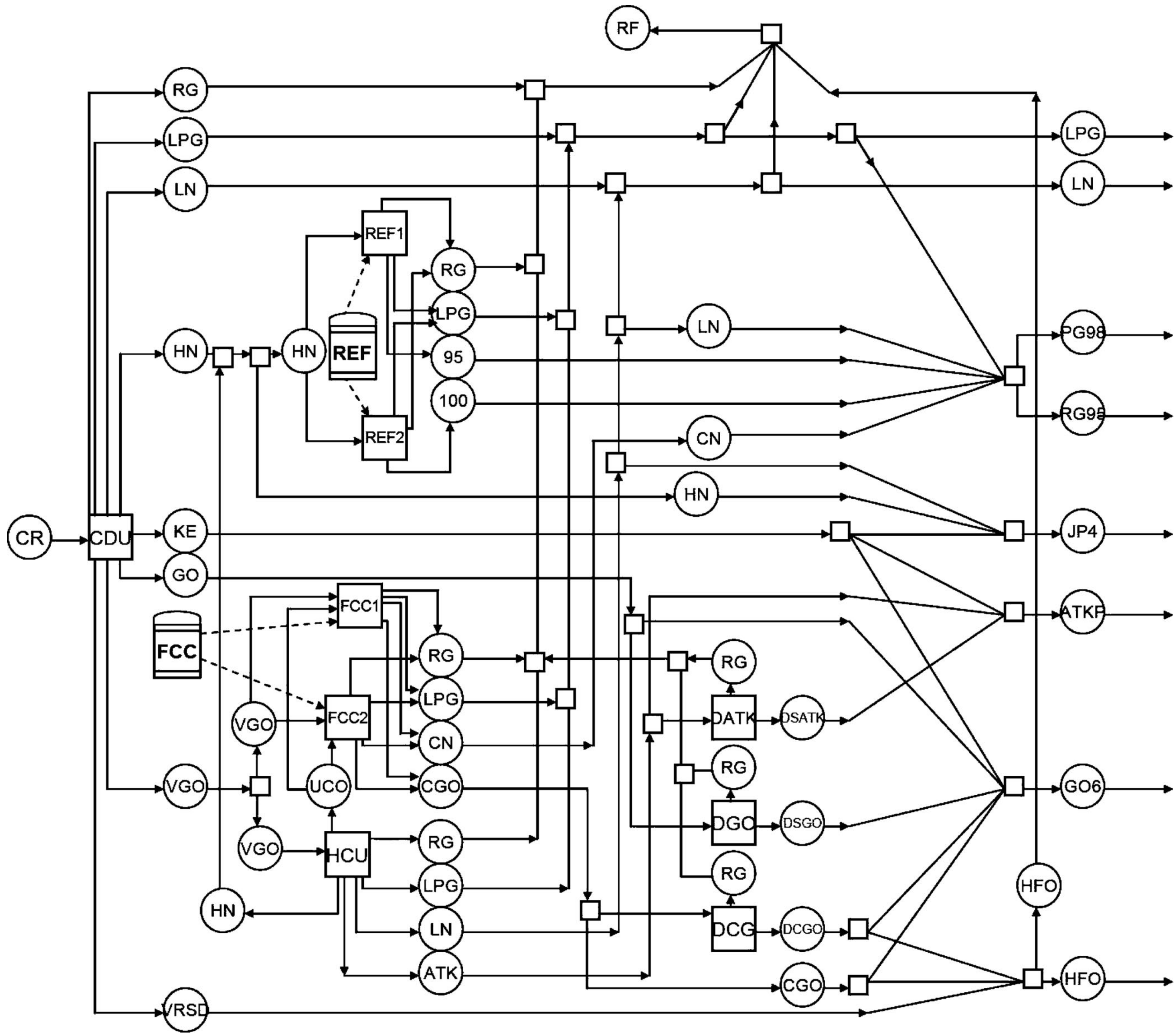

Figure 2. SEN representation of refinery 1.

increase with the increase of reforming severity.[10] Extractive distillation by different solvents, depending on the chosen technology, is used to recover such compounds. BTX recovery consists of an extraction using solvents that enhance the relative volatilities of the preferred compounds followed by a separation process based on the products' boiling points. Further processing of xylenes using isomerization/separation processes is commonly required to produce *o*-, *m*-, and *p*-xylene mixtures depending on market requirements. Benzene, in particular, is a source of a wide variety of chemical products. It is often converted to ethylbenzene, cumene, cyclohexane, and nitrobenzene, which in turn are further processed to other chemicals including styrene, phenol, and aniline.[11] Toluene production, on the other hand, is mainly driven by benzene and mixed xylenes demand. Mixed xylenes, particularly in Asia, are used for producing *p*-xylene and polyester.[12]

The other source of aromatics is pyrolysis gasoline (pygas), which is a byproduct of naphtha or gas oil steam cracking. This presents an excellent synergistic opportunity between refinery, BTX complex, and stream cracking for olefin production.

2.2. Olefins. Olefins are hydrocarbon compounds with at least two carbon atoms having a double bond where their unstable nature and tendency to polymerize makes them one of the very important building blocks for the chemical and petrochemical industry.[10] Although olefins are produced by fluid catalytic cracking in refineries, the main production source is through steam cracking of liquefied petroleum gas (LPG), naphtha, or gas oils.

The selection of steam cracker feedstock is mainly driven by market demand as different feedstock qualities produce different olefin yields. One of the commonly used feed quality assessment approaches in practice is The Bureau of Mines Correlation Index (BMCI).[4] This index is a function of the average boiling point and specific gravity of a particular feedstock. The steam cracker feed quality improves with a decrease in the BMCI value. For instance, vacuum gas oil (VGO) has a high value of BMCI and therefore is not a good steam cracker feedstock. The commonly used feedstocks in industry are naphtha and gas oil.

Steam cracking plays an instrumental role in the petrochemical industry in terms of providing the main petrochemical intermediates for the downstream industry. Steam cracker olefin production includes ethylene, propylene, butylene, and benzene. These intermediates are further processed into a wide range of

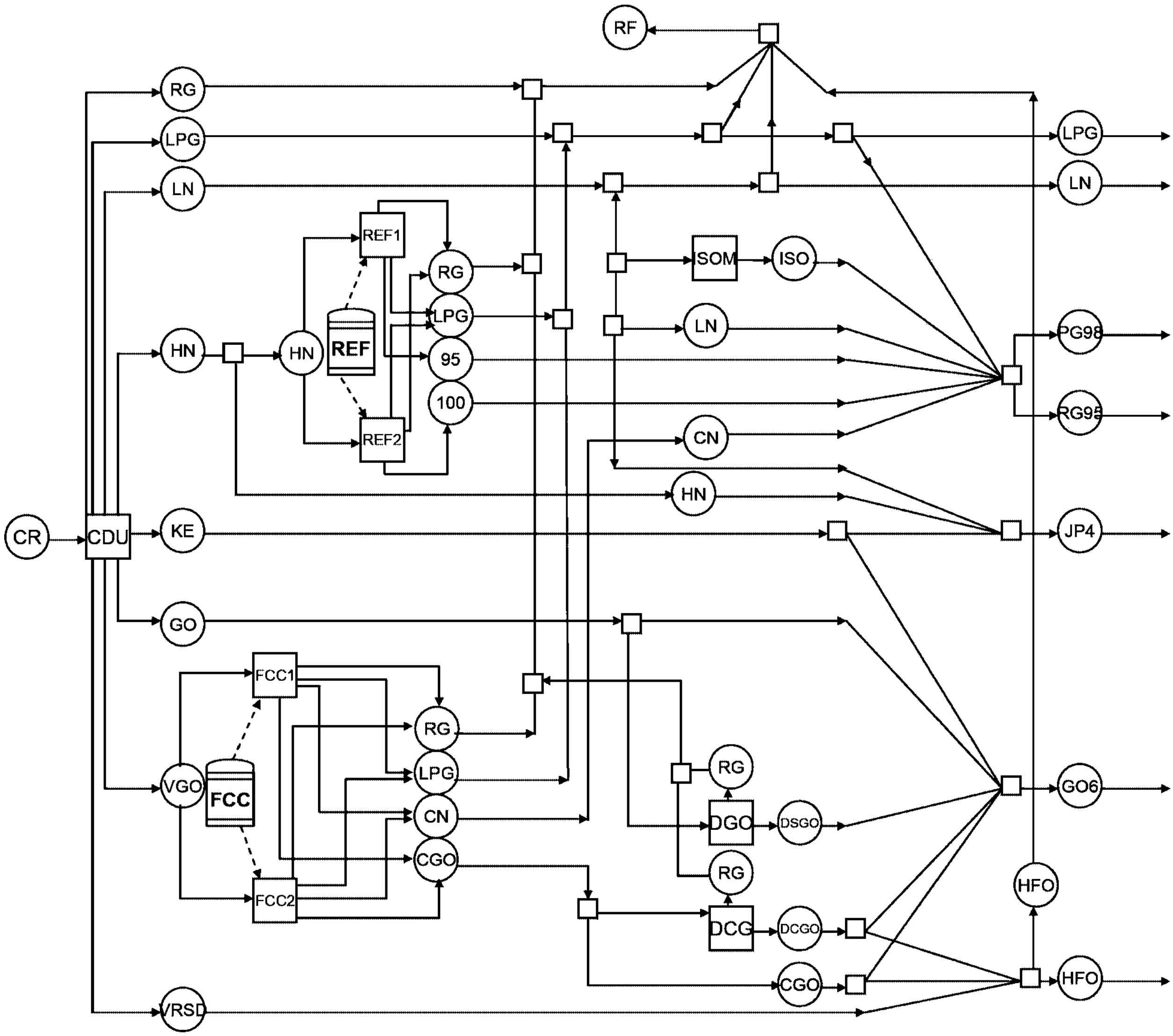

Figure 3. SEN representation of refinery 2.

