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Energy configuration and operation optimization of refinery fuel gas networks

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HIGHLIGHTS

- Complementary formulations are introduced to scheduling model of fuel gas system.
- Physical constraints of the pipes are covered by incorporating detailed pipe model.
- Dynamic multi-component feature of fuel gas system is considered in the model.
- Both the heat value and pressure of demand points are calculated by dynamic model.
- More practical and preferable result is obtained by applying the proposed method.

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ABSTRACT

The production of regular clean fuels is faced with a problem of declining profit under more strict and costly environmental regulations. To satisfy the desire for higher profit and the firm requirements of environmental protection, it is imperative to improve the efficiency of energy systems within refineries. Over the past decade numerous attempts were made to enhance the energy system, addressing the steam power system and hydrogen system in particular. However, the fuel gas system, which serves as the dominant energy source of refineries, has drawn little attention in the research community. Industrial practices indicate that the energy efficiency of the fuel gas systems can be improved remarkably by optimizing the operation schedules. This paper presents a multi-period optimizing model for the scheduling of fuel gas system within refineries. Modeling of the pipeline system is considered important, which was usually ignored in the former studies. Flow reversal and flow transition in the pipe segments are taken into consideration. Pipelines with branching structure and loop structure can be easily modeled and solved with rational computation effort. Complementarity formulations are utilized in modeling of discrete decisions instead of the commonly used binary variables. Application of this method is illustrated with a case study.

1. Introduction

Energy has long been the focus of the research community [1–5]. Energy conservation and energy efficiency are recognized as major topics for the energy-security of US [4]. Additionally, it has also been pointed out that intimate connection exists between energy efficiency and the environmental impact [5]. In a word, it is of great importance to improve the energy efficiency of a system.

The oil refining industry is one of the most energy-intensive manufacturing industries. Energy is a major component of the refinery daily operating costs and is reported to be the second largest contributor to the total costs [6,7]. This energy is consumed by preheating of reactants, product separation or for onsite electricity generation. In refineries producing cleaner fuel, the energy consumption is even higher. Refineries are under pressure to produce cleaner fuels due to the trend in stricter environmental regulations on heavier crude oil processing. Various energy intensive units such as hydrogen production and hydro-treating units are under construction or being revamped to upgrade the quality of gasoline

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Abbreviations: CDU, crude distillation unit; DS, desulfurization unit; FCC, fluid catalytic cracking; HP, high pressure gas; LP, low pressure gas; LPG, liquefied petroleum gas; MPEC, mathematical program with equilibrium constraints; PX, para-xylene.

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Nomenclature

Sets

A	pipeline segments
B	optional backup fuel sources
D	fuel gas consumers
K	components of the gas flow
N	junctions that connect pipes
S	fuel gas suppliers
T	operating periods
$ArctoNode\ (A,N)$	set mapping of pipeline arcs A to node N
$ArcfromNode\ (A,N)$	set mapping of pipeline arcs A out of node N
$Demand\ (D, N)$	set mapping of demand nodes D from nodes N in the network
$Supply\ (S, N)$	set mapping of supply nodes S to nodes N in the network
$Z = \{in, out\}$	inlet/outlet of pipe segment

Parameters

$Area_a$	cross-sectional area in arc a, m^2
C	speed of sound in the gas
ρ	gas density
Dia_a	diameter of pipe segment a, m
L_a	length of pipe segment a, m
ξ	penalty cost of fuel gas, yuan/N m^3
ζ_b	unit price of the backup fuel b
$F^{Capacity}_{s,t}$	producing capacity of the production unit s in period t (based on volume N m^3/h)
Mw_k	molecular weight of component k
$H^{Dem}_{d,t}$	energy demand of consumer b at t, MJ
HV'_b	heating value of backup fuel b based on volume, MJ/N m^3
HV_s	heating value of fuel gas s based on mass, MJ/kg
$y_{s,k,t}$	mass fraction of component k of source s at t
t_t	operating period t
Nt	the last operating period
Δt	time duration of each operating period
μ	gas viscosity, kg/(m s)
$Temp$	temperature, K
P^{Low}	lower bound of the safety pressure level
P^{Up}	upper bound of the safety pressure level
R	gas constant
φ_a	surface roughness of pipe in arc a

Variables

$\frac{dP}{dx}_{a,z,t}$	partial derivative of pressure with respect to position in arc a at endpoint z at t
$F_{s,t}$	fuel gas flowrate of source s at the corresponding source node in period t based on volume, N m^3/h
$F_{b,t}$	consumption of backup fuel b in operating period t
$F_{b,d,t}$	flowrate of backup fuel from source b to consumer d in operating period t
$F_{d,t}$	volume flow of consumer d, N m^3/h
M	mass flowrate, kg/s
$M_{a,in,t}$	inlet mass flowrate of pipe arc a at t, kg/s
$M_{a,out,t}$	outlet mass flowrate of pipe arc a at t, kg/s
$M_{s,t}$	mass flowrate of fuel gas from source s to the corresponding source node at t, kg/s
$M_{d,t}$	mass flowrate of fuel gas fed to consumer d at t, kg/s
$Mw_{a,t}$	molecular weight of the gas stream in arc a at t
$H_{d,t}$	energy provided to consumer d at t, MJ
$HV_{n,t}$	heating value of the gas flow at node n at t based on mass, MJ/kg
$HV_{a,t}$	heating value of the gas flow in pipe arc a at t based on mass, MJ/kg
$HV^{in}_{a,t}$	heating value of the inlet gas flow in pipe arc a at t based on mass, MJ/kg
$HV'_{d,t}$	heating value of the feeding gas flow of consumer d at t based on volume, MJ/N m^3
$HV_{d,t}$	heating value of the feeding gas flow of consumer d at t based on mass, MJ/kg
$\overline{P}_{a,t}$	average pressure in pipe a at t, kPa
P	pressure
$P_{a,in,t}$	pressure at the inlet of pipe a at t
$P_{a,out,t}$	pressure at the outlet of pipe a at t
$Re_{a,z,t}$	Reynolds number for gas in arc a at endpoint z at t
$f_{a,z,t}$	friction factor for gas in arc a at endpoint z at t
$f^{lam}_{a,z,t}$	laminar friction factor for gas in arc a at endpoint z at t
$f^{turb}_{i,z,t}$	turbulent friction factor for gas in arc a at endpoint z at t
$y_{a,k,t}$	mole fraction of component k of the flow in arc a at t
$y^{in}_{a,k,t}$	mole fraction of component k of the inlet flow of arc a at t
$y_{d,k,t}$	mole fraction of component k of the inlet of consumer d at t
$y_{n,k,t}$	mole fraction of component k of the gas flow at node n at t
$TInv_t$	total gas inventory of the pipeline system at t, kg
$Inv_{a,t}$	gas inventory of pipe segment a at t, kg
$TInv_{Nt}$	total gas inventory of the pipeline system at end of the scheduling time horizon, kg
$switch_{a,z,t}$	switching variable
$\lambda 1_{a,z,t}, \lambda 2_{a,z,t}$	auxiliary variables

and diesel products in order to meet the environmental requirement. As the energy consumption is increasing, the overall profit is shrinking. Consequently, refineries are facing challenges of satisfying both the profit demand and environment requirement. Energy conservation and energy efficiency are the keys to solve this challenge.

