

A Novel Process for Natural Gas Liquids Recovery from Oil Field Associated Gas with Liquefied Natural Gas Cryogenic Energy Utilization^{*}

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Abstract A novel process to recovery natural gas liquids from oil field associated gas with liquefied natural gas (LNG) cryogenic energy utilization is proposed. Compared to the current electric refrigeration process, the proposed process uses the cryogenic energy of LNG and saves 62.6% of electricity. The proposed process recovers ethane, liquid petroleum gas (propane and butane) and heavier hydrocarbons, with total recovery rate of natural gas liquids up to 96.8%. In this paper, exergy analysis and the energy utilization diagram method (EUD) are used to assess the new process and identify the key operation units with large exergy loss. The results show that exergy efficiency of the new process is 44.3%. Compared to the electric refrigeration process, exergy efficiency of the new process is improved by 16%. The proposed process has been applied and implemented in a conceptual design scheme of the cryogenic energy utilization for a 300 million tons/yr LNG receiving terminal in a northern Chinese harbor.

Keywords liquefied natural gas, utilization of cryogenic energy, recovery of natural gas liquids, exergy analysis

1 INTRODUCTION

Oil field associated gas (OFG) is composed of methane and natural gas liquids (NGL). NGL is mainly composed of ethane, propane, butane, and heavier hydrocarbons. OFG is dissolved in or volatilize from petroleum due to the changing pressure in the process of crude oil exploitation. Recovery of NGL from OFG will upgrade the quality of commodity natural gas, while heavier hydrocarbons recovered are of higher value than methane.

There are currently a number of NGL recovery processes, including adsorption, oil absorption and conditioning. There are two types of conditioning processes. One is operated at a temperature range between $-20\text{ }^{\circ}\text{C}$ and $-35\text{ }^{\circ}\text{C}$, recovers hydrocarbons heavier than ethane; the other is operated below $-45\text{ }^{\circ}\text{C}$, occasionally down to $-100\text{ }^{\circ}\text{C}$, to recover ethane and heavier hydrocarbons [1]. The conditioning processes consume lots of energy, since most of the current gases conditioning refrigeration processes use the electricity.

With the development of the liquefied natural gas (LNG) industry in China, the import of LNG increases sharply. The import of LNG in 2008 was about 3.3 millions tons, from which about 800 million kWh electricity may be recovered, theoretically. Therefore, the cryogenic energy of LNG, as a cleaner energy, becomes more and more attracting the attention from the industry. Cryogenic energy of LNG can be applied to many

fields, such as air separation [2–4], cryogenic energy power generation [5, 6], waste rubber cryogenic comminution [7], and cold storage [8]. Proposed in this paper, OFG from oil fields in northern China is delivered to a nearby LNG-receiving terminal, and NGL is recovered from OFG together with the LNG cryogenic energy.

2 THE RECOVERY PROCESSES

The components of raw OFG are shown in Table 1. The pressure and the temperature of OFG are 0.9 MPa and $50\text{ }^{\circ}\text{C}$, respectively. The capacity of the process is $1.5 \times 10^6\text{ m}^3\cdot\text{d}^{-1}$, equivalent to about 80% turnout of an oil field in northern China. After NGL is separated from OFG, natural gas is input into the urban natural gas pipeline network with the pressure of 3.0 MPa.

The NGL is separated from OFG with a serial process. Methane is firstly isolated from OFG; then, ethane is extracted from OFG; at last, liquefied petroleum gas (LPG) is distilled from OFG.

2.1 The electric refrigeration process

A NGL recovery process from the OFG is reported with electric refrigeration in Ref. [9]. In this paper, the reported NGL recovery process with electric refrigeration is set as the contrast system to the new process with LNG cryogenic utilization.

Table 1 Components of the oil field associated gas

Components	CH ₄	C ₂ H ₆	C ₃ H ₈	<i>i</i> -C ₄ H ₁₀	<i>n</i> -C ₄ H ₁₀	<i>i</i> -C ₅ H ₁₂	<i>n</i> -C ₅ H ₁₂	C ₆ ⁺	CO ₂	N ₂	H ₂ O	Total
mole fraction/%	79.6	8.6	4.1	0.6	1.2	0.4	0.6	2.2	1.6	0.5	0.15	100

