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Hydrogen distribution in the refinery using mathematical modeling

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ABSTRACT

The increased requirement of hydrogen in the refinery is fulfilled by proper distribution of available hydrogen in the refinery, using additional hydrogen production system and by importing. Amongst these options first one is cost effective as no addition source is required. Thus, the present paper deals with optimum distribution of hydrogen. For this purpose mathematical models are developed based on pressure constraints, source and sinks constraints, compressor flow rate recycle and purity constraints, flow combinations, hydrogen consumption, operating cost, capital cost, payback period etc. The model is developed in stages to show the improvements.

Two case studies, Cases A and B, are considered in this paper and for these linear programming (LP), nonlinear programming (NLP), mixed-integer linear programming (MILP) and mixed-integer nonlinear programming (MINLP) models are developed. The results predicted using these models are compared to select the best network. The optimized network for Case A consumes 20.9% less hydrogen, whereas, for Case B it is reduced by 32.3%.

1. Introduction

The energy starved civilization is always in search of abundant sources of clean and cheap fuels to meet its energy requirement. The demand for clean fuel sources and stringent environment regulation are advocating planners to select hydrogen as fuel for future generation.

It appears from literature that firstly NASAs space program used hydrogen as a fuel for rockets and for fuel cells to generate electricity in 1958 [1]. Further, in response to global oil market disruptions International Energy Agency (IEA) was established in 1977. IEA activities included the research and development of hydrogen energy technologies in the oil industry. Since then several trends are leading to an increased demand of hydrogen in refineries and petrochemical sites. Firstly, stricter legislation on sulfur content in fuels increases the need for hydrotreating. Secondly, processing of heavier crude oils and reduced market for heavy fuel is forcing to greater use of hydrocracking. These factors lead to a deficit in hydrogen balance of a refinery and causes investment for additional hydrogen production facilities e.g. steam reforming or else importing from outside suppliers. These are often the expensive investments. In contrast to it investigators have thought to reduce hydrogen requirement in the refinery.

There are two ways to reduce hydrogen demand of process: (i) to improve design of the individual consumer units and (ii) to generate better hydrogen distribution network for refinery. The first option was considered by Hiller et al. [2]. They modified hydrocrackers in order to reduce the hydrogen usage. Further, Baade et al. [3] discussed methods for expanding steam reformers and Kramer et al. [4] addressed design problems of hydrogen plants. Liu and Zhang [5] proposed a systematic methodology to select appropriate purifiers from pressure swing adsorption (PSA) processes, membranes or hybrid systems for recovering hydrogen from refinery off-gases and showed the cost savings. Liao et al. [6] considered purification technologies as a key solution for refinery hydrogen management. They showed through mixed-integer nonlinear programming (MINLP) model that the performance of the purifier is affected by operating conditions and their placement.

These approaches to reduce hydrogen requirement by improving the performance of individual equipment are all important. However, it is arguably more important to consider all hydrogen suppliers and consumers as part of a refinery and integrate these for maximum possible utilization of available hydrogen.

It appears that limited literature is available for designing and optimization of hydrogen distribution network. Towler et al. [7] proposed a method for analyzing hydrogen network by comparing the cost of recovering hydrogen. It gives the engineer simple and graphical insight into refinery profitability based on

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Nomenclature

F	flowrate
m	binary variable
y	hydrogen purity
Subscripts	
i, sr	source
j, sk	sink
R	recycle
max	maximum
comp	compressor

process design and operational fundamentals. However, they did not consider the physical constraints that influence the distribution network. The most promising and industrially exploited approach was developed by Alves [8] based on pinch technology [9]. The method identifies sources and sinks of hydrogen. Sinks are units which consume hydrogen such as hydrotreater and hydrocracker. However, refinery fulfils demand of hydrogen by generating it in plant using catalytic reformer or purchasing from outside. These are considered as sources. Alves [8] proposed a linear programming (LP) model for optimizing the hydrogen network. The major limitation with this method is that the targets are based only on flowrate and purity requirements. The targeting method assumes that any stream containing hydrogen can be sent to any consumer, regardless of stream pressure. In reality, this will only be true if the stream is at a sufficiently high pressure. Thus these targets are very far from a real design.

Further, Hallale and Liu [10] developed an improved method for hydrogen network accounting pressure constraints as well as economy. They considered that the inlet pressure was equal to lowest pressure amongst all sources which were fed to compressor. Based on this fact the compressor was designed to be able to receive the lowest pressure source and all sources of a higher pressure that also fed the machine were passed through a valve to drop their pressure. However, in actual refinery the input pressure of compressor changes continuously.

The optimization of Porto Refinery of the Galp Energia network using LP model is carried out by Fonseca et al. [11]. However, it is already discussed that all variations in process cannot be accounted through LP model. Khajehpour et al. [12] adopted the approach of Hallale and Liu [10] and proposed reduced superstructure based on heuristic rules, engineering judgment and tuned parameter. This method gave a theoretical solution, which was not necessarily applicable in a real system.

In contrast of these limitations the present paper focused on the development of an improved model for hydrogen distribution in the refinery which considers for pressure constraints, existing compressors and purification system. The method is not limited to minimising hydrogen, but also accounts for hydrogen consumption, operating cost, capital cost, payback period, etc. The model is modified by considering variable inlet and outlet pressures as well as recycle rate of compressor to make it more suitable for the real system. The model is developed in stages to show the improvements.