The energy system of the refineries consists of the following sub-systems: hydrogen, fuel gas and steam system [8,9]. The fuel gas sub-system is the greatest contributor to the total energy system. However, in compare to the well-studied hydrogen system [10–18] and steam system [19–23], only a few work has been focused on optimization of the fuel gas system. The researches cover systems design/revamping and decision making of the fuel gas system. Hasan et al. [24] developed a nonlinear program to tackle the synthesis of optimal fuel gas networks. It was reported that 40–50% of the total energy cost can be saved by applying the proposed method. Many realistic features were incorporated in their model such as auxiliary equipment, nonisobaric operation and nonisothermal mixing, but it was only valid for steady-state operation. Later, Jagannath et al. [25] adapted the model to handle dynamic operation system. In their extended model, a multi-period two stage programming model was reported. Network design including decisions regarding to the existence and sizes of equipments was determined in the first-stage, while the network operation details such as flows and operation duties were calculated in the second-stage. In case of the existing fuel gas systems the increase of efficiency should be first addressed through full utilization of the existing devices. Scheduling is one of the important methods to make competitive operation decisions. Zhang and Rong [26] developed a mixed integer linear programming model for fuel gas scheduling. In their model, the storing capability of the system and the consumption of fuel gas in cogeneration and production

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system were considered ensuring site-wide fuel gas balance. Later, a fuzzy possibilistic model [27] was proposed to deal with the imprecise nature of the system. Subsequently, a logical modeling method, the generalized disjunctive programming, was introduced [28] for the scheduling of fuel gas system considering the pipeline network with loop structure. However, the proposed iterative calculation procedure addressed only simple pipeline systems. Practical circumstance such as pressure gradient along the pipe, flow transition and flow reversal were not considered, which could lead to suboptimal or even infeasible result.

The pipeline system, asserting delivery, plays a key role in daily production. Physical constraints of the pipeline system are important for obtaining practical results. Modeling of the gas pipeline systems has been well investigated by researchers. Wong and Larson [29] developed a dynamic programming technique to deal with the optimization problem of natural gas pipeline system. Their study was dealing with a single gas pipeline controlled by a single compressor. Later, a mathematical model for the dynamic simulation of gas pipeline network was presented [30] and a numerical method for its solution was also developed [31,32]. However this model was limited to turbulent flows and fixed flow direction. Later, van den Heever and Grossmann [33] incorporated a detailed pipeline model into the hydrogen supply network optimization model taking flow transition and flow reversal into consideration. Discrete variables were employed in their model and a mixed integer nonlinear programming model was formulated. Baumrucker et al. [34] introduced the complementarity constraints into modeling of nonsmooth elements, which found its application in the gas pipeline modeling, also. Nonetheless, these models consider only gas systems with constant components or systems where slight change of the components will not affect the feasibility of the result, for example natural gas system. However, the compositions of the fuel gas streams in refinery may change, which will lead to variation of the heating values. Heating value is an important property for fuel gas utilization. Therefore, the pipeline model needs to be more comprehensive to be able to provide useful solutions for such systems.

We address this issue by extending the pipeline model to manage multi-component systems where changes in the components affect the practicability of the result. The extended model is incorporated into the multi-period scheduling model of the fuel gas system in refinery. Flow transition and flow reversal are fully considered which makes it easier to model pipeline network with loop structure. Complementarity formulations are employed to model discrete events, thus binary variables are avoided. This paper is arranged as follows: Section 2 sketches out the fuel gas system in refinery. The fuel gas scheduling problem is stated in Section 3, followed by the description of mathematical formulations in Section 4. Section 5 presents a case study to demonstrate the efficiency of the proposed method. Result and discussion are given in Section 6, and conclusions are summarized in Section 7.

2. Fuel gas system in refinery

The fuel gas system in refinery consists of two sections: low pressure and high pressure section, as shown in Fig. 1. The fuel gas producers include the reforming unit, the FCC unit as well as the gas separation units of the hydrogen system which produce hydrogen-rich waste gases [35]. Purge gases with high pressure are sent to the high pressure fuel gas system, while those with low pressure are sent to the low pressure fuel gas system.

The fuel gas from low pressure system cannot be directly reused as fuel due to its high sulfur content. It can be used for hydrocarbon (C_3 and C_4 olefins) and hydrogen recovery. It can also be used to supplement the high pressure fuel gas system after been compressed to a higher pressure and desulfurized. The rest is stored in a liquid sealed gas tank. Gas inventory in the high pressure system must be kept under a certain safety level. When there's too much fuel gas in the high pressure system, it will be transferred to the low pressure system for safety reasons. Similarly, when the inventory of the storage tank in the low pressure system exceeds its capacity, the gas will be released to flare. However, this should be avoided if possible as sending fuel gas from the high pressure system to the low pressure system will contribute to increase of the operating cost, while flaring to the air will result in resource waste and air pollution.

Usually, the internal fuel gas cannot meet all the energy demand within the system. Thus backup fuel resources are required, such as natural gas. Sometimes, the refinery products are also used for the inner energy supply, for example fuel oil.

3. Problem definition

The scheduling problem of the fuel gas system in refinery is given as a set of fuel gas producers $S = \{s/s = 1, 2, \ldots, Ns\}$ including high pressure fuel gas producers and desulfurization units in which the low pressure fuel gas is desulfurized, and a set of energy consumers $D = \{d/d = 1, 2, \ldots, Nd\}$ including boilers and process units, connected by a set of pipe segments $A = \{a/a = 1, 2, \ldots, Na\}$. There are also backup fuel sources $B = \{b/b = 1, 2, \ldots, Nb\}$, such as natural gas and fuel oil, in case the energy provided by the fuel gas cannot meet the system demand. Different fuel sources have different heating values HV based on their composition. As the operation

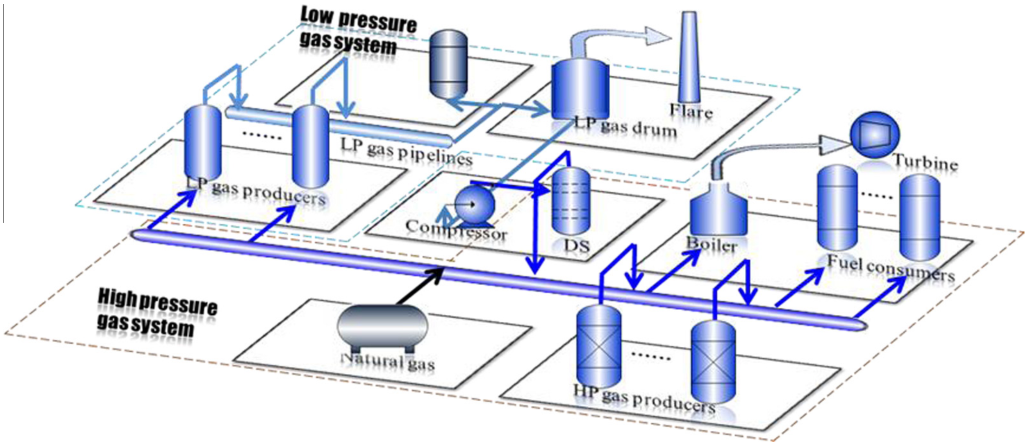

Fig. 1. Sketch map of the fuel gas system in refinery.