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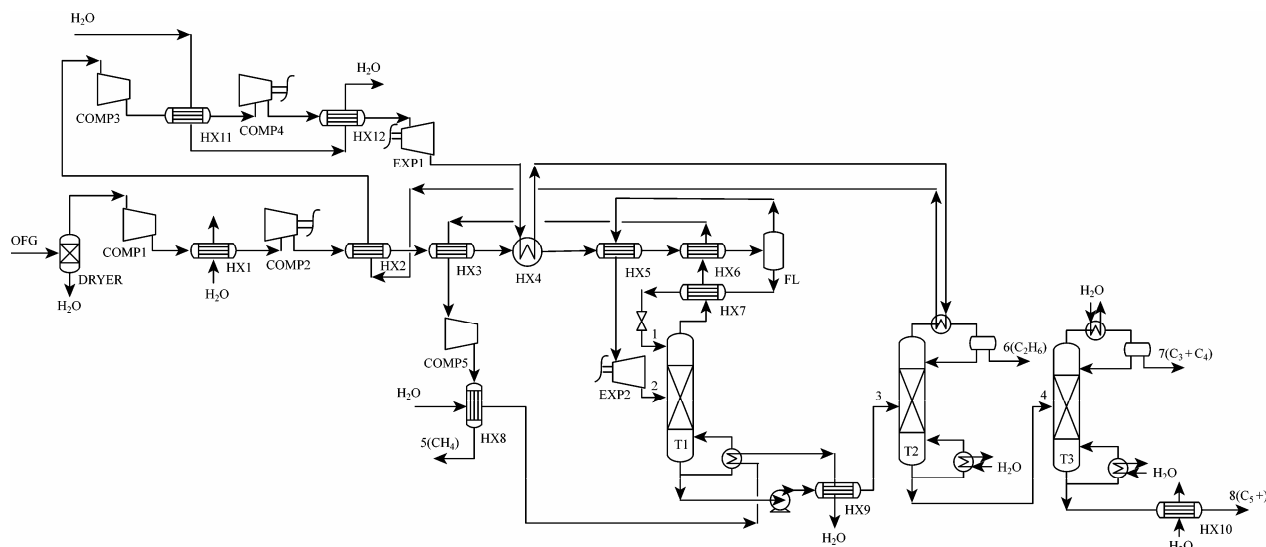


Figure 1 Flow sheet of the process for natural gas liquids recovery from oil field gas with the electric refrigeration
 DRYER—molecular sieve dryer; COMP—compressor; EXP—expander; HX—heat exchanger; FL—flash; T1—de-methane distillation column; T2—de-ethane distillation column; T3—de-butane distillation column

The flowsheet of the process for natural gas liquid recovery from oil field gas with the electric refrigeration is showed in Fig. 1. OFG is introduced into a molecular sieve dryer to separate free water completely. OFG is then fed into the compressor COMP1, compressed to 4.0 MPa. After cooled down to a specified temperature, OFG is fed into another compressor, which is driven up by the expander EXP2, and then compressed to 5.0 MPa. After annealed to $-45\text{ }^{\circ}\text{C}$ by the stream from the electric refrigeration system in the heat exchanger network, OFG is further cooled down to $-60\text{ }^{\circ}\text{C}$. OFG is then fed into a flash, divided into two streams. After its cryogenic energy recovery in the heat exchanger network, the gas stream from the top of the flash is expanded to 1.35 MPa, cooled down to $-100\text{ }^{\circ}\text{C}$. The gas stream is then introduced into a de-methane distillation column at its middle. The liquid stream from the bottom of the flash is cooled down and depressurized to 1.35 MPa, and subsequently fed into the de-methane distillation column from its top. After its cryogenic energy recovery in the heat exchanger network, the product obtained from the top of the de-methane distillation column is compressed to 3.34 MPa and subsequently cooled down to $30\text{ }^{\circ}\text{C}$, then introduced into the urban nature gas pipeline network. The product from the bottom of the de-methane distillation column is pumped to 2.0 MPa and reheated to $31\text{ }^{\circ}\text{C}$, and then fed into the de-ethane distillation column from its middle. Ethane is recovered from the top of the de-ethane distillation column, while the product from its bottom is introduced to the de-butane distillation column. LPG (propane and butane) is exacted from the top the de-butane distillation column, while hydrocarbons heavier than butane are recovered from the bottom of the de-butane distillation column.