2. Problem statement

To develop the modified model an existing hydrogen network is considered from open literature [10]. Henceforth, it is referred as Case A. Hydrogen is supplied from two sources: a catalytic reformer (CRU) and a hydrogen plant. Currently, 45 MMscfd (million standard cubic feet per day) and 23.5 MMscfd of hydrogen are produced in the hydrogen plant and CRU, respectively. Hydrogen is consumed by hydrocracker unit (HCU), diesel hydrotreater (DHT), kerosene hydrotreater (KHT), cracked naphtha hydrotreater (CNHT), naphtha hydrotreater (NHT), hydrodealkylation (HDA) and fuel gas system (Fuel). There are two make-up compressors in the system and all consumers except isomerisation plant have internal recycle compressors. The existing network and process data are shown in Fig. 1 and Table 1, respectively. The cost data is presented in Table 2.

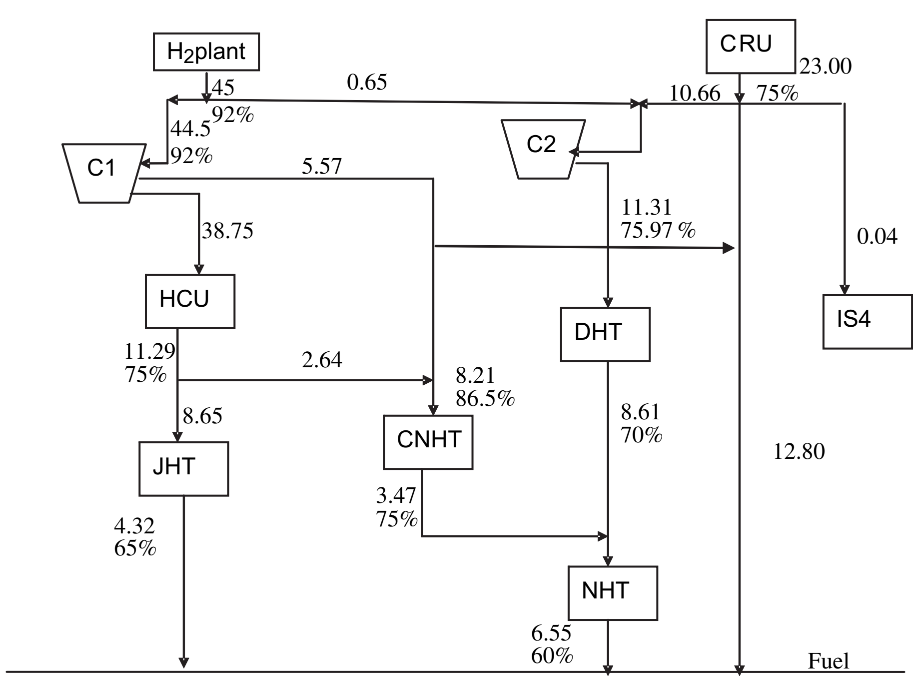

Fig. 1. The existing network for hydrogen.

Table 1
Source and sink data of existing hydrogen network for Case A.

Process	Make-up			Purge			Recycle
	Flow rate (MMscfd)	Purity	Pressure (Psia)	Flow rate (MMscfd)	Purity	Pressure (Psia)	Flow rate (MMscfd)
DHT	11.31	0.7597	600	8.61	0.7	400	1.56
CNHT	8.21	0.8653	500	3.47	0.75	350	36.75
JHT	8.65	0.75	500	4.32	0.65	350	3.6
NHT	12.08	0.7144	300	6.55	0.6	200	3.59
IS4	0.04	0.75	300				
HCU	38.78	0.92	2000	11.29	0.75	1200	85.7
H_2 supply							
	Flow (MMscfd)	Max flow (MMscfd)	Purity	Pressure (Psia)			
H_2 plant	45	50	0.92	300			
CRU	23.5	23.5	0.75	300			

The power co-relation for compressor is taken from Hallele and Liu [10]. Several practical constraints have been imposed by the refinery such as (i) there is only space on the site for one new compressor and one new purification unit; (ii) PSA units are preferred for purification because they do not require feed pre-treatment and (iii) maximum possible payback period for this system is two year.

3. Analysis of hydrogen network

3.1. Nonlinear programming (NLP) model

Alves [8] developed the LP model for optimization of hydrogen distribution based on an assumption that any source can supply hydrogen to any sink. This is the major drawback of LP model as match is feasible between source and sink only when pressure difference exists or at least pressure should be equal. However, for the nonlinear method of analysis, a superstructure is made in such a way that flow rates from sources to sinks are possible only if the pressure of corresponding source is higher or equal to that of the sink. If it is less then the flow rate is taken as zero. For Case A the superstructure is shown in Table 3. The linear constraints of flow and purity are shown through Eqs. (2)–(4) where purity of all sources is known.

For this model compressors are also considered which act both as sinks and sources. As many sources are mixed before entering a compressor the hydrogen purity as well as flow rate of compressors is not known a priori and so the problem is of nonlinear in nature. These constraints are shown through Eqs. (5) and (6). The flow of compressor is not known but it should be lesser than the maximum specified flow rate. So, a flow constraint is used for a compressor which is shown in Eq. (7). For the existing compressor maximum flow is the design flow rate plus a margin of few percent and for a new compressor this is the manufacturer's limitation.

Once all the variables and constraints are known, the system is optimized using the following objective function:

Table 2
Cost data for Case A.

Particular	Cost
Hydrogen cost	\$2000/MMScfd
Electricity cost	\$0.03/kWh
Fuel costs	\$2.5/MMBtu
Compressor costs [18]	115 + 1.91 * power (kWh) (k\$)
PSA [7]	503.8 + 347.4 * flow (MMScfd) (k\$)
Export cost [15]	16.85 \$/GJ

$$\text{Objective function} = \text{Total operating cost} + \text{Total capital costs} \tag{1}$$

where total operating cost = Hydrogen costs + electricity costs − fuel costs; Total capital costs = compressor cost + purifier cost + Piping cost

Fuel cost is function of total operating costs. The negative sign indicates that it replaces certain amount of natural gas which is to be burnt in heaters and other devices.