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conditions of the process units are changing in accordance with the operation plan in real practice, the scheduling time horizon is divided into several operating periods $T = \{t/t = 1, 2, \ldots, Nt\}$. Within each operating period, each process unit is kept in steady state operation. For different operating periods, the fuel gas production rate $F_{s,t}$ of each producer and the energy demand $H_{d,t}^{Dem}$ of each consumer are different. In order to maintain economical and satisfactory operation, the fuel gas produced onsite should be fully utilized and the demand of the consumers must be satisfied. Furthermore, the physical constraints of the pipeline network should be considered and the pressure of the system must be kept under the safety level. It is desired to build up a scheduling model enabling fuel gas distribution in accordance with the energy supply and demand in the system while targeting the minimal operating cost.

4. Mathematical model

A multi-period mathematical model for the scheduling of fuel gas system in refinery is presented in this section.

4.1. Objective function

The goal of the scheduling problem is to reduce the energy waste and minimize the cost while maintain satisfactory performance of the system. The objective function consists of two parts. The first part is the fuel gas penalty cost. The second part is the operating cost, which consists of the purchase cost of the backup fuels and the profit shrink caused by the consumption of onsite fuel product.

$$\text{Min } obj = \sum_d \sum_t \xi F_{d,t} + \sum_b \sum_t \zeta_b F_{b,t} \tag{1}$$

where $F_{d,t}$ represents the fuel gas feeding volume flowrate of consumer d, N m^3/h. ξ stands for the unit penalty of fuel gas. ζ_b represents the unit price of backup fuel b, and $F_{b,t}$ stands for the consumption of backup fuel b in operation period t. It should be stressed out that the first part of the objective function is used to minimize the fuel gas consumption, in order to improve the energy efficiency. The penalty cost will not count in the total network cost though it will affect the total cost.

$$F_{b,t} = \sum_d F_{b,d,t} \quad \forall b \in B, t \in T \tag{2}$$

4.2. Constraints for the fuel gas producers

The gas produced by the onsite processing units should be efficiently utilized since it is a byproduct. What is more, releasing the fuel gas into the atmosphere will lead to severe environmental pollution. Effective utilization of the fuel gas can decrease the expenditure for the backup fuels, and reduce the environmental impact of the plant.

$$F_{s,t} \le F_{s,t}^{Capacity} \quad \forall s \in S, t \in T \tag{3}$$

where $F_{s,t}$ represents the fuel gas available from unit s in period t, N m^3/h. For a high pressure fuel gas producer, it equals to the production amount in each period. For a low pressure fuel gas producer, it equals to the amount that being desulfurized and sent to high pressure fuel gas system.

4.3. Constraints for fuel gas consumers

To maintain productive operation, the energy demand of each consumer must be satisfied.

$$H_{d,t} \ge H_{d,t}^{Dem} \quad \forall d \in D, t \in T \tag{4}$$

The energy acquired by each consumer equals to the energy provided by the fuel gas feeding flow plus the energy from the possible backup fuels.

$$H_{d,t} = F_{d,t} HV'_{d,t} + \sum_b F_{b,d,t} HV'_b \quad \forall d \in D, t \in T \tag{5}$$

4.4. Pipeline network constraints

Pipelines are key elements of the production. They connect the material suppliers with the consumers and transport the materials from upstream to downstream units.

In practice, the pipeline network is composed from a large number of pipe segments. Pipe segments are connected by physical junctions that we present as nodes $n \in N$, as shown in Fig. 2. A node works like a splitter or a mixer, thus there's no mass accumulation. The inlet flowrate equals to the outlet flowrate. The mass balance for node n is formulated as:

$$\sum_{a:(a,n) \in ArctoNode(A,N)} M_{a,out,t} + \sum_{s:(s,n) \in Supply(S,N)} M_{s,t} = \sum_{a:(a,n) \in ArcfromNode(A,N)} M_{a,in,t} + \sum_{d:(d,n) \in Demand(D,N)} M_{d,t} \quad \forall n \in N \tag{6}$$

$$\sum_{a:(a,n) \in ArctoNode(A,N)} M_{a,out,t} y_{a,k,t} + \sum_{s:(s,n) \in Supply(S,N)} M_{s,t} y_{s,k,t} = \sum_{a:(a,n) \in ArcfromNode(A,N)} M_{a,in,t} y_{a,k,t} + \sum_{d:(d,n) \in Demand(D,N)} M_{d,t} y_{d,k,t} \quad \forall n \in N, k \in K \tag{7}$$

where the left side represents the inlet mass flow of node n, while the right side gives the outlet. It should be stressed out that the proposed pipeline model allows reversal flow. When the flow direction is changed in a pipe segment, negative sign will be assigned to the flowrate indicating the flow direction. In this case, the inlet flow became the actual outlet.

The component concentration at each point of the pipe system is calculated as below:

$$y_{n,k,t} = y_{s,k,t} \quad \forall (s,n) \in Supply(S,N), \;\; k \in K, \;\; t \in T \tag{8}$$

$$y_{n,k,t} = \begin{cases} y_{a,k,t}^{in} & \text{if} \quad M_{a,in,t} > 0 \quad \forall (a,n) \in ArcfromNode(A,N), t \in T \\ y_{a,k,t}^{in} & \text{if} \quad M_{a,out,t} < 0 \quad \forall (a,n) \in ArctoNode(A,N), t \in T \end{cases} \tag{9}$$

$$y_{d,k,t} = y_{n,k,t} \quad \forall (d,n) \in Demand(D,N), k \in K, t \in T \tag{10}$$

$$\sum_k y_{n,k,t} = 1 \quad \forall n \in N, t \in T \tag{11}$$

Eq. (8) assigns the component concentration of the source to the corresponding source node. Eq. (9) defines the inlet flow concentration of a pipe arc, which equals to the node concentration of the inlet direction. Eq. (10) gives the component concentration of the demand node.