2.2 The new process with LNG cryogenic energy utilization

The flowsheet of the proposed NGL recovery process from OFG with LNG cryogenic energy utilization is showed in Fig. 2. OFG is introduced into a molecular sieve dryer to separate free water completely. OFG is then fed into a compressor, compressed to 3.73 MPa. After annealed to $7\text{ }^{\circ}\text{C}$ in the heat exchanger network, OFG is fed into a flash, divided into two streams. The gas stream from the top of the flash is cooled down to $-82\text{ }^{\circ}\text{C}$ and fed into a de-methane distillation column from its middle. The liquid stream from the bottom of the flash is cooled down to $-100\text{ }^{\circ}\text{C}$ and subsequently fed into the de-methane distillation column from its top. The product obtained from the top of the de-methane distillation column is reheated to $20\text{ }^{\circ}\text{C}$, and then introduced into the urban nature gas pipeline network. The product from the bottom of the de-methane distillation column is fed into the de-ethane distillation column. Ethane is recovered from the top of the de-ethane distillation column, while the product from its bottom is introduced to the de-butane distillation column. LPG (propane and butane) is exacted from the top the de-butane distillation column, while hydrocarbons heavier than butane are recovered from the bottom of the de-butane distillation column.

The components of LNG in the proposed process are showed in Table 2. The range of LNG temperature is usually between $-158\text{ }^{\circ}\text{C}$ to $2\text{ }^{\circ}\text{C}$ in a LNG receiving terminal. Thus, the input temperature is set at $-158\text{ }^{\circ}\text{C}$, while the output temperature is set at $2\text{ }^{\circ}\text{C}$. The LNG flowrate is set as $38.2\text{ t}\cdot\text{h}^{-1}$, equivalent to about 10% of the throughput of a LNG-receiving terminal

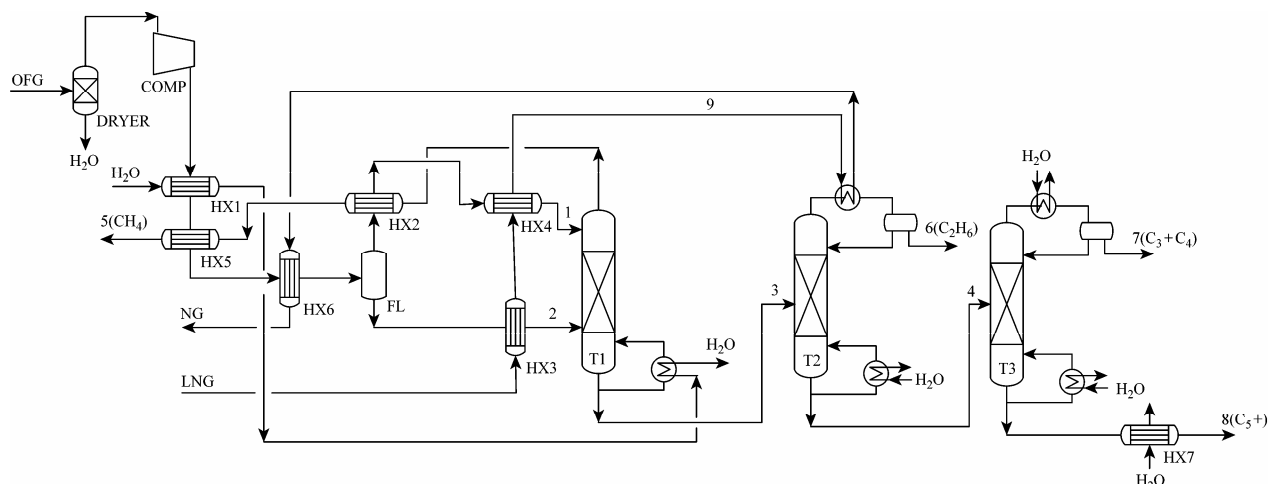


Figure 2 Flowsheet of the proposed NGL recovery process from OFG with LNG cryogenic energy utilization
 DRYER—molecular sieve dryers; COMP—compressor; HX—heat exchanger; FL—flash; T1—de-methane distillation column; T2—de-ethane distillation column; T3—de-butane distillation column

Table 2 Components of liquefied natural gas

Components	CH ₄	C ₂ H ₆	C ₃ H ₈	<i>i</i> -C ₄ H ₁₀	<i>n</i> -C ₄ H ₁₀	CO ₂	N ₂	Total
mole fraction/%	89.39	5.76	3.30	0.78	0.66	0.0	0.11	100

Table 3 Main parameters specification of distillation columns in the electric refrigeration process