polymers (plastics), solvents, fibers, detergents, ammonia, and other synthetic organic compounds for general use in the chemical industry.[11] In a situation where worldwide demand for these basic olefins is souring, more studies are being conducted to maximize steam cracking efficiency.[13] An alternative strategy would be to seek integration possibilities with the refinery as they both share feedstocks and products that can be utilized to maximize profit and processing efficiency.

2.3. Normal Paraffins and Cycloparaffins. Paraffin hydrocarbon compounds contain only single-bonded carbon atoms, which give them higher stability characteristics. Normal paraffin compounds are abundantly present in petroleum fractions but are mostly recovered from light straight-run (LSR) naphtha and kerosene. However, the nonnormal hydrocarbon components of LSR naphtha are of a higher octane number and therefore are preferred for gasoline blending.[14] For this reason, new technologies have been developed to further separate LSR naphtha into higher octane products that can be used in the gasoline pool and normal paraffins that are used as steam cracker feedstock (e.g., UOP IsoSiv Process). The normal paraffins recovered from kerosene, on the other hand, are mostly used in biodegradable detergents manufacturing.

Cycloparaffins, also referred to as naphthenes, are mainly produced by dehydrogenation of their equivalent aromatic compounds, such as the production of cyclohexane by dehydrogenation of benzene. Cyclohexane is mostly used for the production of adipic acid and nylon manufacturing.[11]

3. Refinery and Petrochemical Synergy Benefits

Process integration in the refining and petrochemical industries includes many intuitively recognized benefits of processing higher quality feedstocks, improving value of byproducts, and achieving better efficiencies through sharing of resources. Table 1 illustrates different refinery streams that can be of superior quality when used in the petrochemical industry.[15] The potential integration alternatives for the refining and petrochemical industries can be classified into three main categories: (1) process integration, (2) utility integration, and (3) fuel gas upgrade. The integration opportunities discussed below are for a general refinery and a petrochemical complex. Further details and analysis about the system requirements can be developed based on the actual system infrastructure, market demand, and product and energy prices.

3.1. Process Integration. The innovative design of different refinery processes while considering the downstream petrochemical industry is an illustration of the realization of refining and petrochemical integration benefits. This is demonstrated by the wide varieties of refinery cracking and reforming technologies that maximize olefin production. Some of the available technologies include cracking for high propylene and gasoline production,[16] maximum gasoline and LPG production, and low-pressure combination-bed catalytic reforming for aromatics.[17] Other technologies include different extractive treatments of refinery streams, e.g., aromatic recovery from light straight-run (LSR) naphtha. The normal paraffins of the LSR, on the other hand, are typically used as steam cracker feedstock.[14]

Reforming, as mentioned above, is the main source of aromatics in petroleum refining, where their yield increases with the increase in reforming severity. Aromatics in the reformate streams are recovered by extractive distillation using different solvents, depending on the chosen technology. The benzene−toluene−xylene (BTX) complex is one of the petrochemical processes that leverage a great deal of the integration benefits with petroleum refining. The integration benefits are not only limited to the process side but extend to the utilities as will be explained in the following section.

Pyrolysis gasoline (pygas), a byproduct of stream cracking, can be further processed in the BTX complex to recover the aromatic compounds, and the raffinate after extraction can be blended in the gasoline or naphtha pool.[12] If there is no existing aromatic complex to further process the pygas, it could alternatively be routed to the reformer feed for further processing.[18] However, this alternative may not be viable in general as most reformers run at maximum capacity. Pygas from steam cracking contains large amounts of diolefins, which are undesirable due to their instability and tendency to polymerize yielding filter-plugging compounds. For this reason, hydrogenation of pygas is usually recommended prior to further processing.

3.2. Utility Integration (Heat/Hydrogen/Steam/Power). Petroleum refining and the basic petrochemical industry are the most energy-intensive processes in the chemical process industry.[13] The energy sources in these processes assume different forms including fuel oil, fuel gas, electrical power, and both high- and low-pressure steam. The different energy requirements and waste from the whole range of refinery and petrochemical units present intriguing opportunities for an integrated complex. Integration of energy sources and sinks of steam cracking, for instance, with other industrial processes, particularly natural gas processing, can yield significant energy savings reaching up to 60%.[13] Furthermore, gas turbine integration (GTI) between petrochemical units and ammonia plants can lead to a reduction in energy consumption by up to 10% through exhaust-heat recovery.[3] This can be readily extended to the refinery processing units which span a wide variety of distillation, cracking, reforming, and isomerization processes.

Hydrogen is another crucial utility that is receiving more attention recently mainly due to the more strict environmental regulations on sulfur emissions. Reduction of sulfur emissions is typically achieved by deeper desulfurization of petroleum fuels, which in turn requires additional hydrogen production.[19] A less capital intensive alternative to alleviate hydrogen shortage is to operate the catalytic reformer at higher severity. However, higher severity reforming increases the production of BTX aromatics, which consequently affect the gasoline pool aromatics specification. Therefore, the BTX extraction process becomes a more viable alternative for the sake of aromatics recovery as well as maintaining the gasoline pool within specification.[19] The capital cost for the implementation of such a project would generally be lower as the BTX complex and refinery would share both process and utility streams.

3.3. Fuel Gas Upgrade. Refinery fuel gas is generated from refinery processes and mainly comprises C_1/C_2 fractions and some hydrogen. Considerable amounts of light hydrocarbons are produced from the different conversion units in the refinery and are collected in the common fuel gas system. For instance, FCC off-gas contains significant amounts of ethylene and propylene, which can be extracted and processed as petrochemical feedstocks. A number of integrated U.S. and European refineries have recognized and capitalized on this opportunity by recovering these high-value components.[3] This type of synergy requires proper planning and optimization between the petroleum refining and petrochemical complexes.

The other major component is hydrogen, which typically accounts for 50−80% of the refinery fuel gas.[20] This substantial amount of hydrogen is disposed to the fuel gas system from different sources in the refinery. The most significant source, however, is catalytic reforming. Hydrogen recovery using economically attractive technologies is of great benefit to both refineries and petrochemical systems especially with the increasingly strict environmental regulations on fuels.

4. Problem Statement

The optimization of refining and petrochemical processes involves a wide range of aspects varying from economic analysis and strategic expansions to crude oil selection, process level targets, operating modes, etc. The focus of this study is to develop a mathematical programming tool for the simultaneous design of an integrated network of refineries and petrochemical processes. On the refinery side, the model provides the optimal network integration between the refineries, process expansion requirements, operating policy based on different feedstock combination alternatives, process levels, and operating modes. On the petrochemical side, the model establishes the design of an optimal petrochemical process network from a range of process technologies to satisfy a given demand. The simultaneous network design and optimization of the refining and petrochemical industry provides an appropriate means for improving the coordination across the industrial system and is prone to developing an overall optimal production strategy across the petroleum chain.

The general problem under study can be defined as follows: A set of refinery products $cfr \in CFR$ produced at multiple refinery sites $i \in I$ and a set of petrochemical products $cp \in CFP$ is given. Each refinery consists of different production units $m \in M_{\mathrm{Ref}}$ that can operate at different operating modes $p \in P$ while a set of wide range of petrochemical and chemical process technologies $m \in M_{\mathrm{Pet}}$ to be selected from is given. Furthermore, different crude oil slates $cr \in CR$ are available and given. The petrochemical network selects its feedstock from three main sources: refinery intermediate streams $Fi^{\mathrm{Pet}}_{cr,cir,i}$ of an intermediate product $cir \in RPI$, refinery final products $Ff^{\mathrm{Pet}}_{cr,cfr,i}$ of a final product $cfr \in RPF$, and nonrefinery streams Fn^{Pet}_{cp} of a chemical $cp \in NRF$. The process network across the refineries and petrochemical system is connected in a finite number of ways. An integration superstructure between the refinery processes is defined in order to allow for possible intermediate stream exchange. Market product prices, operating costs at each refinery and petrochemical system, and product demands are assumed to be known.