The constraints are as follows:

$$\text{Sink flow rate} \quad \sum_{i=1}^{\mathrm{sr}} F_{i,j} = F_{\mathrm{sk},j} \tag{2}$$

$$\text{Source flow rate} \quad \sum_{j=1}^{\mathrm{sk}} F_{i,j} + F_{\mathrm{F},i} = F_{\mathrm{sr},i} \tag{3}$$

$$\text{Sink purity} \quad \sum_{j=1}^{\mathrm{sr}} F_{i,j}\left(Y_{\mathrm{sr},i} - Y_{\mathrm{sk},j}\right) = 0 \tag{4}$$

$$\text{Flow rate balance around compressor} \quad \sum_i F_{\mathrm{comp},j} = \sum_j F_{\mathrm{i,comp}} \tag{5}$$

$$\text{Hydrogen balance around compressor} \quad \sum_i F_{\mathrm{comp},j} Y_{\mathrm{comp}} = \sum_i F_{i,\mathrm{comp}} Y_i \tag{6}$$

$$\text{Capacity limit around compressor} \quad \sum_j F_{i,\mathrm{comp}} \le \sum_j F_{\mathrm{max,comp}} \tag{7}$$

One important assumption is considered in this model after referring to various networks discussed by Shahraki et al. [13] that piping cost is less than 15% of total capital cost. So in the optimization problem, capital costs are taken as sum of compressor and PSA costs and then assumed that this is 85% of total capital cost. The power of compressor is computed using the method shown in the work of Hallale and Liu [10].

3.1.1. Results predicted using NLP model

The existing network shown in Fig. 1 is optimized using NLP model which gives hydrogen flow as 45.6 MMscfd which is close to the literature value [10]. Further, to improve this network a number

Table 3
Superstructure for Case A.

Sources	No	Pressure	Sinks	No	Pressure	Non zero flow between sources and sinks
H_2 plant	1	300	HCU	1	2000	F1,5; F1,6; F1,7; F1,8; F1,9; F1,10
CRU	2	300	CNHT	2	500	F2,5; F2,6; F2,7; F2,8; F2,9; F2,10
HCU	3	1200	DHT	3	600	ALL except F3,1
CNHT	4	350	JHT	4	500	F4,5; F4,6; F4,7; F4,8; F4,9; F4,10
DHT	5	400	NHT	5	300	F5,5; F5,6; F5,7; F5,8; F5,9; F5,10;
JHT	6	350	IS4	6	300	F6,5; F6,6; F6,7; F6,8; F6,9; F6,10;
NHT	7	200	Compressor 1	7	300	F7,9; F7,10
Compressor 1	8	2000	Compressor 2	8	300	All
Compressor 2	9	600	Fuel	9	–	All except F9,1
New compressor	10	300	New compressor	10	200	F10,5; F10,6; F10,7; F10,8; F10,9; F10,10

Table 4
Results of NLP model for Case A.

Cases (modification done on existing network)	Hydrogen consumed (MMscfd)	Operating cost ($/year)	Capital cost for new investment ($/year)	Total annual cost ($/year)	Payback (year)
Existing	45.62208	21,946,947.72	–	–	–
One new compressor	45.10778	21,904,161.39	383,243.86	22,188,046	0.643
One new compressor and new PSA	29.55064	16,500,277.63	11,328,853.5	24,892,021	1.888
One new PSA	32.83961	18,032,959.71	7,598,037.67	23,661,136	1.7

Table 5
Revised results of NLP model for Case A.

Cases (modification done on existing network)	Hydrogen consumed (MMscfd)	Operating cost	Capital cost for new investment	Export cost of hydrogen	Total annual cost
Existing	45.62208	21,946,947.72	–	–	–
One new compressor	45.10778	21,904,161.39	383,243.86	836,663.2	21,450,742
One new compressor and new PSA	29.55064	16,500,277.63	11,328,853.5	26,145,019	1,684,113
One new PSA	32.83961	18,032,959.71	7,598,037.67	20,794,522	4,836,475

of modifications are considered based on combinations of new compressor and PSA units.

A comparative study of different modifications with the existing system is presented in Table 4 which shows hydrogen consumption, operating cost, capital cost and payback period for each modification. Payback period is computed as the ratio of capital costs needed for modification to saving in operating costs. In this table only hydrogen supplied from the plant is mentioned assuming that hydrogen provided by CRU remains fixed to a value of 23.5 MMscfd. Capital cost mainly includes the costs of new compressor and/or PSA as cost of exiting system remains same for all modification. Based on the data shown in Table 4 it is found that existing network with one new compressor has a minimum total annual cost assuming that total service life of a compressor is 10,000 h [14]. It also has a minimum payback period. However, the hydrogen requirement is almost remain same as compared to existing network.

On the other hand, if the production of hydrogen in H_2 plant is considered to be equal to 45.6 MMscfd then the remaining amount of hydrogen from each modification may be exported. The export cost of hydrogen is taken from the work of Shiga et al. [15]. This cost is a profit so it must be deducted from total annual cost. The revised data are shown in Table 5 which indicates that optimum case is the existing system with new compressor and PSA unit. The respective network is presented through Fig. 2. This network is required substantial repiping. Here new compressor is used to accommodate the increased recycle requirement of NHT as well as to compress one of the feeds to the Fuel. Total payback period for this modification is only 0.35 years.

The above discussion shows that NLP model depicts the real refinery system and it is ensured that a cost benefit refinery can be designed considering the suggested modifications. The only disadvantage is the time taken for solving it and one should use more advanced versions of solver because this model has more variables and constraints.