Unit	Number of stages	Feed plate	Temperature/°C	Pressure/MPa	Distillate rate/kmol·h ⁻¹	Reflux ratio
de-methane distillation column	31	1, 12	-100, -97.0	1.25	2310	0.3
de-ethane distillation column	32	17	31.0	1.85	235.6	1.4
de-butane distillation column	42	16	83.4	1.2	151.1	1.2

Table 4 Main parameters specification of distillation columns in the new process

Unit	Number of stages	Feed plate	Temperature/°C	Pressure/MPa	Distillate rate/kmol·h ⁻¹	Reflux ratio
de-methane distillation column	31	1, 12	-100, -82.0	3.4	2274	0.12
de-ethane distillation column	32	17	31.0	1.85	259.1	1.2
de-butane distillation column	42	16	83.4	1.2	147.8	0.91

Table 5 Products of the process in the electric refrigeration process

Products	Temperature/°C	Pressure/MPa	Flow/kg·h ⁻¹	Mole fraction/%	Recovery efficiency/%
ethane	-14.8	1.85	7600	84.3	82.5
LPG	46.5	1.17	7425	—	91.1
light hydrocarbon	38.0	1.41	7184	—	98.1
natural gas	30.0	3.34	38494	96.5	100
NGL					90.3

Table 6 Products of the process in the new process

Products	Temperature/°C	Pressure/MPa	Flow/kg·h ⁻¹	Mole fraction/%	Recovery efficiency/%
ethane	-18.6	1.85	8266	86.8	93.4
LPG	44.0	1.17	7835	—	97.3
light hydrocarbon	38.0	1.41	7305	—	99.5
natural gas	20.2	3.34	37298	98.0	100
NGL					96.8

Table 7 Compositions and exergy values of key streams in the electric refrigeration process

Stream	Flow rate×10 ⁻³ /kg·h ⁻¹	T /°C	P /MPa	Composition/% (by mol)										Enthalpy /kJ·kg ⁻¹	Entropy /kJ·kg ⁻¹ ·K ⁻¹	Exergy×10 ⁻⁴ /kJ·h ⁻¹	
				CH ₄	C ₂ H ₆	C ₃ H ₈	i-C ₄ H ₁₀	n-C ₄ H ₁₀	i-C ₅ H ₁₂	n-C ₅ H ₁₂	C ₆ ⁺	CO ₂	N ₂				H ₂ O
OFG	61.01	50.0	0.9	79.6	8.6	4.1	0.6	1.2	0.4	0.6	2.2	1.6	0.5	0.15	-3911	-5.60	1806
1	34.44	-97.0	1.35	60.6	15.7	9.2	1.4	2.9	1.0	1.5	5.4	2.3	2.0	—	-3895	-8.19	1125
2	28.26	-100	1.35	93.6	3.8	0.6	376×10 ⁻⁶	535×10 ⁻⁶	70×10 ⁻⁶	86×10 ⁻⁶	88×10 ⁻⁶	1.2	0.7	—	-4873	-7.41	1338
3	22.21	31.0	2.0	—	41.9	23.7	3.5	7.1	2.4	3.5	13.0	4.9	—	—	-2935	-6.98	239.5
4	14.61	87.0	1.4	—	14×10 ⁻⁶	41.4	7.0	14.1	4.7	7.0	25.8	—	—	—	-2299	-6.84	76.9
5	38.50	30.0	3.34	96.5	1.8	977×10 ⁻⁶	24×10 ⁻⁶	26×10 ⁻⁶	1×10 ⁻⁶	1×10 ⁻⁶	404×10 ⁻⁶	0.9	0.6	—	-4663	-6.59	1970
6	7.60	-14.8	1.85	—	84.3	5.9	24×10 ⁻⁶	3×10 ⁻⁶	—	—	—	0.98	—	—	-4015	-7.22	198.7
7	7.43	46.5	1.17	—	22×10 ⁻⁶	65.4	11.1	22.2	1.1	0.3	27×10 ⁻⁶	—	—	—	-2613	-7.39	65.3
8	7.18	38.0	1.44	—	—	—	258×10 ⁻⁶	21×10 ⁻⁶	10.9	18.7	70.3	—	—	—	-2305	7.41	1.91