The problem consists of determining the optimal integration and coordination strategy for the overall refinery network and designing the optimal petrochemical network required to satisfy a given demand from the available process technologies. The proposed approach will also provide overall refinery and petrochemical process production levels and detailed blending levels at each refinery site. The objective function is a minimization of the annualized cost over a given time horizon among the refineries by improving the coordination and utilization of excess capacities in each facility and maximization of the added value in the petrochemical system. Expansion requirements to improve production flexibility and reliability in the refineries are also considered.

For all refinery and petrochemical processes within the network, we assume that all material balances are expressed in terms of linear yield vectors. Even though this might sound restrictive as most if not all refinery and petrochemical processes are inherently nonlinear, this practice is commonly applied with such large-scale systems. Moreover, the decisions in this study are of a strategic level in which such linear formulation is adequate to address the required level of details involved at this stage. However, for operational planning and scheduling problems consideration of process nonlinearity is essential, and with that problem complexity becomes another issue to address. In this study, it is also assumed that processes have fixed capacities and the operating cost of each process and production mode is proportional to the process inlet flow. In the case of refinery product blending, quality blending indices are used to maintain model linearity. It is also assumed that all products that are in excess of the local demand can be exported to a global market. Piping and pumping installation costs to transport refinery intermediate streams between the different refinery sites as well as the operating costs of the new system are lumped into one fixed-charge cost. This was adjusted to the current year using the Marshall and Swift index and was found to be 8.9 US\$/ft. The cost used in the optimization problem was the uniform equivalent annual cost (UEAC) based on 20 years lifetime, 7% interest rate, and a salvage value of 5% of the installation cost. No inventories will be considered since the model is addressing strategic decisions which usually cover a long period of time. We also assume perfect mixing in the refineries and that the properties of each crude oil slate are decided by specific key components.

5. Model Formulation

The proposed formulation addresses the problem of simultaneous design of an integrated network of refineries and petrochemical processes. The proposed model is based on the formulations proposed by Al-Qahtani and Elkamel[8] and Al-Qahtani et al.[21] All material balances are carried out on a mass basis with the exception of refinery quality constraints of properties that only blend by volume where volumetric flow rates are used instead. The model is formulated as follows:

$$\min \sum_{cr\in CR}\sum_{i\in I} CrCost_{cr} S^{\mathrm{Ref}}_{cr,i} + \sum_{p\in P} OpCost_p \sum_{cr\in CR}\sum_{i\in I} z_{cr,p,i} + \sum_{cir\in CIR}\sum_{i\in I}\sum_{i'\in I} InCost_{i,i'} y_{\mathrm{pipe}^{\mathrm{Ref}}_{cir,i,i'}} + \sum_{i\in I}\sum_{m\in M_{\mathrm{Ref}}}\sum_{s\in S} InCost_{m,s} y_{\exp^{\mathrm{Ref}}_{m,i,s}} - \sum_{cfr\in PEX}\sum_{i\in I} Pr^{\mathrm{Ref}}_{cfr} e^{\mathrm{Ref}}_{cfr,i} - \sum_{cp\in CP}\sum_{m\in M_{\mathrm{Pet}}} Pr^{\mathrm{Pet}}_{cp}\delta_{cp,m} x^{\mathrm{Pet}}_{m} \quad \text{where } i\neq i' \tag{1}$$

subject to

$$z_{cr,p,i} = S^{\mathrm{Ref}}_{cr,i} \qquad \forall cr\in CR,\ i\in I \quad \text{where } p\in P' = \{\text{set of CDU processes } \forall \text{ plant } i\} \tag{2}$$

$$\sum_{p\in P}\alpha_{cr,cir,i,p} z_{cr,p,i} + \sum_{i'\in I}\sum_{p\in P}\xi_{cr,cir,i',p,i} xi^{\mathrm{Ref}}_{cr,cir,i',p,i} - Fi^{\mathrm{Pet}}_{cr,cir\in RPI,i} - \sum_{i'\in I}\sum_{p\in P}\xi_{cr,cir,i,p,i'} xi^{\mathrm{Ref}}_{cr,cir,i,p,i'} - \sum_{cfr\in CFR} w_{cr,cir,cfr,i} - \sum_{rf\in FUEL} w_{cr,cir,rf,i} = 0 \qquad \forall cr\in CR,\ cir\in CIR,\ i' \text{ and } i\in I \quad \text{where } i\neq i' \tag{3}$$

$$\sum_{cr\in CR}\sum_{cir\in CB} w_{cr,cir,cfr,i} - \sum_{cr\in CR}\sum_{rf\in FUEL} w_{cr,cfr,rf,i} - \sum_{cr\in CR} Ff^{\mathrm{Pet}}_{cr,cfr\in RPF,i} = x^{\mathrm{Ref}}_{cfr,i} \qquad \forall cfr\in CFR,\ i\in I \tag{4}$$

$$\sum_{cr\in CR}\sum_{cir\in CB}\frac{w_{cr,cir,cfr,i}}{sg_{cr,cir}} = xv^{\mathrm{Ref}}_{cfr,i} \qquad \forall cfr\in CFR,\ i\in I \tag{5}$$

$$\sum_{cir\in FUEL} cv_{rf,cir,i} w_{cr,cir,rf,i} + \sum_{cfr\in FUEL} w_{cr,cfr,rf,i} - \sum_{p\in P}\beta_{cr,rf,i,p} z_{cr,p,i} = 0 \qquad \forall cr\in CR,\ rf\in FUEL,\ i\in I \tag{6}$$

$$\sum_{cr\in CR}\sum_{cir\in CB}\left(att_{cr,cir,q\in Qv}\frac{w_{cr,cir,cfr,i}}{sg_{cr,cir}} + att_{cr,cir,q\in Qw}\left[w_{cr,cir,cfr,i} - \sum_{rf\in FUEL} w_{cr,cfr,rf,i} - \sum_{cr\in CR} Ff^{\mathrm{Pet}}_{cr,cfr\in RPF,i}\right]\right) \geq q^{\mathrm{L}}_{cfr,q\in Qv} xv^{\mathrm{Ref}}_{cfr,i} + q^{\mathrm{L}}_{cfr,q\in Qw} x^{\mathrm{Ref}}_{cfr,i} \qquad \forall cfr\in CFR,\ q=\{Qw,Qv\},\ i\in I \tag{7}$$

$$\sum_{cr\in CR}\sum_{cir\in CB}\left(att_{cr,cir,q\in Qv}\frac{w_{cr,cir,cfr,i}}{sg_{cr,cir}} + att_{cr,cir,q\in Qw}\left[w_{cr,cir,cfr,i} - \sum_{rf\in FUEL} w_{cr,cfr,rf,i} - \sum_{cr\in CR} Ff^{\mathrm{Pet}}_{cr,cfr\in RPF,i}\right]\right) \leq q^{\mathrm{U}}_{cfr,q\in Qv} xv^{\mathrm{Ref}}_{cfr,i} + q^{\mathrm{U}}_{cfr,q\in Qw} x^{\mathrm{Ref}}_{cfr,i} \qquad \forall cfr\in CFR,\ q=\{Qw,Qv\},\ i\in I \tag{8}$$

$$MinC_{m,i} \leq \sum_{p\in P}\gamma_{m,p}\sum_{cr\in CR} z_{cr,p,i} \leq MaxC_{m,i} + \sum_{s\in S} AddC_{m,i,s} y_{\exp^{\mathrm{Ref}}_{m,i,s}} \qquad \forall m\in M_{\mathrm{Ref}},\ i\in I \tag{9}$$

$$\sum_{cr\in CR}\sum_{p\in P}\xi_{cr,cir,i,p,i'} xi^{\mathrm{Ref}}_{cr,cir,i,p,i'} \leq F^{\mathrm{U}}_{cir,i,i'} y_{\mathrm{pipe}^{\mathrm{Ref}}_{cir,i,i'}} \qquad \forall cir\in CIR,\ i' \text{ and } i\in I \quad \text{where } i\neq i' \tag{10}$$

$$\sum_{i\in I}\left(x^{\mathrm{Ref}}_{cfr,i} - e^{\mathrm{Ref}}_{cfr',i}\right) \geq D_{\mathrm{Ref}_{cfr}} \qquad \forall cfr \text{ and } cfr' \quad \text{where } cfr\in CFR,\ cfr'\in PEX \tag{11}$$

$$IM^{\mathrm{L}}_{cr} \leq \sum_{i\in I} S^{\mathrm{Ref}}_{cr,i} \leq IM^{\mathrm{U}}_{cr} \qquad \forall cr\in CR \tag{12}$$