3.2. MINLP model

A wide range of nonlinear optimization problems involve integer or discrete variables in addition to the continuous variables. These classes of optimization problems are denoted as MINLP problems [16]. It refers to nonlinearities in the objective function and constraints with continuous and discrete variables. MINLPs have been used in various applications, including the process industry, finance, engineering, management science and operation research sectors. MINLP problems are comparatively difficult to solve as these combine all difficulties of both subclasses: the combinatorial nature of mixed-integer programs (MIP) and NLP.

In MINLP model binary variables are used to simulate the existence or non-existence of units (in its 0, 1 combination) whereas continuous variables represent input–output interaction relationships among individual units and operation.

In the NLP formulation compressor is modeled based on inlet and outlet variables but did not consider the recycle rate. However, sometimes it might be specified by the refinery that an additional capacity of 5% or 10% for the compressor can be considered in the recycle rate, such that recycle rate could not exceed a maximum value. Thus, recycle rate is also an optimization variable. It is done to avoid the problem of recycle compressors which affects the entire network. This helps not to shut down all units instead of one defective unit. This factor is accounted in MINLP formulation where recycle rate along with compressor is considered as a single system.

The objective function is same as Eq. (1). However, the constraints are modified as given below:

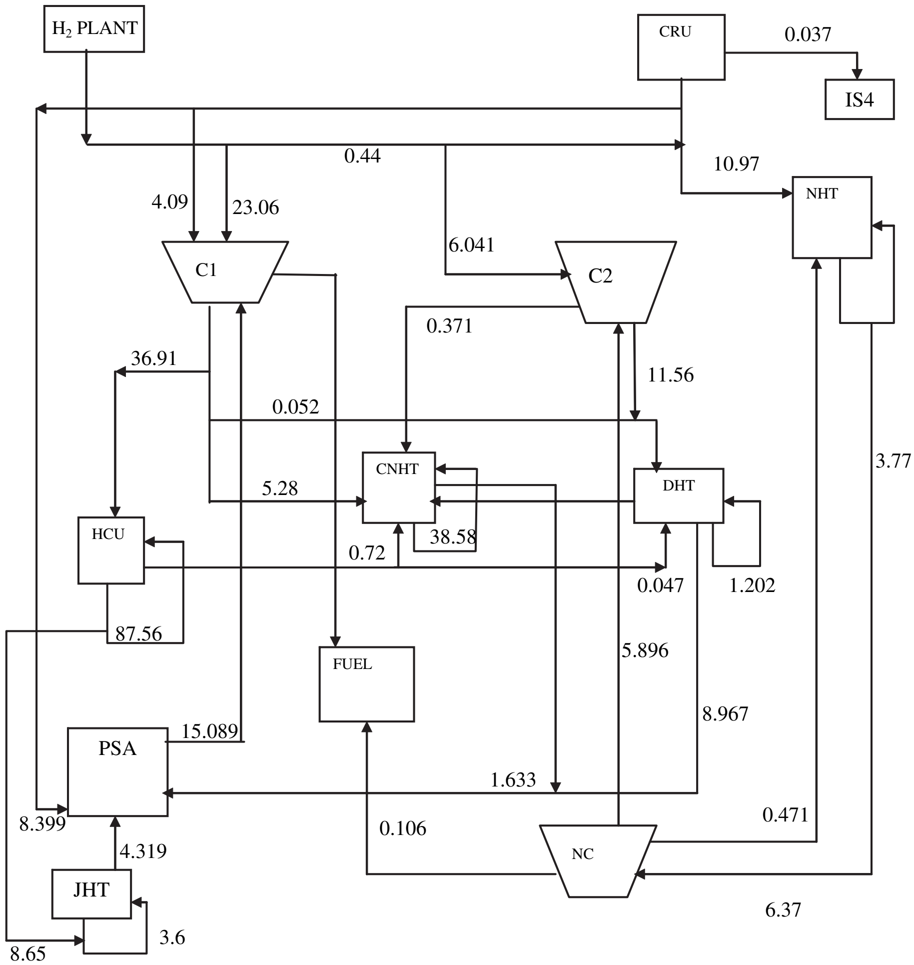

Fig. 2. Optimized network for Case A with PSA and compressor using NLP model.

Flow rate balance: For a compressor the flowrates of gas entering and leaving must equal:

$$\sum_i \left[\left(F_{\mathrm{comp},j} + F_{\mathrm{comp,R}}\right) \times m(\mathrm{comp},j)\right] = \sum_i \left[F_{i,\mathrm{comp}} \times m(i,\mathrm{comp})\right] \tag{8}$$

Hydrogen balance: The amount of pure hydrogen entering the compressor must equal the amount leaving.

$$\sum_i \left[F_{\mathrm{comp},j} Y_{\mathrm{comp}} + F_{\mathrm{comp,R}} Y_{\mathrm{R}}\right] \times m(\mathrm{comp},j) = \sum_i \left[F_{i,\mathrm{comp}} Y_i \times m(i,\mathrm{comp})\right] \tag{9}$$

where

$$0 \le \sum_i F_{\mathrm{comp,R}} \le \sum_i \left(0.1 \times F_{i,\mathrm{comp}}\right) \tag{10}$$

Table 6
MINLP result for Case A.