Table 8 Compositions and exergy values of key streams in the new process

Stream	Flow rate×10 ⁻³ /kg·h ⁻¹	T /°C	P /MPa	Composition/% (by mol)											Enthalpy /kJ·kg ⁻¹	Entropy /kJ·kg ⁻¹ ·K ⁻¹	Exergy×10 ⁻⁴ /kJ·h ⁻¹
				CH ₄	C ₂ H ₆	C ₃ H ₈	i-C ₄ H ₁₀	n-C ₄ H ₁₀	i-C ₅ H ₁₂	n-C ₅ H ₁₂	C ₆ ⁺	CO ₂	N ₂	H ₂ O			
LNG	38.20	-158	6.2	89.39	5.76	3.3	0.78	0.66	—	—	—	—	0.11	—	-5083	-11.18	3607
1	51.37	-82.0	3.58	84.0	8.5	3.5	0.4	0.7	0.2	0.2	0.3	1.7	0.5	—	-3754	-8.57	2215
2	9.334	-100	3.58	17.4	10.4	13.5	3.7	9.2	4.4	7.2	33.3	0.9	0.04	—	-2887	-8.49	122.6
3	23.41	31.0	2.0	—	44.2	21.8	3.2	6.5	2.2	3.3	12.1	6.6	—	—	-3053	-6.80	271.9
4	15.14	83.4	1.4	—	0.3	44.2	6.6	13.2	4.4	6.7	24.5	—	—	—	-2315	-6.87	82.14
5	37.30	20.0	3.34	98.0	0.7	0.2	113×10 ⁻⁶	150×10 ⁻⁶	21×10 ⁻⁶	17×10 ⁻⁶	0.5	0.6	—	—	-4686	-6.75	1940
6	8.266	-18.6	1.85	546×10 ⁻⁶	86.8	991×10 ⁻⁶	—	—	—	—	—	13.1	—	—	-4360	-7.01	226.9
7	7.835	44.0	1.17	—	0.5	68.6	10.2	20.4	0.2	701×10 ⁻⁶	2×10 ⁻⁶	19×10 ⁻⁶	—	—	-2622	-7.43	72.31
8	7.305	38.0	1.41	—	—	—	15×10 ⁻⁶	0.2	12.2	18.6	69.0	—	—	—	-2317	-7.42	2.106
9	38.20	-57.8	6.14	89.39	5.76	3.3	0.78	0.66	—	—	—	—	0.11	—	-4456	-7.77	2429

in a northern China of $3 \times 10^6 \text{ t} \cdot \text{a}^{-1}$ capacity typically.

The proposed process is an electricity-saving and alternative process to the electric refrigeration process. The process has been applied in a conceptual design scheme of LNG cryogenic energy utilization, and implemented in a $300 \times 10^6 \text{ t} \cdot \text{a}^{-1}$ LNG receiving terminal in a northern Chinese harbor.

Since OFG and raw natural gas are basically same composition, this new process can also be applied to recovery of the liquids from the raw natural gas.

3 SIMULATIONS OF THE PROCESSES

Flowsheet simulations are implemented with the commercial Aspen Plus software [10]. To simplify the simulations, it is assumed that the systems are steady state flowing. As most of the inputs involved in the two processes are hydrocarbons, PR equation is chosen as the model for the thermodynamic properties and phase equilibrium computation in this paper. In the simulations, it is assumed that the isentropic efficiency of the compressor is 80% and its mechanical efficiency 90%. Heat loss in the heat exchangers is ignored, while the specific pressure drop in one heat exchanger is 0.03 MPa. The operation conditions of the distillation columns of the electric refrigeration process and the new process are illustrated in Tables 3 and 4, respectively. In addition, there is no condenser on the top of the de-methane distillation column, while there is a reboiler at its bottom in two processes. The statuses of the products generated in two processes are showed in Tables 5 and 6, respectively. The NGL recovery rate in the electric refrigeration process is 90.3%, and the NGL recovery rate in the new process is 96.8%.

Since the operation temperatures of de-methane column and de-ethane column are low, the cryogenic energy of their products should be reused through the integration of the heat exchanger network. To improve the energy efficiency of the process and reduce the consumption of utilities, the pinch technique is used to integrate the heat exchanger networks of the two processes. It is assumed that the minimum temperature difference in the heat exchangers networks is 3°C .