$$Fn^{\mathrm{Pet}}_{cp\in NRF} + \sum_{i\in I}\sum_{cr\in CR} Fi^{\mathrm{Pet}}_{cr,cp\in RPI,i} + \sum_{i\in I}\sum_{cr\in CR} Ff^{\mathrm{Pet}}_{cr,cp\in RPF,i} + \sum_{m\in M_{\mathrm{Pet}}}\delta_{cp,m} x^{\mathrm{Pet}}_{m} \geq D_{\mathrm{Pet}^{\mathrm{L}}_{cp\in CFP}} \qquad \forall cp\in CP \tag{13}$$

$$Fn^{\mathrm{Pet}}_{cp\in NRF} + \sum_{i\in I}\sum_{cr\in CR} Fi^{\mathrm{Pet}}_{cr,cp\in RPI,i} + \sum_{i\in I}\sum_{cr\in CR} Ff^{\mathrm{Pet}}_{cr,cp\in RPF,i} + \sum_{m\in M_{\mathrm{Pet}}}\delta_{cp,m} x^{\mathrm{Pet}}_{m} \leq D_{\mathrm{Pet}^{\mathrm{U}}_{cp\in CFP}} \qquad \forall cp\in CP \tag{14}$$

$$B^{\mathrm{L}}_{m} y_{\mathrm{proc}^{\mathrm{Pet}}_{m}} \leq x^{\mathrm{Pet}}_{m} \leq K^{\mathrm{U}} y_{\mathrm{proc}^{\mathrm{Pet}}_{m}} \qquad \forall m\in M_{\mathrm{Pet}} \tag{15}$$

$$\sum_{cp\in CIP} y_{\mathrm{proc}^{\mathrm{Pet}}_{m}} \leq 1 \qquad \forall m\in M_{\mathrm{Pet}} \text{ that produce } cp\in CIP \tag{16}$$

$$\sum_{cp\in CFP} y_{\mathrm{proc}^{\mathrm{Pet}}_{m}} \leq 1 \qquad \forall m\in M_{\mathrm{Pet}} \text{ that produce } cp\in CFP \tag{17}$$

$$Fn^{\mathrm{Pet}}_{cp} \leq S^{\mathrm{Pet}}_{cp} \qquad \forall cp\in NRF \tag{18}$$

The above objective function (1) represents a minimization of the annualized cost which consists of crude oil cost, refinery operating costs, refinery intermediate exchange piping costs, and refinery production system expansion cost, less the refinery export revenue and added value by the petrochemical processes.

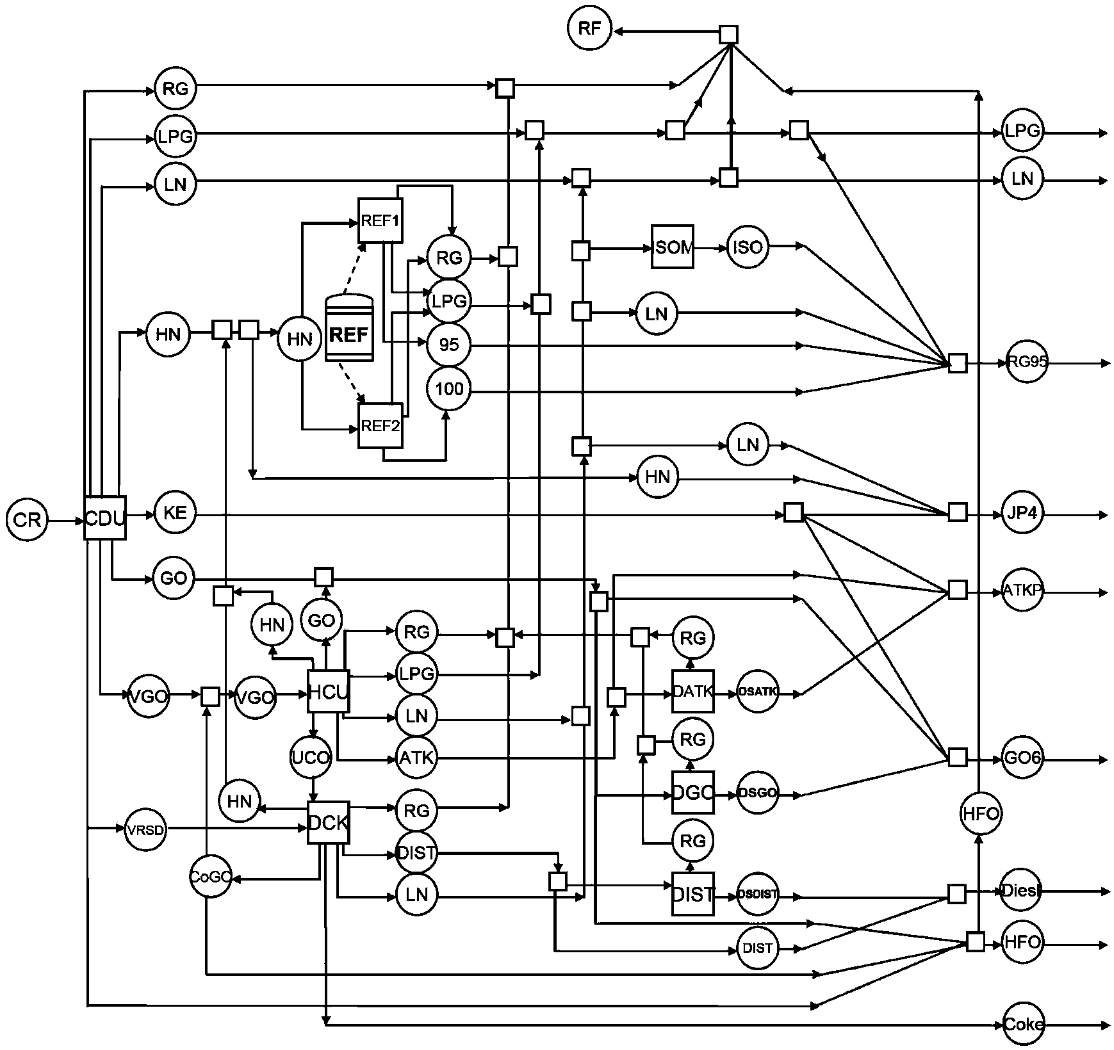

Figure 4. SEN representation of refinery 3.

The operating cost of each refinery process is assumed to be proportional to the process inlet flow and is expressed on a yearly basis. Inequality 2 corresponds to each refinery raw materials balance where throughput to each distillation unit $p \in P'$ at plant $i \in I$ from each crude type $cr \in CR$ is equal to the available supply $S_{cr,i}$. Constraint 3 represents the intermediate material balances within and across the refineries, where the coefficient $\alpha_{cr,cir,i,p}$ can assume either a positive sign if it is an input to a unit or a negative sign if it is an output from a unit. The multirefinery integration matrix $\xi_{cr,cir,i,p,i'}$ accounts for all possible alternatives of connecting intermediate streams $cir \in CIR$ of crude $cr \in CR$ from refinery $i \in I$ to process $p \in P$ in plant $i' \in I'$. Variable $xi^{\text{Ref}}_{cr,cir,i,p,i'}$ represents the transshipment flow rate of crude $cr \in CR$, of intermediate $cir \in CIR$ from plant $i \in I$ to process $p \in P$ at plant $i' \in I$. Constraint 3 also considers the petrochemical network feedstock from the refinery intermediate streams $Fi^{\text{Pet}}_{cr,cir,i}$ of each intermediate product $cir \in RPI$. The material balance of final products in each refinery is expressed as the difference between flow rates from intermediate streams $w_{cr,cir,cfr,i}$ for each $cir \in CIR$ that contribute to the final product pool and intermediate streams that contribute to the fuel system $w_{cr,cfr,rf,i}$ for each $rf \in FUEL$ less the refinery final products $Ff^{\text{Pet}}_{cr,cfr,i}$ for each $cfr \in RPF$ that are fed to the petrochemical network as shown in constraint 4. In constraint 5 we convert the mass flow rate to volumetric flow rate by dividing it by the specific gravity $sg_{cr,cir}$ of each crude type $cr \in CR$ and intermediate stream $cir \in CB$. This is needed in order to express the quality attributes that blend by volume in blending pools. Constraint 6 is the fuel system material balance, where the term $cv_{rf,cir,i}$ represents the caloric value equivalent for each intermediate $cir \in CB$ used in the fuel system at plant $i \in I$. The fuel production system can consist of either a single or combination of intermediates $w_{cr,cir,rf,i}$ and products $w_{cr,cfr,rf,i}$. The matrix $\beta_{cr,rf,i,p}$ corresponds to the consumption of each processing unit $p \in P$ at plant $i \in I$ as a percentage of unit throughput. Constraints 7 and 8, respectively, represent lower and upper bounds on refinery quality constraints for all refinery products that either blend by mass $q \in Q_w$ or by volume $q \in Q_v$. Constraint 9 represents the maximum and minimum allowable flow rates to each processing unit. The coefficient $\gamma_{m,p}$ is a zero−one matrix for the assignment of production unit $m \in M_{\text{Ref}}$ to process operating mode $p \in P$. The term $AddC_{m,i,s}$ accounts for the additional refinery expansion capacity of each production unit $m \in M_{\text{Ref}}$ at refinery $i \in I$ for a specific expansion size $s \in S$. In this formulation, we only allow the addition of predetermined capacities whose pricing can be acquired a priori through design companies' quotations. The integer variable $y_{\exp^{\text{Ref}}_{m,i,s}}$ represents the decision of expanding a production unit, and it can take a value of 1 if the unit expansion is required or 0 otherwise. Constraint 10 sets an upper bound on intermediate stream flow rates between the different refineries. The integer variable $y_{\text{pipe}^{\text{Ref}}_{cir,i,i'}}$ represents the decision of exchanging intermedi-