Costs, \$	NLP model	MINLP model
Hydrogen consumption, MMscfd	69.1	54.6
Operating cost, \$/yr	21,946,947.92	17,127,341.78
Electricity cost, \$/yr	1,780,395.003	990,670
Fuel cost, \$/yr	13,137,564.32	16,713,000

Capacity limit:

$$\sum_j F_{i,\mathrm{comp}} m(i,\mathrm{comp}) \le \sum_j F_{\max,\mathrm{comp}} \tag{11}$$

Sink flow rate constraint:

$$\sum_{i=1}^{\mathrm{sr}} F_{i,j} m(i,j) = F_{\mathrm{sk},j} \tag{12}$$

Source flow rate constraint:

$$\sum_{j=1}^{\mathrm{sk}} F_{i,j} m(i,j) = F_{\mathrm{sr},i} \tag{13}$$

Sink purity:

$$\sum_{j=1}^{\mathrm{sr}} F_{i,j}\left(Y_{\mathrm{sr},i} - Y_{\mathrm{sk},j}\right) = 0 \tag{14}$$

Pressure constraints: Inlet pressure to the compressor is greater or equal to lowest pressure over all sources that feed the compressor. Similarly, the discharge design pressure is lesser or equal to the highest pressure over all sinks that are fed to new compressor.

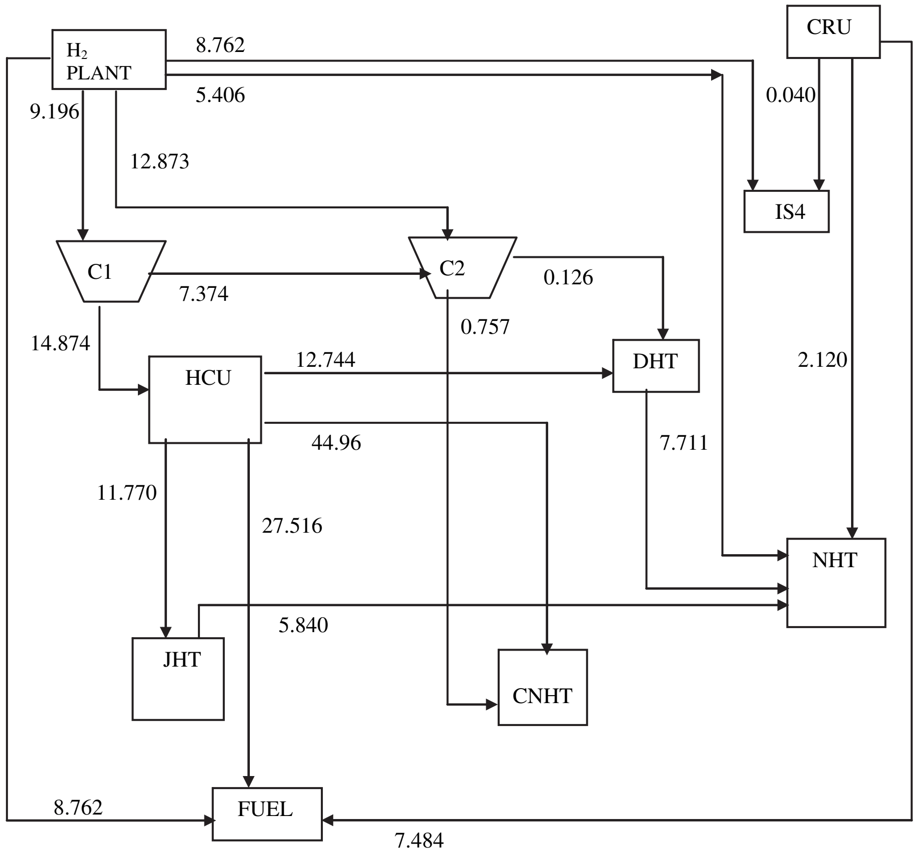

Fig. 3. Optimized network for Case A using MINLP.

The uncertainty is shown through Eq. (10) as for each match $F_{comp,R}$ is between zero and ($0.1 \times F_{i,comp}$). In this model nonlinearity arises due to constraints of hydrogen balance around compressor. The capacity limit around the compressor is also defined where maximum flow rate is either design value or from manufacturer's limitation. Moreover, in this model inlet and outlet pressures are also optimization variables.

It must be pointed out that the mathematical program now becomes a MINLP. Integer variables are required to account the existence of streams entering to the compressors. So, multiplication of binary variables with continuous determines the existence or non-existence of flow combination in this optimized hydrogen network depending on the value of 0 or 1.

Mathematical modeling software, GAMS, is used for solving the present MINLP model [17]. Structural and parameter optimization can be performed simultaneously. This is clearly a very important point since in process synthesis discrete decisions that are involved in the selection of a flowsheet also implies continuous decisions such as selection of flowrates, which in some cases can have a very important effect on the structure. Discrete and logical constraints can be handled explicitly with the binary variables. This feature is very powerful because it allows the designer to specify structural constraints or conditions that will yield realistic flowsheet configurations.

3.2.1. Results predicted using MINLP model

Case A is optimized using MINLP model and results of comparison with NLP model are summarized in Table 6. It shows that for MINLP model the predicted operating cost is 21.9% less in comparison to NLP model. However, the capital costs of networks in both models are same as no extra units are involved in either case. The revised network, based on MINLP model is presented in Fig. 3, which is considerably simple than the existing network, shown in Fig. 1.

Further, it is seen that the same problem is optimized using model developed by Hallale and Liu [10]. It is found that their network consumes 6% more hydrogen in comparison to present network. It also includes PSA as additional equipment. Both operating and capital costs are decreased using present MINLP model and piping cost is also less in comparison to Hallale and Liu [10] network. Thus, present model gives more cost effective design based on pressure constraints, source and sinks constraints, compressor flow rate recycle and purity constraints as well as flow combination.