The heating value of the fuel gas flow is determined by its components. When different fuel gas flows are mixed together, the heating value of the new stream is calculated as follows:

$$\sum_{a:(a,n) \in ArctoNode(A,N)} M_{a,out,t} HV_{a,t} + \sum_{s:(s,n) \in Supply(S,N)} M_{s,t} HV_s = \sum_{a:(a,n) \in ArcfromNode(A,N)} M_{a,in,t} HV_{a,t} + \sum_{d:(d,n) \in Demand(D,N)} M_{d,t} HV_{d,t} \quad \forall n \in N \tag{12}$$

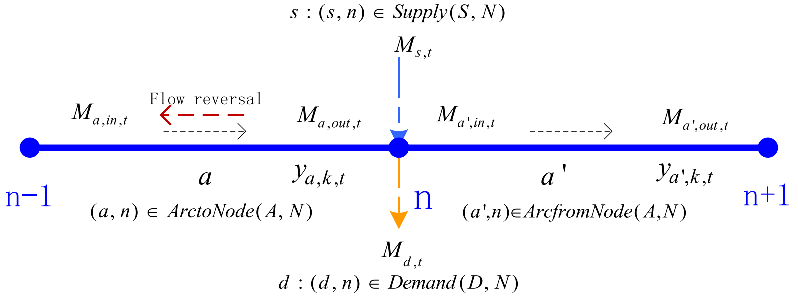

Fig. 2. Representation of pipe connection.

$$HV_{n,t}=\begin{cases}HV^{in}_{a,t} & if \quad M_{a,in,t}>0 \quad \forall(a,n)\in ArcfromNode(A,N),t\in T\\ HV^{in}_{a,t} & if \quad M_{a,out,t}<0 \quad \forall(a,n)\in ArctoNode(A,N),t\in T\end{cases} \tag{13}$$

$$HV_{d,t}=HV_{n,t} \quad \forall(d,n)\in Demand(D,N),t\in T \tag{14}$$

$$HV_{s}=HV_{n,t} \quad \forall(s,n)\in Supply(S,N),t\in T \tag{15}$$

Assuming that, (1) the flow is one-dimensional plug flow and (2) isothermal conditions and ideal gas behavior. Given that the fuel gas system is normally kept under 0.7–0.8 MPa, the ideal gas behavior assumption is rational. The mass balance for the fuel gas flow inside a pipe segment is described as a linear first order partial differential equation [31].

$$\frac{Area}{c^2}\frac{\partial P}{\partial t}=-\frac{\partial M}{\partial x}$$

where c is the speed of sound in the gas and $c^2=\frac{P}{\rho}$ (the isothermal process, ρ—gas density). The following formulation is derived:

$$\frac{MwArea}{RTemp}\frac{\partial P}{\partial t}=-\frac{\partial M}{\partial x}$$

By integrating both sides of the equation and applying the trapezoidal rule, the mass balance equation is reformulated to the following form:

$$\frac{Area_aL_a(Mw_{a,t+1}\overline{P}_{a,t+1}-Mw_{a,t}\overline{P}_{a,t})}{RTemp}=\frac{1}{2}\left[(M_{a,in,t}-M_{a,out,t})\right.$$
$$\left.+(M_{a,in,t+1}-M_{a,out,t+1})\right](t_{t+1}-t_t) \quad \forall a\in A,t\in T \tag{16}$$

where $Area_a$ and L_a stand for the cross sectional area and the length of the pipe segment, respectively. $Mw_{a,t}$ is the molecular weight of the fuel gas stream in pipe segment a at t. The molecular weight of the fuel gas stream is changing along with time and the location.

$$Mw_{a,t}=\frac{1}{\sum_k^{y_{a,k,t}}/Mw_k} \quad \forall a\in A,t\in T \tag{17}$$

Mass balance for the whole pipeline system is formulated as:

$$\frac{\sum_a Area_aL_a(Mw_{a,t+1}\overline{P}_{a,t+1}-Mw_{a,t}\overline{P}_{a,t})}{RTemp}$$
$$=\frac{1}{2}\left[\sum_s(M_{s,t}+M_{s,t+1})-\sum_d(M_{d,t}+M_{d,t+1})\right](t_{t+1}-t_t)$$
$$\forall t\in T,t\neq Nt \tag{18}$$

where $\overline{P}$ presents the average pressure of a pipe and is calculated by the following equations [36]:

$$\overline{P}_{a,t}=\frac{\int_0^{L_a}P(x')dx'}{\int_0^{L_a}dx}=\frac{\left(\frac{1}{2}\right)P_{a,in,t}+\left(\frac{1}{12}\right)\frac{dP}{dx_{a,in,t}}L_a+\left(\frac{1}{2}\right)P_{a,out,t}-\left(\frac{1}{12}\right)\frac{dP}{dx_{a,out,t}}L_a}{L_a} \tag{19}$$

$$\forall a\in A,t\in T$$

$$P_{a,out,t}=P_{a,in,t}+\frac{L_a}{2}\left(\frac{dP}{dx_{a,in,t}}+\frac{dP}{dx_{a,out,t}}\right) \quad \forall a\in A,t\in T \tag{20}$$

The momentum conservation for one-dimensional plug flow is described as a nonlinear differential equation. To compromise between the model accuracy and the computational cost, the following simplified equation is derived.

$$\frac{dP}{dx}=\frac{-fRTempM|M|}{2DiaArea^2MwP}$$

The equation is evaluated at both ends of each pipe segment, and $Z=\{in, out\}$ is introduced for notational convenience.

$$\frac{dP}{dx_{a,z,t}}=\frac{-f_{a,z,t}RTempM_{a,z,t}|M_{a,z,t}|}{2Dia_aArea_a^2Mw_{a,t}P_{a,z,t}} \quad \forall a\in A,t\in T,z\in Z \tag{21}$$

The momentum conservation equation gives the pressure gradient of the fuel gas along a pipe. Here, the friction dominates the other contribution factors. Both laminar and turbulent flows are considered.

For laminar flow:

$$f^{lam}_{a,z,t}=64/\mathrm{Re}_{a,z,t} \quad \forall a\in A,z\in Z,t\in T \tag{22}$$

For turbulent flow:

$$f^{turb}_{a,z,t}=1.326\left[\ln\left(\frac{1}{\varphi_a/(3.7Dia_a)+2.51\Big/\left(Re_{a,z,t}\sqrt{f^{turb}_{a,z,t}}\right)}\right)\right]^{-2} \tag{23}$$

$$\forall a\in A,z\in Z,t\in T$$

The Reynolds number is calculated as:

$$Re_{a,z,t}=\frac{|M_{a,z,t}|Dia_a}{\mu Area_a} \quad \forall a\in A,z\in Z,t\in T \tag{24}$$

where μ represents for the gas viscosity at the given condition, kg/(m s).