The temperature difference between the cold stream and the hot stream in the heat exchanger 2 is showed in Fig. 3, in which the minimum temperature difference is 3.2°C , and the temperature of the cold

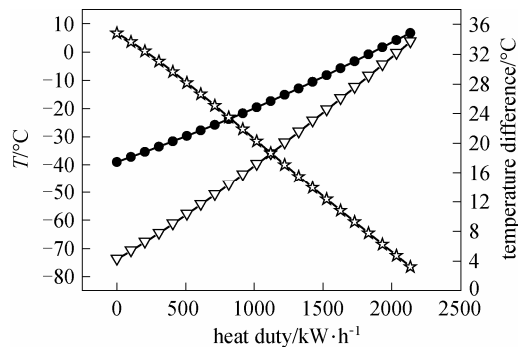


Figure 3 The temperature difference between the cold stream and the hot stream in heat exchanger HX2

▽ CH₄; ● OFG; ☆ temperature difference

and hot streams is 3.6°C and 6.8°C , respectively. And the simulation results show that the minimum temperature differences of Heat exchanger 3 and Heat exchanger 4 are 57°C and 10.9°C , respectively.

In the electric refrigeration process, the mixture of ethane and propane is chose as the refrigerant, whose molar composition of ethane is 0.5. The refrigeration cycle temperature is set as -50°C . The compositions of key streams in the electric refrigeration process are showed in Table 7.

In the new process with LNG cryogenic energy utilization, LNG provides the cryogenic energy for OFG conditioning, while electricity is used to compress OFG to meet the specific pressure. The compositions of key streams are showed in Table 8.

4 THERMODYNAMIC ANALYSIS

4.1 Electricity-saving efficiency analysis

The comparison of energy usage between the electric refrigeration process and the proposed process is given in Fig. 4. All of the electricity consumption in the electric refrigeration process is set as W_{OE} , which includes four electricity consumptions, namely the net work required of OFG compressor COMP1, W_1 ; the electricity required of the electric refrigeration system COMP3, W_2 ; the net work required of the product compressor COMP5, W_3 ; and the work of the pump P1, W_4 . In the new process, there is only one type of electricity consumption, namely the net work required

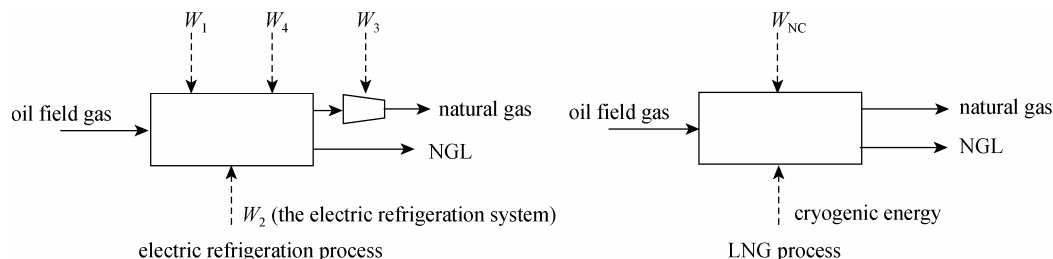


Figure 4 Energy comparison between the electric refrigeration process and the LNG process

of OFG compressor COMP, W_{NC} . Electricity-saving efficiency of the new process can be formulated as follows:

$$\eta_E = \frac{W_{\text{OE}} - W_{\text{NC}}}{W_{\text{OE}}} \quad (1)$$

Following analysis is based on the case of $1.5 \times 10^6 \text{ m}^3 \cdot \text{d}^{-1}$ OFG treatment capacity. The simulation results show that in the electric refrigeration process, W_1 , W_2 , W_3 , and W_4 are 4304 kW, 5398 kW, 2377 kW and 11 kW, respectively. In the new process, the net work required of OFG compressor COMP, W_{NC} , is 4526 kW. Compared to the electric refrigeration process, the new process can save 7564 kW electricity, and the electricity-saving efficiency of the new process is 62.6%.

4.2 Exergy analysis

Exergy can be used to account for the irreversibility of a process, and provide a more detailed tracking mechanism for energy usage. The exergy of a material stream is defined by Kotas as the maximum work available, when the stream is brought from its initial state to the final state by a process during which the stream may interact only with the environment [11]. Exergy can be considered as being composed of two components: physical exergy and chemical exergy [12]:

$$E_X = E_{X,\text{phys}} + E_{X,\text{chem}} \quad (2)$$

where E_X is the total exergy, $E_{X,\text{phys}}$ the physical exergy, and $E_{X,\text{chem}}$ the chemical exergy, respectively.

In a steady state flowing and open system, it is assumed that potential exergy and kinetic exergy are ignored, then the exergy of the stream is calculated by

$$E_X = \Delta H - T_0 \Delta S = H - H_0 - T_0 (S - S_0) \quad (3)$$

where ΔH is enthalpy change, ΔS is entropy change,

T_0 is the reference temperature 298.15 K, H_0 and S_0 are enthalpy and entropy at the reference temperature, respectively.