4. Case study: case B

The study presented above is applied to optimize a complex refinery hydrogen network taken from Alves [8]. The exiting network diagram is shown in Fig. 4 comprising five compressors. It consists of purge, recycle and make-up of hydrogen for the present hydrogen distribution system. The necessary data is shown in Table 7. The current system has following hydrogen consuming processes: HCU, catalytic cracking feed hydrotreater (CCHT), CNHT, two lubricant units (Lube1 and Lube2), DHT, KHT and NHT. Here hydrogen is supplied by Import, H_2 plant and CRU. The import cost of hydrogen is 1.67 \$/kmol H_2.

4.1. Network optimized using LP, NLP and MILP

To show the effectiveness of present work the hydrogen network is developed through LP, NLP and mixed-integer linear programming (MILP) for case study and the results are compared.

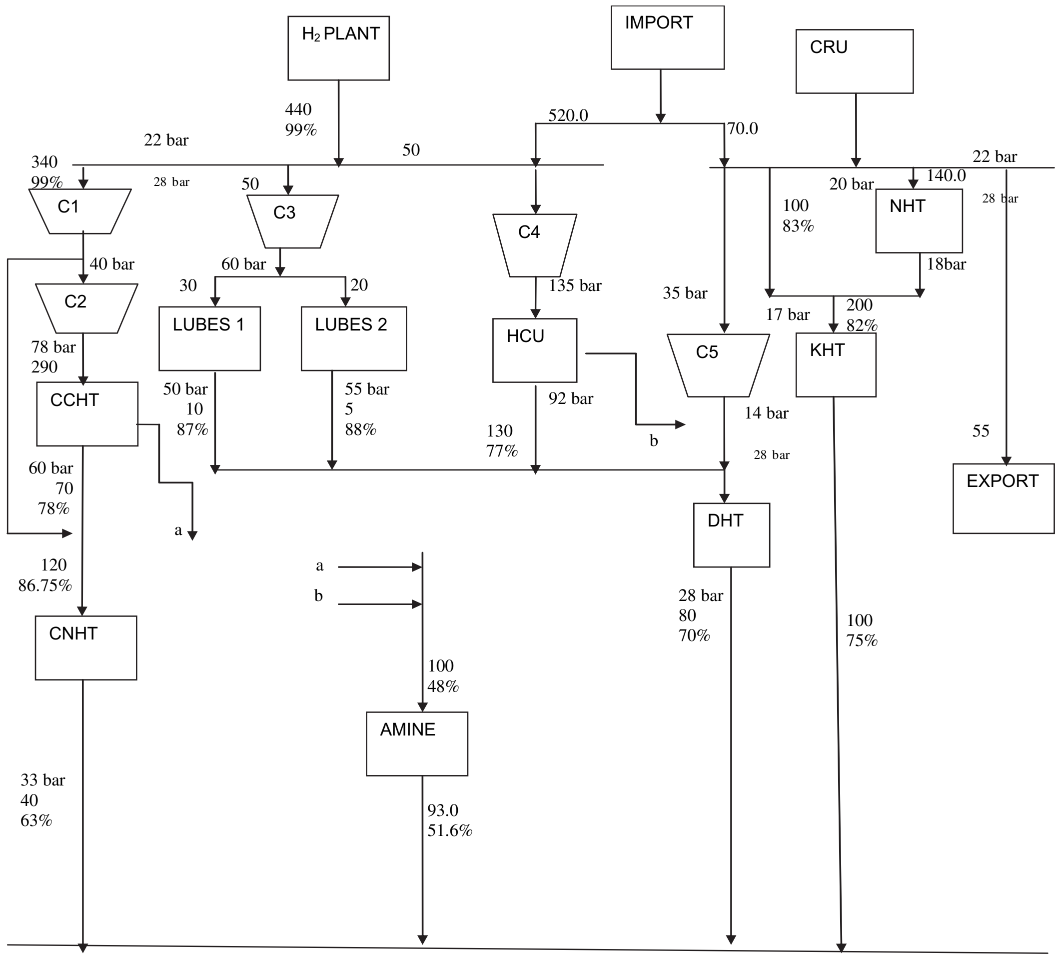

Fig. 4. Existing refinery network for Case B.

Table 7
Source and sink data for Case B.

Units	Make-up		Purge		Recycle	Source		Sink	
	Flow rate (mol/s)	Purity	Flow rate (mol/s)	Purity	Flow rate (mol/s)	Flow rate (mol/s)	Purity	Flow rate (mol/s)	Purity
Lubes1	30	0.99	10	0.87	150	180	0.89	160	0.87
Lubes 2	20	0.99	5	0.88	100	120	0.8983	105	0.88
HCU	570	0.935	130	0.77	2000	2570	0.8067	2130	0.77
CCHT	290	0.99	70	0.78	800	1090	0.8359	870	0.78
CNHT	120	0.867	40	0.63	400	520	0.6848	440	0.63
DHT	270	0.829	80	0.70	250	520	0.7642	330	0.7
KHT	200	0.82	100	0.75	0	200	0.82	100	0.75
NHT	140	0.83	100	0.81	0	140	0.83	100	0.81
H_2 plant						440	0.99		
Import						590	0.93		
CRU						350	0.83		
LPPs						93	0.5161		
Export								55	0.83

Table 8
Results of LP, NLP and MILP model for Case B.