For the long distance pipeline operation, the spread of the transient behavior can be relatively slow, such as natural gas system. The inlet and outlet flow direction of a pipe may be different when the inlet flow/pressure experiences a sudden change. However, for the fuel gas system within refineries, the pipe segments are relatively short compare to that of the natural gas system. The transient behavior can be completed within several minutes, which is quite short comparing to the scheduling time interval. Hence we assume that the inlet and outlet flow direction are the same during each period.

$$M_{a,in,t}M_{a,out,t}\geq 0 \quad \forall a\in A,t\in T \tag{25}$$

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It is assumed that the fuel gas is uniformly mixed in each pipe segment. The component concentration and the heating value of the gas flow inside a pipe are defined as:

when $t = 0$

$$y_{a,k,0} = y^{in}_{a,k,0} \quad \forall a \in A, k \in K \tag{26}$$

$$HV_{a,0} = HV^{in}_{a,0} \quad \forall a \in A \tag{27}$$

when $t > 0$, $M_{a,in,t+1} > 0$

$$y_{a,k,t+1}(|M_{a,in,t+1}|\Delta t + Inv_{a,t}) = y^{in}_{a,k,t+1}|M_{a,in,t+1}|\Delta t + y_{a,k,t}Inv_{a,t} \quad \forall a \in A, k \in K, t \in T \tag{28}$$

$$HV_{a,t+1}(|M_{a,in,t+1}|\Delta t + Inv_{a,t}) = HV^{in}_{a,t+1}|M_{a,in,t+1}|\Delta t + HV_{a,t}Inv_{a,t} \quad \forall a \in A, t \in T \tag{29}$$

when $t > 0$, $M_{a,in,t+1} < 0$

$$y_{a,k,t+1}(|M_{a,out,t+1}|\Delta t + Inv_{a,t}) = y^{in}_{a,k,t+1}|M_{a,out,t+1}|\Delta t + y_{a,k,t}Inv_{a,t} \quad \forall a \in A, k \in K, t \in T \tag{30}$$

$$HV_{a,t+1}(|M_{a,out,t+1}|\Delta t + Inv_{a,t}) = HV^{in}_{a,t+1}|M_{a,out,t+1}|\Delta t + HV_{a,t}Inv_{a,t} \quad \forall a \in A, t \in T \tag{31}$$

$$\sum_k y^{in}_{a,k,t} = 1 \quad \forall a \in A, t \in T \tag{32}$$

$$\sum_k y_{a,k,t} = 1 \quad \forall a \in A, t \in T \tag{33}$$

The network inventory considered in this study is the linepack in the pipeline.

$$Inv_{a,t} = Area_a L_a Mw_{a,t}\overline{P}_{a,t}/RTemp \quad \forall a \in A, t \in T \tag{34}$$

$$Inv_{a,t+1} = \begin{cases} Inv_{a,t} + (M_{a,in,t+1} - M_{a,out,t+1})\Delta t & \text{if} \quad M_{a,in,t+1} > 0 \\ Inv_{a,t} - (M_{a,in,t+1} - M_{a,out,t+1})\Delta t & \text{if} \quad M_{a,in,t+1} < 0 \end{cases} \quad \forall a \in A, t \in T \tag{35}$$

It is necessary to ensure that the gas inventory at the end of the scheduling time horizon equal to or not less than the original amount of the gas inventory. The following equations are formulated:

$$TInv_t = \sum_{a \in A} Inv_{a,t} \quad \forall t \in T \tag{36}$$

$$TInv_{Nt} \geq TInv_0 \tag{37}$$

4.5. Complementarity formulations and equilibrium constraints

Both the piecewise function and the absolute value operator are presented in the mathematical model. These are nonsmooth functions. Usually, mixed integer programming is used to handle these logical disjunctions. But the associate computational cost might be high for large systems with lots of discrete events. Especially for the dynamic system in the present work, flow reversal may occur at any point of the pipe system in any time. Additionally, nonlinear terms exist in the model. For the worst case, the solution time of a mixed integer nonlinear programming problem may grow exponentially with the number of discrete decisions. Zhang et al. [28] introduced a general disjunctive programming (GDP) to model the logical disjunctions, which is eventually converted to mixed integer form and solved by a simulation based iterative approach. In the present work, we adopt an alternative way to model the disjunctive events—complementary formulations.

Complementarity is a relationship between variables where either or both must be at their bound. Complementarity formulations offer an alternative way to model discrete decisions without the use of binary variables. Usually, compared to a mixed integer nonlinear program, less computational effort is required to obtain local optimal solutions. It is useful with certain classes of problems [34], such as absolute value operator and piecewise functions.

The absolute value operator in the momentum conservation equation is reformulated as:

$$|M_{a,z,t}| = M^+_{a,z,t} + M^-_{a,z,t} \quad \forall a \in A, z \in Z, t \in T \tag{38}$$

$$M_{a,z,t} = M^+_{a,z,t} - M^-_{a,z,t} \quad \forall a \in A, z \in Z, t \in T \tag{39}$$

$$0 \leqslant M^+_{a,z,t} \perp M^-_{a,z,t} \geqslant 0 \quad \forall a \in A, \ z \in Z, \ t \in T \tag{C.1}$$

Eq. (C.1) represents a complementarity relationship, in which $\perp$ is the complementarity operator compelling at least one of the complementing bounds to be active. The complementarity formulations are then reformulated into equilibrium constraints in order to be incorporated into the mathematical program, thus a mathematical program with equilibrium constraints (MPEC) is created. The applications of MPEC in chemical engineering optimization is introduced by Baumrucker et al. [34].

$$M^+_{a,z,t}M^-_{a,z,t} = 0 \quad \forall a \in A, z \in Z, t \in T \tag{40a}$$

$$M^+_{a,z,t}, M^-_{a,z,t} \geq 0 \quad \forall a \in A, z \in Z, t \in T \tag{40b}$$

The friction factor is calculated piecewisely based on the flow pattern:

$$f_{a,z,t} = \begin{cases} f^{lam}_{a,z,t} & \text{Re} < 2300 \\ f^{turb}_{a,z,t} & \text{Re} > 2300 \end{cases} \quad \forall a \in A, z \in Z, t \in T \tag{41}$$

With the use of complementarity formulations, the piecewise function is reformulated as:

$$f_{a,z,t} = switch_{a,z,t}f^{lam}_{a,z,t} + (1 - switch_{a,z,t})f^{turb}_{a,z,t} \quad \forall a \in A, z \in Z, t \in T \tag{41a}$$

$$(Re_{a,z,t} - 2300) - \lambda 1_{a,z,t} + \lambda 2_{a,z,t} = 0 \quad \forall a \in A, z \in Z, t \in T \tag{41b}$$

$$0 \leq switch_{a,z,t} \perp \lambda 1_{a,z,t} \geq 0 \quad \forall a \in A, z \in Z, t \in T \tag{C.2}$$

$$0 \leq (1 - switch_{a,z,t}) \perp \lambda 2_{a,z,t} \geq 0 \quad \forall a \in A, z \in Z, t \in T \tag{C.3}$$

where $switch_{a,z,t}$ is the switching variable determined by the Reynolds number. By inspection it is easy to see that if the Reynolds number is smaller than 2300, auxiliary variable $\lambda 2_{a,z,t}$ will be enforced to be positive by Eq. (41b), and the switching variable $switch_{a,z,t}$ will then be set to 1 by the complementarity formulation (C.3). In this way, the friction factor is calculated by laminar flow, vice versa. The friction factor calculation is set valued when the Reynolds number is exactly 2300. The complementary formulations (C.2) and (C.3) are then reformulated as equilibrium constraints.