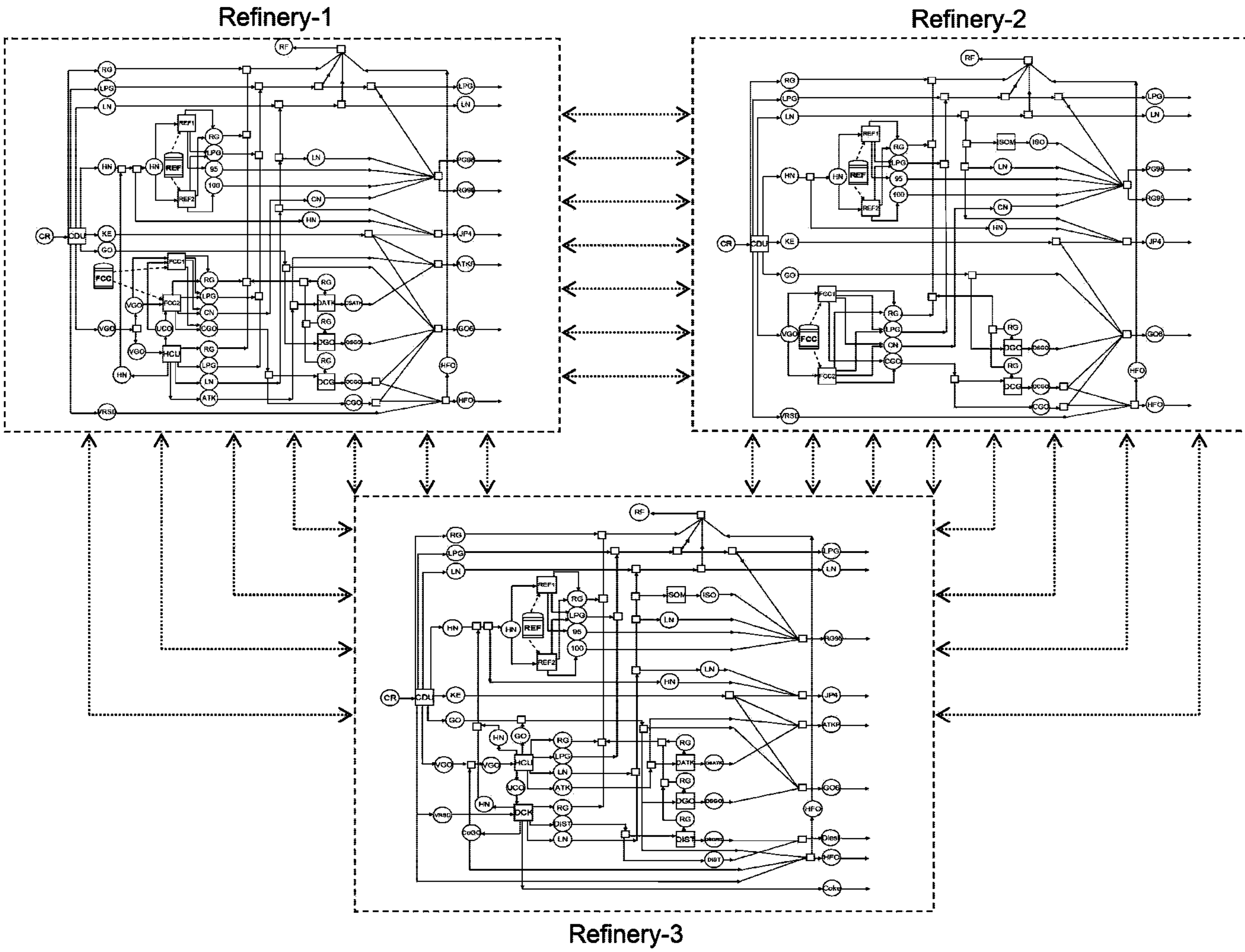

Figure 5. SEN representation of the refinery integration network.

Table 2. Major Refinery Network Capacity Constraints

	higher limit (10^3 tons/year)		
	R1	R2	R3
Production Capacity			
distillation	4500.0	12000.0	9900.0
reforming	700.0	2000.0	1800.0
isomerization	200.0	–	450.0
fluid catalytic cracker	800.0	1400.0	–
hydrocracker	–	1800.0	2400.0
delayed coker	–	–	1800
desulfurized gas oil	1300.0	3000.0	2400.0
desulfurized cycle gas oil	200.0	750.0	–
desulfurized ATK	–	1200.0	1680.0
desulfurized distillates	–	–	450.0
Crude Availability			
Arabian Light		31200.0	
Local Demand			
LPG		432.0	
LN		–	
PG98		540.0	
PG95		4440.0	
JP4		2340.0	
GO6		4920.0	
ATK		1800.0	
HFO		200.0	
Diesl		400.0	
Coke		300.0	

ate products between the refineries and takes on the value of 1 if the commodity is transferred from plant $i \in I$ to plant $i' \in I$ or 0 otherwise, where $i \neq i'$. When an intermediate stream is selected to be exchanged between two refineries, its flow rate must be below the transferring pipeline capacity $F^{\mathrm{U}}_{cir,i,i'}$. Constraint 11 stipulates that the final products from each refinery $x^{\mathrm{Ref}}_{cfr,i}$ less the amount exported $e^{\mathrm{Ref}}_{cfr',i}$ for each exportable product $cfr' \in PEX$ from each plant $i \in I$ must satisfy the domestic demand $D_{\mathrm{Ref}_{cfr}}$. Resources are limited by constraint 12, which imposes upper and lower bounds on the available feedstock $cr \in CR$ to the refineries.

Constraints 13 and 14 represent the material balance that governs the operation of the petrochemical system. The petrochemical network receives its feed from potentially three main sources. These are (1) refinery intermediate streams $Fi^{\mathrm{Pet}}_{cr,cir,i}$ of an intermediate product $cir \in RPI$, (2) refinery final products $Ff^{\mathrm{Pet}}_{cr,cfr,i}$ of a final product $cfr \in RPF$, and (3) nonrefinery streams Fn^{Pet}_{cp} of a chemical $cp \in NRF$. For a given subset of chemicals $cp \in CP$, the proposed model selects the feed types, quantity, and network configuration based on the final chemical and petrochemical lower and upper product demands $D_{\mathrm{Pet}^{\mathrm{L}}_{cp}}$ and $D_{\mathrm{Pet}^{\mathrm{U}}_{cp}}$ for each $cp \in CFP$, respectively. In constraint 15, defining a binary variable $y_{\mathrm{proc}^{\mathrm{Pet}}_{m}}$ for each process $m \in M_{\mathrm{Pet}}$ is required for the process selection requirement as $y_{\mathrm{proc}^{\mathrm{Pet}}_{m}}$ will equal 1 only if process m is selected or 0 otherwise. Furthermore, if only process m is selected, its production level must be at least equal to the process minimum economic capacity B^{L}_{m} for each $m \in M_{\mathrm{Pet}}$, where K^{U} is a valid upper bound. In the case where it is preferred to choose only one process technology to produce a chemical, constraints 16 and 17 can be included for each

Table 3. Major Products and Process Technologies Considered in the Petrochemical Complex

product	sale price[a] ($/ton)	process technology	process index	min econ prod (10^3 tons/year)
ethylene (E)	1570	pyrolysis of naphtha (low severity)	1	250
		pyrolysis of gas oil (low severity)	2	250
		steam cracking of naphtha (high severity)	3	250
		steam cracking of gas oil (high severity)	4	250
ethylene dichloride (EDC)	378	chlorination of ethylene	5	180
		oxychlorination of ethylene	6	180
vinyl chloride monomer (VCM)	1230	chlorination and oxychlorination of ethylene	7	250
		dehydrochlorination of ethylene dichloride	8	125
poly(vinyl chloride) (PVC)	1600	bulk polymerization	9	50
		suspension polymerization	10	90

[a] All chemical prices in this study were obtained from the latest CW Price Reports in *Chemical Week*.

intermediate product $cp \in CIP$ and final product $cp \in CFP$, respectively. Finally, we can specify limitations on the supply of feedstock Fn_{cp}^{Pet} for each chemical type $cp \in NRF$ though constraint 18. Bear in mind that the limitations on the refinery intermediate product $Fi_{cr,cir,i}^{\text{Pet}}$ and final product $Ff_{cr,cfr,i}^{\text{Pet}}$ that are fed to the petrochemical network are dictated by the model based on both refinery and petrochemical demand and price structure.