Parameter	Existing system	LP model		NLP model	MILP model
Units		PSA	PSA	NC, PSA	PSA
Operating cost, \$/yr	30,000,000	27,000,000	21,090,439	21,412,804.6	24,561,261.7
Hydrogen import, mol/s	590	497	446.3	399	497

Firstly, a LP model is developed. The objective function and constraints are written in similar manner as shown for the case of Hallale and Liu [10] problem. In the problem it is assumed that flow exists from all sources to all sinks and they are optimization variables. Following this assumption 13 sources (import, H_2 plant, CRU, LPPs, Lube1, Lube2, HCU, CCHT, CNHT, DHT, KHT, NHT and PSA) and 11 sinks (Lube1, Lube2, HCU, CCHT, CNHT, DHT, KHT, NHT, Export, PSA feed and Fuel) have been identified. So, there are total 143 variables. A linear model is solved for Case B using Microsoft excel 2007 and the results are compared with the existing model. The

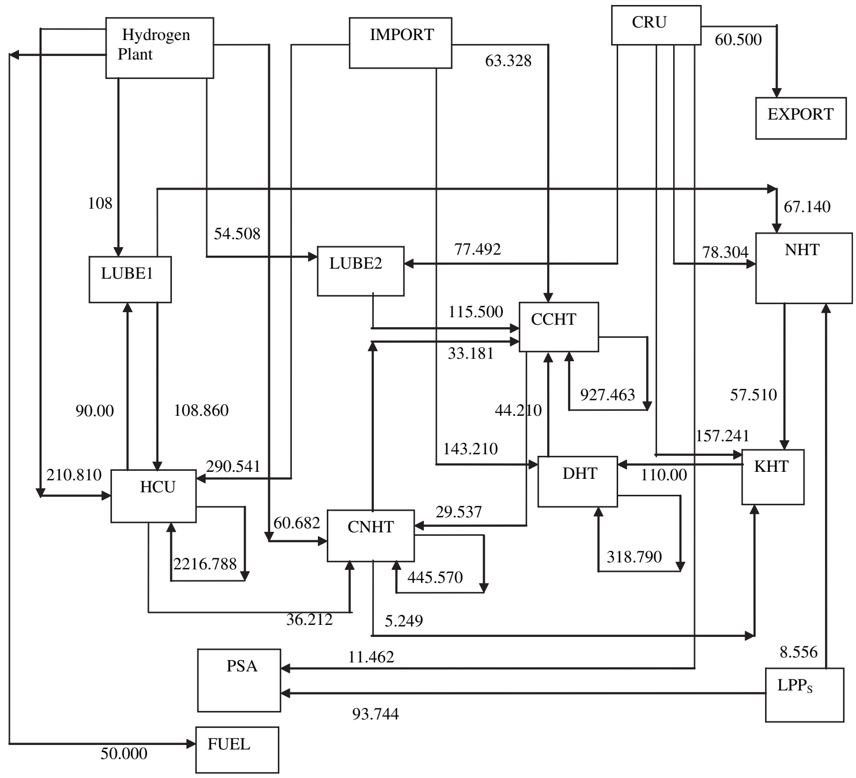

Fig. 5. Optimized network for Case B using LP model.

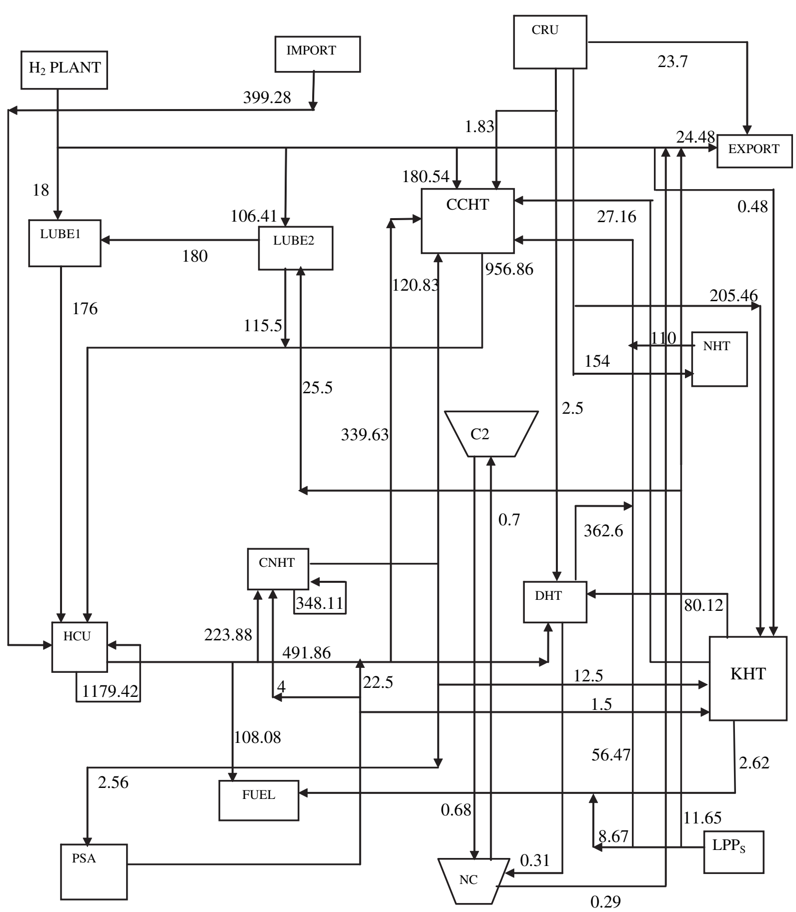

Fig. 6. Optimized network for Case B using NLP model.