$$switch_{a,z,t}\lambda 1_{a,z,t} = 0 \quad \forall a \in A, z \in Z, t \in T \tag{41c}$$

$$(1 - switch_{a,z,t})\lambda 2_{a,z,t} = 0 \quad \forall a \in A, z \in Z, t \in T \tag{41d}$$

$$switch_{a,z,t}, \lambda 1_{a,z,t}, \lambda 2_{a,z,t} \geq 0 \quad \forall a \in A, z \in Z, t \in T \tag{41e}$$

$$1 - switch_{a,z,t} \geq 0 \quad \forall a \in A, z \in Z, t \in T \tag{41f}$$

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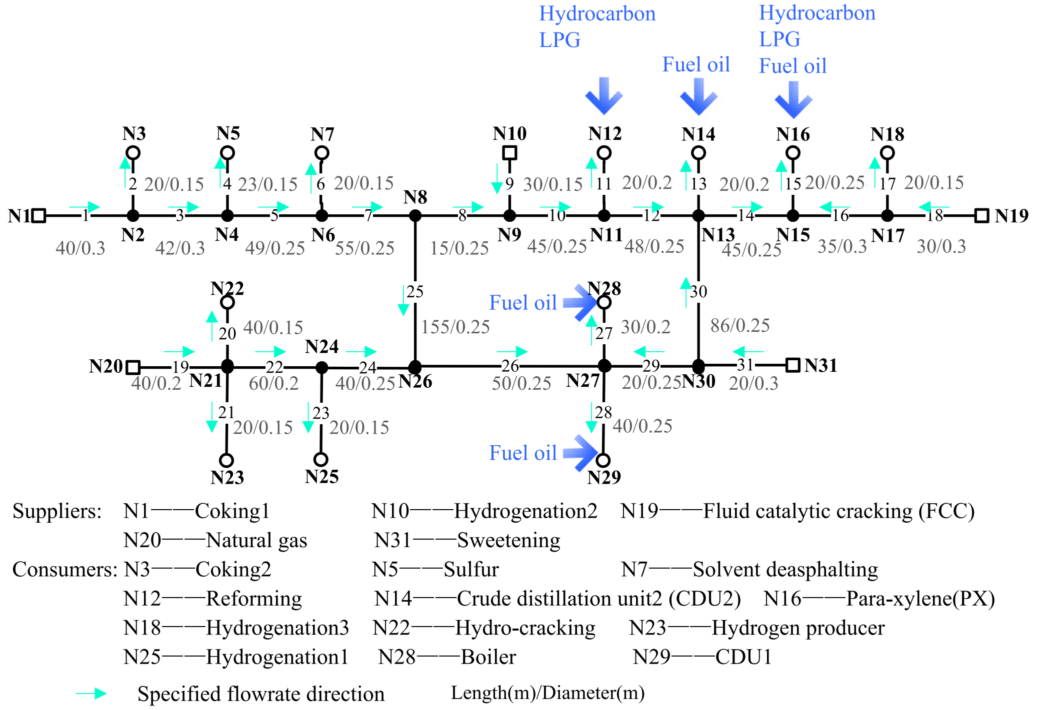

Fig. 3. Layout of the pipeline system of the case study. Nodes are labeled N#, with source nodes (suppliers) represented as small boxes, and demand nodes (customers) marked as small round boxes.

Table 1
Prediction of fuel gas output for the in-plant producers (N m^3/h).

s	Operation periods					
	1	2	3	4	5	6
Coking	10,600	11,800	11,900	11,300	11,100	11,400
Hydrogenation2	1200	1200	1250	1210	1190	1200
FCC	7100	7080	6900	7000	7090	6060
Sweetening	10,000	10,500	9800	10,200	10,800	10,500
Low-pressure gas	10,800	10,600	17,000	16,000	13,700	10,500

Table 2
Heating value (based on volume) of fuel gas and backup fuels (MJ/N m^3).

HV'	Coking	Hydrogenation2	FCC	Sweetening
	33.85	65.049	30.157	39.436
	Natural gas	Hydrocarbon	LPG	Fuel oil (MJ/t)
HV'	33.324	103.928	125.738	40,000

Table 3
Component concentration and viscosity of each fuel gas source.

	Coking	Hydrogenation2	FCC	Sweetening	Natural gas
H_2	0.1151	0.1303	0.311	0.2503	0
O_2	0.0164	0.037128	0.008	0.009324	0
N_2	0.0064	0.139672	0.1308	0.035076	0.0554
CH_4	0.5004	0.312	0.2593	0.4867	0.8553
C_2H_4	0.0329	0.1237	0.1237	0.0196	0
C_2H_6	0.148	0.1789	0.115	0.1653	0.00025
C_3H_8	0.0853	0.0014	0.0141	0.0039	0.0002
C_3H_6	0.0328	0.0043	0	0.0031	0
C_4H_8	0.0072	0.0108	0.0045	0.0011	0
C_4H_{10}	0.0205	0.0101	0	0.0026	0
C_5H_{12}	0.0164	0.004	0	0.004	0
CO	0	0.0268	0.0056	0.012	0.04995
CO_2	0.0186	0.0209	0.028	0.007	0.0389
μ	1.10×10^{-5}	1.18×10^{-5}	1.08×10^{-5}	1.08×10^{-5}	1.18×10^{-5}

4.6. Pressure constraints

The fuel gas system is required to be maintained under the safety pressure range.

$$P^{Low} \le P_{a,z,t} \le P^{Up} \quad \forall a \in A, z \in Z, t \in t \tag{42}$$

As ideal gas behavior is assumed, the relationship between volumetric flowrate and mass flowrate is give as follow:

$$M_{s,t} = F_{s,t} Mw_{s,t} P / RTemp \tag{43}$$

5. Case study

The application of the proposed approach is illustrated with a case study taken from the literature [28]. The fuel gas system in the case study consists from 3 high pressure fuel gas producers, a sweetening unit which upgrades low pressure fuel gas to qualified high pressure fuel gas, 4 backup fuel sources, 1 self-producing-self-consuming unit and 12 energy consumers. The layout of the pipe-

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Table 4
Price of the backup fuels (yuan/N m^3).

	Natural gas	Hydrocarbon	LPG	Fuel oil (yuan/t)	HP to LP	Flare
ζ	2	7	10	2000	5	6

HP—high pressure gas, LP—low pressure gas.

Table 5
Prediction of energy demanded by the consumers in the case study (KMJ).

Consumers	Operating periods					
	1	2	3	4	5	6
Coking2	77	79	78	74	78	77
Sulfur	67	65	66	67	69	66
Solvent deasphalting	50	50	51	48	52	51
Reforming	330	335	331	329	334	338
CDU2	290	293	291	288	289	290
PX	680	684	682	679	677	681
Hydrogenation3	38	37	39	40	37	39
Hydro-cracking	56	56.5	57	56	57.5	56
Hydrogen producer	56	56.5	57	56	57.5	56
Hydrogenation1	39	37	38	39	40	38
Boiler	146.4	145.5	146.8	145.9	146.4	147.2
CDU1	198.6	197.5	199.2	198.1	198.6	199.8
Hydrogenation2	65	66	68	64	65	66

line system as well as the length and diameter of the pipelines are given in Fig. 3.