6. Illustrative Case Study

In this section we demonstrate the performance of the proposed model on an industrial-scale case study. Instead of considering the full-scale petrochemical network which may have limited application, we consider a special case of the integration problem. Although the proposed formulation covers the full-scale refinery network and petrochemical systems, the case study will consider the integration of a petrochemical complex for the production of poly(vinyl chloride) (PVC) with a multirefinery network. PVC is one of the major ethylene derivatives that has many important applications and uses including pipe fittings, automobile bumpers, toys, bottles, and many others.[11]

Direct integration of refining and ethylene cracking is considered the essential building block in achieving the total petrochemical integration.[6,22] This problem has received more attention lately due to soaring motor gasoline prices and the directly related prices of ethylene feedstocks.[23] This kind of volatility in prices has prompted a shift in ethylene feedstock selection and economics to either lighter or heavier refinery product slates.[24] Shifting from one feedstock to another will mainly depend on the market price structure and demand for refinery products. Some researchers believe that the tendency of ethylene feedstock shift would mainly be toward heavier refinery streams including heavy and vacuum gas oils due to the diminishing reserves of sweet crudes and decreasing demand of heavy fraction fuels.[25,26] This change in ethylene feedstock selection and the direct effect on the refinery products requires adequate decision making and analysis that takes into account both refining and petrochemical markets.

In the case study, we consider the planning for three complex refineries by which we demonstrate the performance of our model in devising an overall production plan as well as an integration strategy among the refineries. The state equipment network (SEN) representation of the three refineries are shown in Figures 2, 3, and 4, while the overall topology of the refinery network is illustrated in Figure 5. The atmospheric crude unit separates crude oil into several fractions including LPG, naphtha, kerosene, gas oil, and residues. The heavy residues are then sent to the vacuum unit, where they are further separated into vacuum gas oil and vacuum residues. Depending on the production targets, different processing and treatment processes are applied to the crude fractions. In this example, the naphtha is further separated into heavy and light naphtha. Heavy naphtha is sent to the catalytic reformer unit to produce high octane reformates for gasoline blending, and light naphtha is sent to the light naphtha pool and to an isomerization unit to produce isomerate for gasoline blending, too, as in refineries 2 and 3. The middle distillates are combined with other similar intermediate streams and sent for hydrotreating and then for blending to produce jet fuels and gas oils. Atmospheric and vacuum gas oils are further treated by catalytic cracking, as in refinery 2, or by hydrocracking, as in refinery 3, or by both, as in refinery 1, to increase the gasoline and distillate yields. In refinery 3, vacuum residues are further treated using cooking and thermal processes to increase light products yields. The final products of the three-refinery network consists of liquefied petroleum gas (LPG), light naphtha (LT), two grades of gasoline (PG98 and PG95), No. 4 jet fuel (JP4), military jet fuel (ATKP), No. 6 gas oil (GO6), diesel fuel (Diesl), heating fuel oil (HFO), and petroleum coke (Coke). The major capacity constraints for the refinery network are given in Table 2. Furthermore, the three refineries are assumed to be in one industrial area, which is a common situation in many locations around the world, and are coordinated through a main headquarters sharing the feedstock supply. The cost parameters for pipeline installation were calculated as cost per distance between the refineries, and then multiplied by the required pipe length in order to connect any two refineries. The pipeline diameters considered in all cases was 8 in.

The petrochemical complex for the production of PVC starts with the production of ethylene from the refinery feedstocks. The main feedstocks to the ethylene plant in our study are light naphtha (LN) and gas oil (GO). The selection of the feedstocks and hence the process technologies is decided based on the optimal balance and trade-off between the refinery and petrochemical markets. The process technologies considered in this study for the production of PVC are listed in Table 3. The overall topology of all petrochemical technologies considered for the PVC production is shown in Figure 6.

From the refinery side, the proposed model will provide the optimal network integration between the refineries, process expansion requirements, operating policy based on different feedstock combination alternatives, process levels, and operating modes. On the petrochemical side, the model will establish the design of an optimal petrochemical process network for the production of PVC from the range of process technologies and feedstocks available to satisfy a given demand. This problem was formulated as an MILP with the overall objective of minimizing the total annualized cost of the refinery and maximizing the added value from the PVC petrochemical network. Maximizing the added value of the petrochemical network is appropriate since the feedstock costs contribute to the majority of the total cost. For instance, the feedstock cost

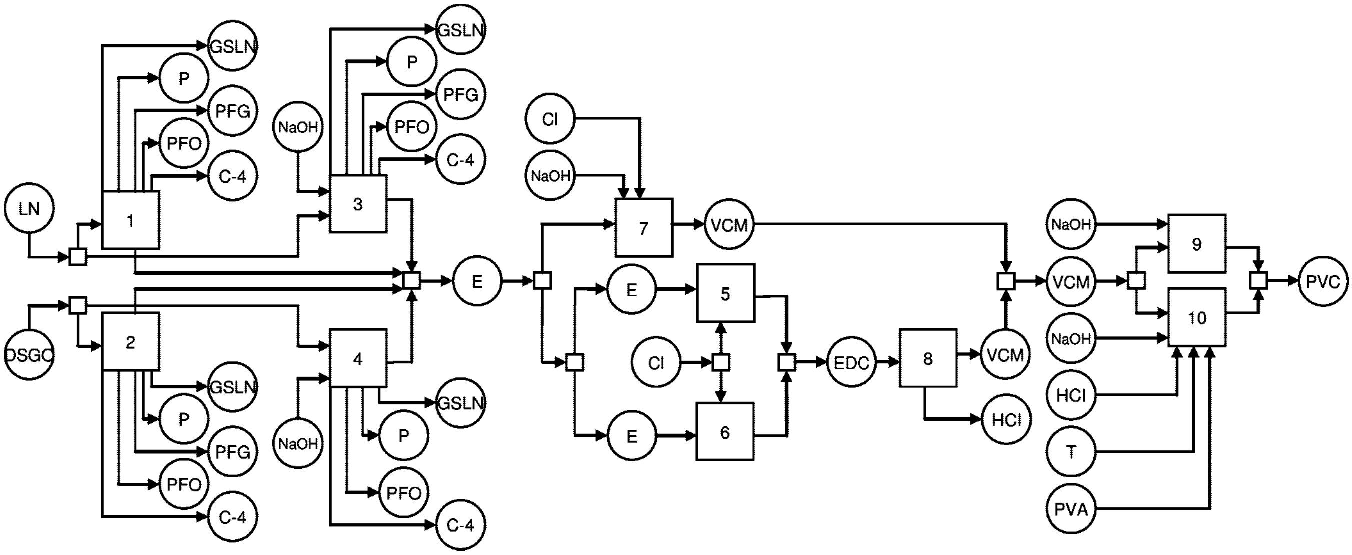

Figure 6. SEN representation of the PVC petrochemical complex possible routes.

Table 4. Model Results of Multirefinery Network

process variables			results (10^3 tons/year) R1	R2	R3
crude oil supply			4500.0	12000.0	9900.0
production levels	crude unit		4500.0	12000.0	9900.0
	reformer		573.1	1824.5	1800.0
	isomerization		200.0	–	450.0
	FCC		640.0	1400.0	–
	hydrocracker		–	1740.4	2400
	delayed coker		–	–	1484
	desulfurized gas oil		1084.6	2763.7	2383.8
	desulfurized cycle gas oil		200.0	600.0	–
	desulfurized ATK		–	1200.0	1654.8
	desulfurized distillates		–	–	360
intermediate streams exchange	from R1	VGO	–	–	446.0[a]
		VRSD	–	–	380.4[b]
	from R2	LN	–	–	340.0[c]
	from R3	LN	86.1[c]	–	–
		VGO	–	144.7[d]	–
		UCO	–	130.1[d]	–

	process variables	results (10^3 tons/year)
exports	PG95	273.2
	JP4	1359.5
	GO6	3695.7
	HFO	1579.8
	ATK	1767.5
	Coke	203.9
	Diesl	109.7
total cost ($/year)	$9,331,000	

[a] To HCU. [b] To coker. [c] To isomerization. [d] To FCC.

of the ethylene plant contributes to more than 87% of the total cost when naphtha is used and 84% and 74% when propane and ethane are used, respectively.[27]

The modeling system GAMS[28] is used for setting up the optimization models. The computational tests were carried out on a Pentium M processor 2.13 GHz, and the MILP problems were solved with CPLEX.[29]

The problem was first solved for the refinery network separately in order to compare and illustrate the effect of considering the PVC complex on the refinery network design and operating policies in terms of refinery product blending and process modes. Table 4 shows the optimal network integration design and operating policies of the refineries. The three refineries collaborated to satisfy a given local market demand, and the model proposed the production and blending level targets for the individual sites. The annual production cost across the facilities was found to be $9,331,000. The model required 2.4 CPU s to converge to the optimal solution.