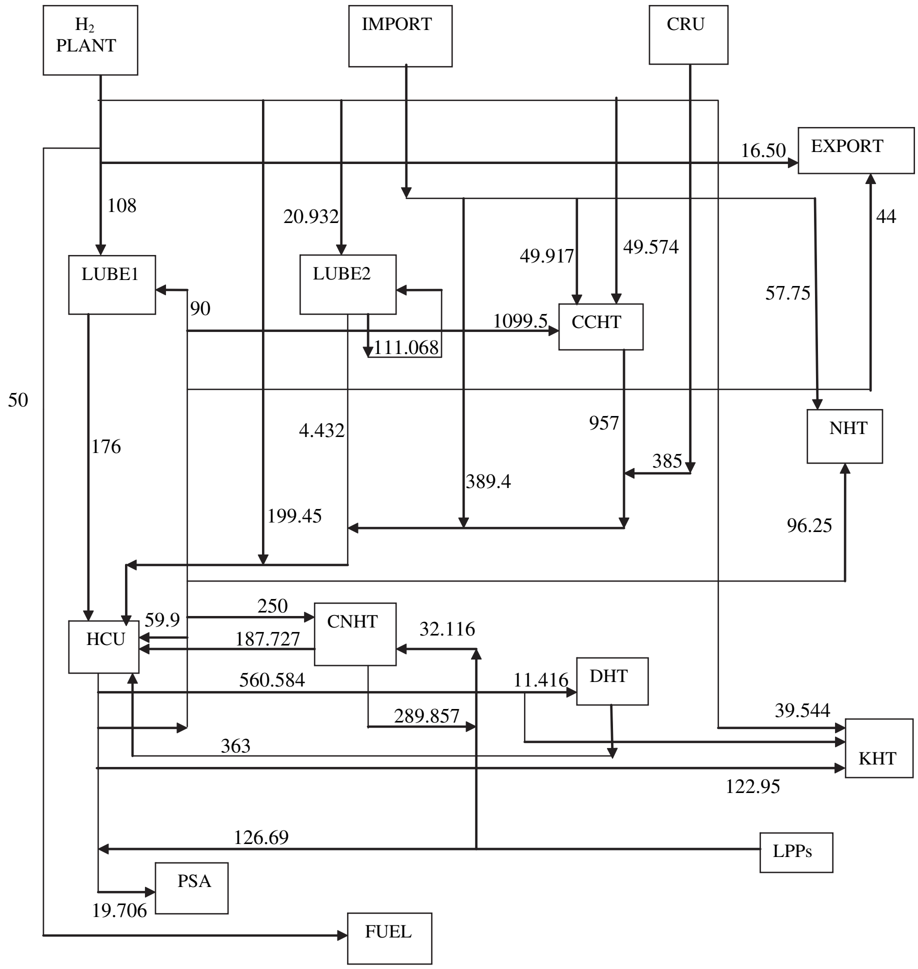

Fig. 7. Optimized network for Case B using MILP model.

results of LP model are summarized in Table 8 and network is shown in Fig. 5 which has no compressor.

In LP method of analysis it is assumed that any source can supply hydrogen to any other sink unless otherwise specified. This assumption is relaxed by considering the pressure constraints. For flow to happen, the pressure of source should be greater or at least equal to that of sink. Here compressor model is also included which makes the system nonlinear. So, total 11 sources and 11 sinks are identified including the new compressors and PSA units. In this problem number of variables is reduced to 121. The optimized results of NLP model and respective network are presented in Table 8 and Fig. 6, respectively. The operating cost is reduced by 20.7% using NLP model. For Case B only three compressors (C1, C2 and NC) are used in comparison to five involved in Fig. 4. However, the flow rate of C1 comes out as 0.0006 mol/s, which is negligible. Therefore, in Fig. 6 only compressors, C2 and NC exist. On the other hand it is considered as compressor C1 is repositioned in place of NC. So no new compressor is required.

Further, a mixed-integer model is developed for the Case B and results are compared with LP and NLP models. For this analysis hydrogen network is modeled as a superstructure that contains all the feasible connections between hydrogen sources and sinks of the process. The decision variables for the MILP model are hydrogen flow rate between sink and source $F_{i,j}$, the connection existence between source and sink $F_{i,j}(0,1)$. In this model, the purity and pressures are assumed to be constant. The continuous variables are associated to flowrates, equipment sizes, pressures and temperatures, while 0–1 binary variables are connected to the existence of combination. In this model all equation are linear in nature.

For Case B there are 13 sources and 11 sinks. So, maximum 143 flow combinations may exist in this refinery but all combination cannot occur simultaneously. By declaring these flow combinations as binary variable different hydrogen network can be obtained. In this way most profitable and feasible structure out of all networks is selected. After optimization thirty flow combinations are obtained and network is shown in Fig. 7. The results of MILP model are summarized in Table 8.

Table 8 shows that hydrogen import is reduced to 32.4% using NLP model with 2 compressors and one PSA. For LP and MILP model though amount of hydrogen import is same in both cases, MILP gives less complex network. The same amount of hydrogen in both models is due to the linear objective function and constraints. These models are not included compressor so that less purity hydrogen can be compressed to give high purity source and utilize in the process. The capital cost of network obtained from NLP is higher than that of LP and MILP as two compressors are included in it. However, this cost is considerably less in comparison to existing system where five compressors are installed, as shown in Fig. 4. Moreover, NLP gives more realistic network than LP and MILP.

Based on above discussion it is seen that mathematical modeling is the better option to optimize the network. As we move to complicated models more constraints can be considered which cause the problem more realistic and in this case network can directly be used as final design. For example MINLP model has pressure constraints, source and sinks constraints, compressor flow rate recycle and purity constraints as well as flow combination considerations which gives realistic and simplified network.

5. Conclusions

The salient conclusions of the present study are as follows:

1) Mixed-integer linear and nonlinear programming technique is considerably better than linear techniques as it provides the less complex and more realistic refinery system. It directly gives the optimized network which saves the considerable time of the designer.
2) Based on two case studies it is found that the average reduction of hydrogen consumption in refinery is 26.6% using present modeling techniques.
3) The MINLP model is able to account many complexities of real refinery system such as pressure constraints, source and sinks constraints, compressor flow rate recycle and purity constraints as well as flow combinations.
4) The present technique can be applied to any refinery systems if the refinery network is available with appropriate data.

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