The scheduling problem considered in this work is composed of 6 operation periods. The temperature of the system is set to be at 298.15 K. Table 1 gives the prediction of the fuel gas output for the producers, while Table 2 presents the heating values of the energy sources. The heating values of the energy sources are considered to be constant during the scheduling time horizon. The component concentration and the viscosity of each energy source are given in Table 3. In order to cut down the calculation effort, the viscosity of the fuel gas is set to be 1.18×10^{-5}. Backup fuels include natural gas, hydrocarbon and fuel oil. Prices of the backup fuels are shown in Table 4. The penalty cost of fuel gas ξ is set to be 1. Prediction of the energy demanded by the consumers is shown in Table 5. In order to maintain safety operation, the pressure of the system is required to be kept within the range of (0.7, 0.8) MPa.

Table 6
Comparison of the proposed model and the model proposed by Zhang et al. [28]

	Model proposed by Zhang et al. [28]	Proposed model
Model type	MILP	MPEC
Continuous variables	384	2618
Binary variables	162	–
Constraints	720	2656
Practical considerations	Loop and branching pipeline network	Loop and branching pipeline network; flow reversal; flow transition; dynamic heating value and pressure calculation
Solve strategy	Simulation-based iterative approach	Non-iterative method

It can be concluded from Tables 1, 2 and 5 that the energy produced by the self-producing-self-consuming unit (i.e. the hydrogenation2 unit) is larger than its demand during the whole time horizon. Hence it is treated as an energy supplier, as shown in Fig. 3.

6. Result and discussion

GAMS (General Algebraic Modeling System) software [37] is used to model the scheduling problem and NLPEC is chosen as the solver. The MPEC model contains 2618 variables and 2656 constraints. A comparison of the proposed model and the model proposed by Zhang et al. [28] is given in Table. 6. The MPEC model is transformed into NLP model by the NLPEC solver and then solved with a standard NLP solver. In our case study, we apply the active set NLP solver, CONOPT. It took 275.581 s to get the optimal result (based on the PC specification: Intel D CPU 3.00 GHz, 4 GB RAM). The model is first solved as a steady state problem under the operation condition of the starting period. The solution from the first solving is then used to initiate the variables of the multi-period model.

By solving the multi-period model, an optimal scheduling result including a rational distribution of the high pressure fuel gas and the corresponding supplementation strategy of the backup fuels is obtained. Table 7 gives the result status of the input fuel gas flow of each consumer in each period, including the flowrate and the heating value. The supplementation scheme of the backup fuels is shown in Table 8. Only three of the backup fuels are chosen by the proposed procedure to complement the energy deficit of the system, which are natural gas, fuel oil and liquefied petroleum gas (LPG). It should be noted that, natural gas is first sent to the high pressure fuel gas pipeline system and then to each consumer, while the other two backup fuels are directly fed to the corresponding consumers.

A comparison is made to illustrate the advantage of the proposed method. Fig. 4 compares the natural gas consumption

Table 7
Result status of the input fuel gas flow of each consumer.

D	*t*											
	1		2		3		4		5		6	
	F	HV'	*F*	HV'	*F*	HV'	*F*	HV'	*F*	HV'	*F*	HV'
Coking2	2274.7	33.85	2333.8	33.85	2304.3	33.85	2186.1	33.85	2304.3	33.85	2274.7	33.85
Sulfur	1979.3	33.85	1920.2	33.85	1949.8	33.85	1979.3	33.85	2038.4	33.85	1949.8	33.85
Solvent deasphalting	1477.1	33.85	1477.1	33.85	1506.6	33.85	1418.0	33.85	1536.2	33.85	1506.6	33.85
Reforming	5062.6	35.04	6243.9	34.74	6345.2	34.82	5922.5	35.00	5402.0	34.91	5837.2	34.81
CDU2	3016.5	39.43	1805.8	39.42	3925.3	39.43	1472.3	39.41	3775.3	39.43	1707.3	39.42
PX	5844.8	30.16	5851.6	30.16	5625.2	30.16	5661.1	30.16	5888.5	30.16	4749.3	30.17
Hydrogenation3	1260.1	30.16	1226.9	30.16	1293.2	30.16	1326.4	30.16	1226.9	30.16	1293.2	30.16
Hydro-cracking	1680.5	33.32	1695.5	33.32	1710.5	33.32	1680.5	33.32	1725.5	33.32	1680.5	33.32
Hydrogen producer	1680.5	33.32	1695.5	33.32	1710.5	33.32	1680.5	33.32	1725.5	33.32	1680.5	33.32
Hydrogenation1	1170.3	33.32	1110.3	33.32	1140.3	33.32	1170.3	33.32	1200.3	33.32	1140.3	33.32
Boiler	1954.8	39.43	3690.2	39.43	2914.5	39.42	3700.1	39.43	2492.2	39.41	3733.1	39.43
CDU1	5037.0	39.43	5009.1	39.43	2987.5	39.42	5023.9	39.43	4581.9	39.41	5067.0	39.43

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Table 8
Supplementation strategy of the backup fuels for some of the units.

B	*D*	1	2	3	4	5	6
Fuel oil/t	CDU2	4.2768	5.5454	3.4059	5.7492	3.5037	5.5675
	PX	12.5925	12.6874	12.8081	12.7060	12.4845	13.4434
	Boiler	1.7331	0	0.7978	0	1.2046	0
	CDU1	0	0	2.0358	0	0.4508	0
LPG/N m^3	Reforming	1213.6784	939.1038	875.4462	968.0937	1156.6839	1072.1104
Natural gas/N m^3	–	4541.864	4507.425	4582.340	4527.756	4694.080	4480.283

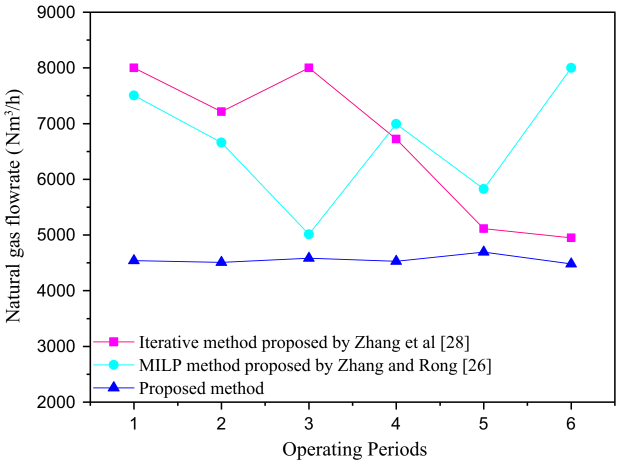

Fig. 4. Natural gas consumption obtained by different methods.