The model was then solved for the total refinery network and the PVC petrochemical complex. As shown in Table 5, the proposed model redesigned the refinery network and operating policies and also devised the optimal production plan for the PVC complex from all available process technologies. The

Table 5. Model Results of Refinery and Petrochemical Networks

Refinery

process variables			R1 (results, 10^3 tons/year)	R2	R3
crude oil supply			4500.0	12000.0	9900.0
production levels	crude unit		4500.0	12000.0	9900.0
	Reformer		573.1	1824.6	1793.5
	isomerization		200.0	–	450.0
	FCC		640.0	1400.0	–
	hydrocracker		–	1740.4	2400.0
	delayed coker		–	–	1440.0
	desulfurized gas oil		1300.0	3000.0	2400.0
	desulfurized cycle gas oil		200.0	600.0	–
	desulfurized ATK		–	1200.0	1654.8
	desulfurized distillates		–	–	360.0
intermediate streams exchange	from R1	VGO	–	204.4[a]	301.2[a]
	from R2	LN	–	–	321.2[b]
		VRSD	–	–	267.6[c]
	from R3	LN	68.0[c]		
		UCO	–	100.3[d]	–

process variables		results (10^3 tons/year)
exports	PG95	265.2
	JP4	1365.5
	GO6	1503.4
	HFO	1658.9
	ATK	1767.5
	Coke	178.8
	Diesel	84.0

Petrochemical Network

process variables		R1 (results, 10^3 tons/year)	R2	R3
refinery feed to PVC complex	gas oil	1162.4	920.0	71.3
production levels	steam cracking of gas oil (4)		552.2	
	Cl and OxyCl E (7)		459.1	
	bulk polymer (9)		204.0	
final product	PVC		204.0	
total cost ($/year)	$8,948,000			

[a] To HCU. [b] To isomerization. [c] To coker. [d] To FCC.

model selected gas oil, an intermediate refinery stream, as the feedstock to the ethylene cracker as opposed to the more commonly used light naphtha feedstock. In fact, this selection provided the optimal strategy as the light naphtha stream was used instead in the gasoline pool for maximum gasoline production. PVC production was proposed by first high severity steam cracking of gas oil to produce ethylene. Vinyl chloride monomer (VCM) is then produced through the chlorination and oxychlorination of ethylene and finally, VCM is converted to PVC by bulk polymerization. The simultaneous optimization of the refinery and petrochemical network had an impact on both the refinery intermediate exchange network and production levels, as shown in Tables 4 and 5. For example, the capacity utilization of the desulfurization of gas oil process increased from 83%, 92%, and 99% in refineries 1, 2, and 3, respectively, to 100% in all refineries when the petrochemical network was considered in the model. The annual production cost across the facilities was found to be $8,948,000, and the model converged to the optimal solution after 3.1 CPU s.

7. Conclusion

A mixed-integer programming model for designing an integration and coordination policy among multiple refineries and a petrochemical network was presented. The objective was to develop a simultaneous methodology for designing a process integration network between petroleum refining and the petrochemical industry. A three large-scale refinery network and a PVC petrochemical complex were integrated to illustrate the performance of the proposed design methodology and to show the economic potential and trade-offs involved in the optimization of such systems. The study showed that the optimization of the downstream petrochemical industry has an impact on the multirefinery network integration and coordination strategies. This result emphasizes the importance of developed methodology.

In our study, however, all parameters were assumed to be known with certainty. Nevertheless, the current situation of fluctuating and high petroleum crude oil prices, changes in demand, and the direct effect this can have on the downstream petrochemical system underlines the importance of considering uncertainties. For example, the availability of crude oils, feedstock and chemical prices, and market demands for finished products will have a direct impact on the output of the highly strategic decisions involved in our study. Therefore, and acknowledging the shortcomings of deterministic models, the next phase of our investigation will be the consideration of uncertainties in the integration problem.

Acknowledgment

This work was supported in part by Saudi Aramco and by the Natural Sciences and Engineering Research Council of Canada (NSERC).

Nomenclature

Indices

i = plant
m = production units
p = processes
s = refinery units predetermined expansion size
j = iterations in the SAA method
q = quality specification
cir = refinery intermediate stream
cr = refinery raw material
cp = petrochemical commodities
cfr = refinery final product
rf = refinery fuel

Sets

I = plants
M_{Ref} = refinery units
M_{Pet} = petrochemical units
P = processes
S = refinery units predetermined expansion sizes
CR = refinery raw materials
CP = petrochemical commodities
CB = streams for blending refinery products
CFR = refinery final products
CIR = refinery intermediate streams
CIP = petrochemical intermediate streams
CFP = petrochemical final products
RPI = refinery intermediates as petrochemical feed
RPF = refinery final products as petrochemical feed
NRF = nonrefinery petrochemical feed
RF = refinery fuel streams
PEX = products for exports
$FUEL$ = streams comprising refinery fuel
Qw = quality of products that blend by weight
Qv = quality of products that blend by volume

Parameters

$\alpha_{cr,cir,i,p}$ = refinery input−output coefficients of intermediate stream cir from crude cr at plant i by process p
$\beta_{cr,rf,i,p}$ = refinery fuel consumption coefficients of refinery fuel rf from crude cr at plant i by process p
$\gamma_{m,p}$ = refinery assignment of process p to equipment m
$\xi_{cr,cir,i,p,i'}$ = refinery integration superstructure of all possible alternatives to transfer crude cr of commodity cir from plant i to process p in plant i'
$\delta_{cp,m}$ = petrochemical input−output coefficients of commodity cp in process m
$cv_{rf,cir,i}$ = refinery caloric value equivalent of refinery fuel rf by commodity cir at plant i
$MaxC_{m,i}$ = max capacity of production unit m at plant i
$MinC_{m,i}$ = min capacity required of production unit m at plant i
$AddC_{m,i}$ = additional capacity of production unit m at plant i
$OpCost_p$ = refinery operating cost of process p
$InCost_{i,i'}$ = insulation cost of piping to transfer refinery commodity cir from plant i to plant i'
$InCost_{m,s}$ = installation cost of a refinery production unit m
$CrCost_{cr}$ = price of crude cr
Pr^{Ref}_{cfr} = export price of product cfr
Pr^{Pet}_{cp} = price of petrochemical commodity cp
$D_{\mathrm{Ref}_{cfr}}$ = refinery demand of products cfr
$D_{\mathrm{Pet}^{\mathrm{L}}_{cp}}$ = petrochemical lower demand of product $cp \in CFP$
$D_{\mathrm{Pet}^{\mathrm{U}}_{cp}}$ = petrochemical upper demand of product $cp \in CFP$
$q^{\mathrm{L}}_{ci,q}$ = quality bounds of commodity ci of property q
$q^{\mathrm{U}}_{ci,q}$ = upper level bounds of commodity ci of property q
IM^{U}_{cr} = import upper bound level of commodity cr
IM^{L}_{cr} = import lower bound level of commodity cr
$att_{cr,ci,q}$ = attributes of intermediate streams ci produced from crude cr blending of property q
$sg_{cr,ci}$ = specific gravity of commodity ci by crude cr
$F^{\mathrm{U}}_{ci,i,i'}$ = flow rate upper bound of intermediate ci from plant i to i'
B^{L}_{m} = petrochemical process minimum economic capacity
K^{U} = petrochemical process production upper bound
S^{Pet}_{cp} = nonrefinery petrochemical feed supply of commodity $cp \in NRF$