As all the streams in two processes have only physical change and no chemical changes involved, and Eq. (3) applies. The exergy values of part of key streams in the electric refrigeration process are shown in Table 7, and the exergy values of part of key streams in the new process are shown in Table 8.

Gray box model is a network model of black box units, which is chosen to assess the two processes. In the gray box model, each operation unit is assumed to be a black box unit, which is connected with others by the exergy flow. For black box model, the input total exergy of one system is equal to the sum of output total exergy of the system:

$$E_{X,\text{in}} - E_{X,\text{out}} - E_{X,\text{loss}} = 0 \quad (4)$$

where $E_{X,\text{in}}$ is the total input exergy; the total output exergy includes two forms of exergy, namely, the effective exergy $E_{X,\text{out}}$ and the exergy loss $E_{X,\text{loss}}$. The exergy efficiency is calculated as a ratio of the effective exergy, $E_{X,\text{out}}$, to the total input exergy, $E_{X,\text{in}}$:

$$\eta = \frac{E_{X,\text{out}}}{E_{X,\text{in}}} \quad (5)$$

Correspondingly, the exergy loss rate is calculated as

$$\xi = 1 - \eta = \frac{E_{X,\text{loss}}}{E_{X,\text{in}}} \quad (6)$$

Exergy efficiencies of key units of the electric refrigeration process are shown in Table 9, and exergy efficiencies of key units of the new process are shown in Table 10.

From Table 9, the exergy efficiency of Heat exchanger 2 and Heat exchanger 10 are lower than the other operation units, 25.3% and 19.6%, respectively. Temperature differences between hot and cool streams

Table 9 Exergy efficiency of operation units in the electricity refrigeration process

Unit	Efficiency of exergy (η)/%	Unit	Efficiency of exergy (η)/%
Compressor 1	86.2	Heat exchanger 5	89.3
Compressor 2	82.7	Heat exchanger 6	76.1
Compressor 3	86.5	Heat exchanger 7	62.6
Compressor 4	84.8	Heat exchanger 8	96.5
Compressor 5	78.2	Heat exchanger 9	96.7
Expander 1	73.1	Heat exchanger 10	19.6
Expander 2	68.2	Heat exchanger 11	75.2
Molecular sieve dryer	99.4	Heat exchanger 12	83.2
Heat exchanger 1	91.3	flash	97.6
Heat exchanger 2	25.3	de-methane distillation column	78.2
Heat exchanger 3	93.4	de-ethane distillation column	67.3
Heat exchanger 4	56.1	de-butane distillation column	63.2
		the whole process	38.2

Table 10 Exergy efficiency of operation units in the new process

Unit	Efficiency of exergy (η)/%	Unit	Efficiency of exergy (η)/%
compressor	80.5	Heat exchanger 6	96.7
molecular sieve dryer	99.4	flash	97.3
Heat exchanger 1	92.1	de-methane distillation column	86.2
Heat exchanger 2	97.7	de-ethane distillation column	85.5
Heat exchanger 3	94.1	de-butane distillation column	53.1
Heat exchanger 4	94.0	Heat exchanger 7	29.3
Heat exchanger 5	98.9	the whole process	44.3

in those two heat exchangers are 102 °C and 137 °C, respectively. Large temperature differences cause lower exergy efficiencies.

The results in Table 10 show that the exergy efficiency of the de-butane distillation column and Heat exchanger 7 are lower than other operation units, 53.1% and 29.3%, respectively, because of the large temperature differences between hot and cool streams of the de-butane distillation column and Heat exchanger 7.

The exergy efficiency of the new process is 44.3% in Table 10, and the exergy efficiency of the electric refrigeration process is 38.2% in Table 9. Compared to the electric refrigeration process, the new process has a higher efficiency of energy utilization, and exergy efficiency of the new process is improved by 16%.

4.3 Energy utilization diagram analysis

The conventional exergy analysis is based on the exergy balance calculation. The energy utilization diagram (EUD) analysis [13,14] is a better approach to analyze the process energy usage efficiency, which was proposed by Ishida in 1982, and can reveal fundamental mechanism of the exergy loss in the operation units. EUD analysis could provide a way to improve the performance of the process.

In EUD, the energy transfer process consists of an energy donor and an energy acceptor. The energy level, A , a new intensive parameter, is introduced in the EUD, which is a ratio of the exergy change, ΔE , to the enthalpy change, ΔH , in the process:

$$A = \Delta E / \Delta H \quad (7)$$

The energy level difference between the energy donor and the energy acceptor drives the process. The area between the energy donor curve and the energy acceptor curve in the EUD represents the exergy loss of the process.