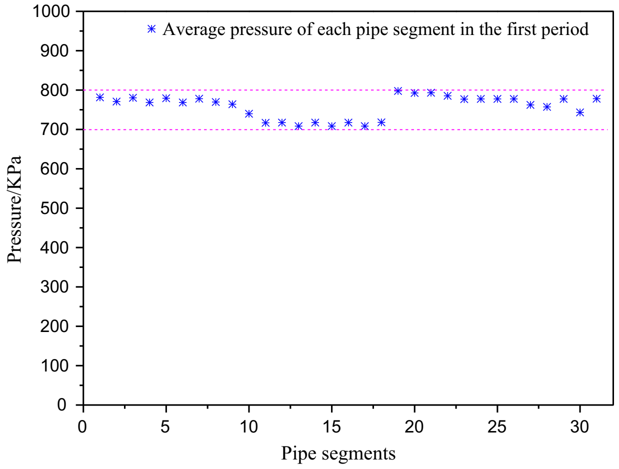

Fig. 6. Average pressure of each pipe segment in the first operating period.

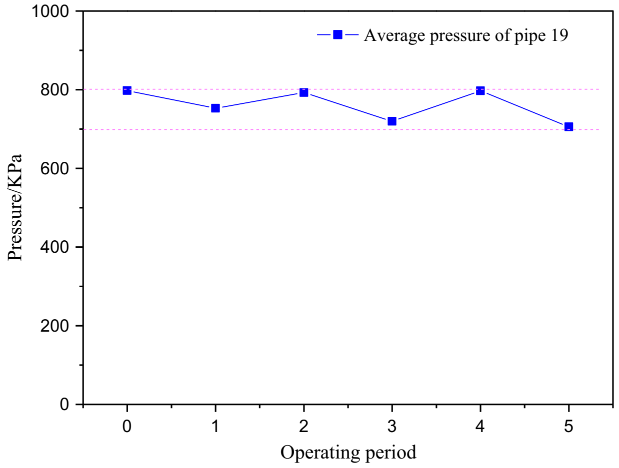

Fig. 5. Average pressure of pipe segment 19 in each operation period.

obtained by the present work with the results presented by Zhang et al. [28]. As it is shown, the result obtained by the proposed method uses less natural gas. Instead, more fuel oil is consumed. This is because the price of the fuel oil (0.05 yuan/MJ) is lower than that of the natural gas (0.06 yuan/MJ), based on unit heating value. Furthermore, the supply schedule obtained by the present study is more stable during the entire time horizon. A stable supply schedule of the external fuel is easier to manage and better for maintaining the safety system pressure. The average pressure of the pipe connecting the natural gas source to the network is given in Fig. 5. Fig. 6 illustrates the average pressure of each pipe in the network during the first operating period. As shown, the pressure of each pipe segment is kept under the safety level.

By applying the proposed method, fuel resources are high efficiently utilized. There is no low pressure fuel gas being released to flare or high pressure fuel gas being sent to low pressure system. Fig. 7 illustrates the comparison between the energy supply and the energy demand of each consuming unit in the 1st operating period. By comparing the energy supply scheme with the energy demand, it can be concluded that: (1) for the result obtained by the proposed method, the energy supplied to each consuming unit satisfies the actual demand just with the right amount. Neither energy shortage occurred nor large amount of superfluous energy were consumed. (2) For the result obtained by Zhang et al. [28], however, both energy shortage and excessive energy consuming occurred. For example, the hydro-cracking and the hydrogen producer unit, the energy provided are less than the demand, while

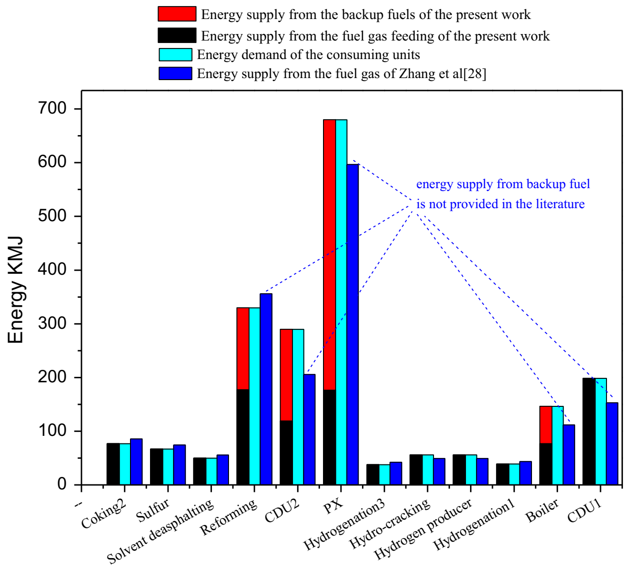

Fig. 7. The energy supply and demand of each unit in the 1st operating period.

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Table 9
Comparison of the optimization result for different methods.

Items	Cost/yuan	Consumption	HV/MJ	Total HV/MJ	Total cost/yuan
Proposed method					
Natural gas	54667.497	27333.735 N m^3	910869.8	6,133,304	338903.8
Fuel oil	221985.161	110.993 t	4,439,700		
LPG	62251.164	6225.116 N m^3	782733.7		
Zhang et al. [28]					
Natural gas	80,000	40,000 N m^3	1,332,960	4,352,887	–
Fuel oil	113,092	56.546 t	2,261,840		
LPG	60,291	6029.1 N m^3	758,087		

Total energy demand: 12,589,000 MJ; total internal energy supply: 6457308.2 MJ.

for the coking2 and the sulfur unit the energy supplied is greater than the demand. It should be noted that the energy supply of the backup fuels of the result in literature is not shown, because it is not given by Zhang et al. [28]. These units for which the backup fuels are available are not compared.

Table 9 gives the comparison of the optimization result between the present work and the method proposed in literature [28]. The total energy demand of all the consumers during the whole period is 12,589,000 MJ, which can be calculated from Table 5. It can also be derived from Tables 1 and 2 that the energy provided by all the fuel gas produced during the whole time horizon is 6457308.2 MJ. If we compare the total energy demand with the total energy supply during the whole time horizon, we can see that for the proposed method, the total energy supply 12590612.2 MJ exceeds the total demand only by 0.0128%. The effective utilization rate of energy is up to 99.987%. However, for the result obtained in literature [28], the total energy supply 10810195.2 MJ cannot fully cover the total energy demand. As a result, the overall operation cost of the two optimization schedule is not compared. It can be concluded that, the scheduling result obtained by the proposed method is practical and preferable. For the proposed method, isothermal conditions and ideal gas behavior is assumed. So the model is only valid for the systems operating under a relatively low pressure and temperature.

7. Conclusion

A multi-period MPEC optimizing model for the scheduling of the fuel gas system in refinery is presented. The proposed optimization tool is based on detailed pipeline network model considering reversal and transition flow. The multi-component nature of the gas system is also taken into consideration. Pipeline systems with branching structure as well as loop structure can be solved simultaneously with the scheduling problem. Complementarity formulations are used to model the discrete elements instead of the commonly used binary variables. The case study illustrates that by applying the developed model practical and favorable scheduling scheme can be obtained with rational computational effort.

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