Variables

$z_{cr,p,i}$ = process input flow rate of crude cr to process p at plant i
$w_{cr,ci,cfr,i}$ = blending levels of crude cr that produces intermediate cir to yield a product cfr at plant i
Fn^{Pet}_{cp} = petrochemical feed from nonrefinery commodity cp
$Fi^{\mathrm{Pet}}_{cr,cp,i}$ = petrochemical feed from refinery intermediates $cir \in RPI \in CP$ from plant i
$Ff^{\mathrm{Pet}}_{cr,cp,i}$ = petrochemical feed from refinery final product $cfr \in RPF \in CP$ from plant i
$x^{\mathrm{Ref}}_{cfr,i}$ = mass flow rate of refinery final product cfr by refinery i
$xv^{\mathrm{Ref}}_{cfr,i}$ = volumetric flow rate of refinery final product cfr by refinery i
x^{Pet}_{m} = mass flow rate of petrochemical production unit m
$xi^{\mathrm{Ref}}_{cr,cir,i,p,i'}$ = refinery transshipment level of crude cr of commodity cir from plant i to process p at plant i'
$S^{\mathrm{Ref}}_{cr,i}$ = refinery supply of raw material cr to refinery i
$e^{\mathrm{Ref}}_{cfr',i}$ = refinery exports of final product $cfr \in PEX$ from refinery i
$y_{\mathrm{pipe}^{\mathrm{Ref}}_{cir,i,i'}}$ = binary variable representing transshipment of refinery commodity cir from plant i to plant i'
$y_{\mathrm{exp}^{\mathrm{Ref}}_{m,i,s}}$ = binary variable representing refinery expansions of production unit m at plant i for a specific expansion size s
$y_{\mathrm{proc}^{\mathrm{Pet}}_{m}}$ = binary variable representing petrochemical selection of unit m

Processing Unit

DATK = desulfurization of aviation turbine kerosene
DCG = desulfurization of cycle gas oil
DCK = delayed coker
DGO = desulfurization of gas oil
CDU = crude distillation
DIST = desulfurization of delayed coker distillates
FCC1 = fluid catalytic cracker (gasoline mode)
FCC2 = fluid catalytic cracker (gas oil mode)
HCU = hydrocracker
ISOM = isomerization
REF1 = reformer (95% severity)
REF2 = reformer (100% severity)

Stream

100 = reformate at 100% severity
95 = reformate at 95% severity
ATK = aviation turbine kerosene intermediates
ATKP = aviation turbine kerosene product
C-4 = C-4 fractions (mixed butanes, butenes, etc.)
CGO = cycle gas oil
Cl = chlorine

CN = fluid catalytic cracker gasoline
CoGO = coker gas oil
Coke = petroleum coke
CR = crude oil
DCGO = desulfurized cycle gas oil
Diesl = petroleum diesel product
DIST = distillate
DSATK = desulfurized aviation turbine kerosene
DSDIST = desulfurized distillate
DSGO = desulfurized gas oil
E = ethylene
EDC = ethylene dichloride
GO = gas oil
GO6 = No. 6 gas oil
GSLN = petrochemical gasoline
HCI = hydrochloric acid
HN = heavy naphtha
HFO = petroleum heating fuel oil
ISO = isomerate
JP4 = No. 4 jet fuel
KE = kerosene
LN = light naphtha
LPG = liquefied petroleum gas
NaOH = sodium hydroxide
P = propylene
PFG = petrochemical fuel gas
PFO = petrochemical fuel oil
PG95 = refinery gasoline with 95 octane number
PG98 = refinery gasoline with 98 octane number
PVA = poly(vinyl alcohol)
PVC = poly(vinyl chloride)
RF = refinery fuel
RG = refinery gas
T = toluene
UCO = unconverted gas oil
VCM = vinyl chloride monomer
VGO = vacuum gas oil
VRSD = desulfurized vacuum residue

Literature Cited

(1) Sahinidis, N. V.; Grossmann, I. E.; Fornari, R. E.; Chathrathi, M. Optimization model for long range planning in chemical industry. *Comput. Chem. Eng.* **1989**, *13*, 1049.

(2) Shah, N. Single-and multisite planning and scheduling: current status and future challenges. *Proceedings of the Third International Conference of the Foundations of Computer-Aided Process Operations, Snowbird, Utah, USA, July 5−10,1998*; AIChE Symposium Series 94; AIChE: New York, 1998; p 75.

(3) Swaty, T. E. Consider over-the-fence product stream swapping to raise profitability. *Hydrocarbon Process.* **2002**, *81*, 37.

(4) Gonzalo, M. F.; Balseyro, I. G.; Bonnardot, J.; Morel, F.; Sarrazin, P. Consider integrating refining and petrochemical operations. *Hydrocarbon Process.* **2004**, *83*, 61.

(5) Sadhukhan, J.; Zhang, N.; Zhu, X. X. Analytical optimization of industrial systems and applications to refineries, petrochemicals. *Chem. Eng. Sci.* **2004**, *59*, 4169.

(6) Li, C.; He, X.; Chen, B.; Chen, B.; Gong, Z.; Quan, L. Integrative optimization of refining and petrochemical plants *Proceedings of the16th European Symposium on Computer Aided Process Engineering and 9th International Symposium Process Systems Engineering, Garmisch-Partenkirchen, Germany, July 9−13, 2006*; p 21.

(7) Kuo, T.; Chang, T. Optimal planning strategy for the supply chains of light aromatic compounds in petrochemical industries. *Comput. Chem. Eng.* **2008**, *32*, 1147.

(8) Al-Qahtani, K.; Elkamel, A. Multisite Facility Network Integration Design and Coordination: An Application to the Refining Industry. *Comput. Chem. Eng.* **2008**, *32*, 2198.

(9) Bell, J. T. Modeling of the Global Petrochemical Industry. Ph.D. Thesis, University of Wisconsin−Madison, 1990.

(10) Gary, J. H.; Handwerk, G. E. *Petroleum Refining: Technology and Economics*; Marcel Dekker Inc.: New York, 1994.

(11) Rudd, D. F.; Fathi-Afshar, S.; Trevino, A. A.; Statherr, M. A. *Petrochemical Technology Assessment*; John Wiley & Sons: New York, 1981.

(12) Balaraman, K. S. Maximizing Hydrocarbon Value Chain by Innovative Concepts. *J. Petrotech Soc.* **2006**, *3*, 26.

(13) Ren, T.; Patel, M.; Blok, K. Olefins from conventional and heavy feedstocks: Energy use in steam cracking and alternative processes. *Energy* **2006**, *31*, 425.

(14) Meyers, R. A. *Handbook of Petroleum Refining processes*, *2nd* ed.; McGraw-Hill: New York, 1997.

(15) Anonymous. Petrochem complex shields refining profits. *Oil Gas J.* **1998**, *96*, 31.

(16) Fujiyama, Y.; Redhwi, H. H.; Aitani, A. M.; Saeed, M. R.; Dean, C. F. Demonstration plant for new FCC technology yields increased propylene. *Oil Gas J.* **2005**, *103*, 54.

(17) Wang, J. China's Petrochemical Technologies. *Chem. Int.* **2006**, (May-June), 4.

(18) Philpot, J. Integration Profitability. *Hydrocarbon Eng.* **2007**, *12*, 41.

(19) Crawford, C. D.; Bharvani, R. R.; Chapel., D. G. Integrating Petrochemicals into the Refinery. *Hydrocarbon Eng.* **2002**, *7*, 35.

(20) Patel, N.; Baade, W.; Fong, L. W.; Khurana, V. Creating Value Through Refinery Hydrogen Management. http:// www. h2alliance.com /pdf/ Creating Value.pdf (accessed Feb 27, 2008).

(21) Al-Qahtani, K.; Elkamel, A.; Ponnambalam, K. Robust Optimization for Petrochemical Network Design under Uncertainty. *Ind. Eng. Chem. Res.* **2008**, *47*, 3912.

(22) Joly, M.; Moro, L. F. L.; Pinto, J. M. Planning and scheduling for petroleum refineries using mathematical programming. *Braz. J. Chem. Eng.* **2002**, *19*, 207.

(23) Lippe, D. Gasoline price spike shifts ethylene feeds. *Oil Gas J.* **2007**, *105*, 56.

(24) Lippe, D. US olefins see improving feed economics, demand. *Oil Gas J.* **2008**, *106*, 56.

(25) Singh, J.; Kumar, M. M.; Saxena, A. K.; Kumar, S. Reaction pathways and product yields in mild thermal cracking of vacuum residues: A multi-lump kinetic model. *Chem. Eng. J.* **2005**, *108*, 239.

(26) Van Geem, K. M.; Reyniers, M. F.; Marin, G. B. Challenges of Modeling Steam Cracking of Heavy Feedstocks. *Oil Gas Sci. Technol.* **2008**, *63*, 79.

(27) NBK MENA Equity Research. *Petrochemical Primer*; http:// cm2.zawya.com/researchreports/nbk/20070923_NBK_105729.pdf (accessed March 20, 2008).

(28) Brooke, A.; Kendrick, D.; Meeraus, A.; Raman, R. *GAMS−A User's Guide*; GAMS Development Corp.: Washington DC, 1996.

(29) CPLEX Optimization, Inc. *Using the CPLEX Callable Library and CPLEX Mixed Integer Library*; CPLEX Optimization, Inc.: Incline Village, NV, 1993.

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