In a steady state system, the exergy change can be calculated with

$$\Delta E = \Delta H - T_0 \Delta S \quad (8)$$

Substituting Eq. (8) into Eq. (7), the energy level is

$$A = 1 - T_0 \Delta S / \Delta H \quad (9)$$

For the heat exchanger subsystem, ΔH is the heat duty Q , and ΔS is Q/T . Eq. (9) changes to

$$A = \Delta E / \Delta H = 1 - T_0 / T \quad (10)$$

In the power subsystem, ΔH is equal to W and $\Delta S = 0$, Eq. (9) becomes

$$A = \Delta E / \Delta H = 1 - T_0 \Delta S / \Delta H = 1 \quad (11)$$

This means that the energy level of the energy donor in the compressor is always equal to one; its energy level of the energy acceptor can be calculated by Eqs. (7) and (8).

Energy level is a function of the temperature in the heat exchange process, as shown in Eq. (10). Therefore, the energy levels are calculated with the temperatures of the cool stream and the hot stream in the heat exchanger. From the results of the process simulation, we can obtain the temperature of the cool stream, the hot stream, and the corresponding heat duty of the temperature change, and then the energy levels of the cool stream and the hot stream can be calculated from Eq. (10). As most of the exergy loss occurs in condensers and reboilers, a distillation process can be approximate to a heat exchange process, in terms of energy utilization. The EUD of the proposed process is shown in Fig. 5, in which the energy levels are Y-axis, while the corresponding heat duty is X-axis.

Exergy utilization diagram analysis of the new process could be made in details from Fig. 5 with separation of different operation units as noted. Exergy loss of each unit can be calculated from the area confined within each pair of curves, the results are shown in Table 11.

As shown in Fig. 5 and Table 11, the exergy loss of the OFG compressor is 0.878 million kW, 16.4% of the total exergy loss, much lower than the exergy loss of the heat exchanger and distillation subsystem.

Heat exchanger 4 and the de-ethane distillation column are the units with the largest exergy loss in the heat exchange subsystem from Fig. 5 and Table 11.

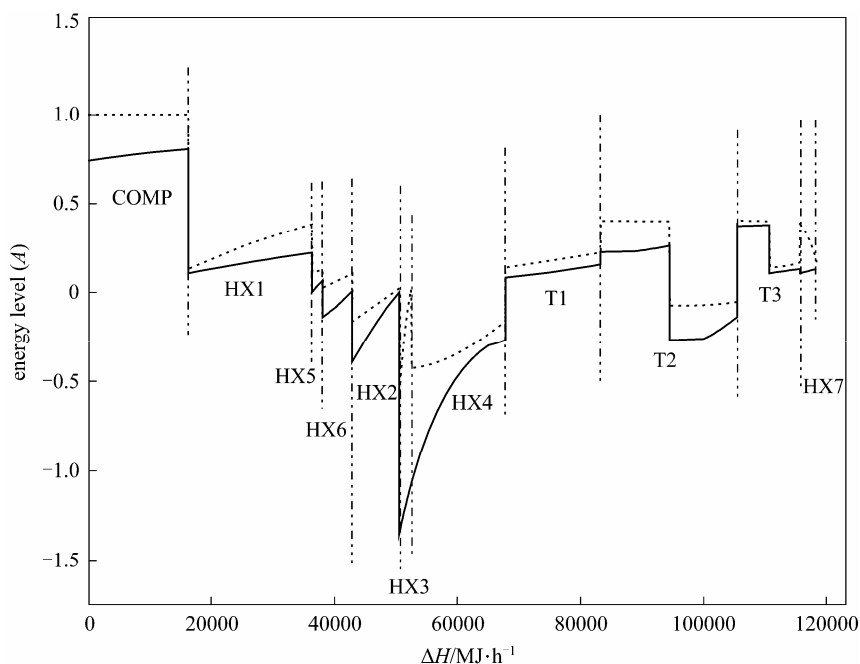


Figure 5 EUD diagram of the proposed process

COMP—compressor; HX—heat exchanger; T1—de-methane distillation column; T2—de-ethane distillation column; T3—de-butane distillation column; — acceptor; donor

Table 11 Exergy loss of heat exchangers and distillation columns in the new process

Item	Exergy loss $\times 10^6/\text{kW}$	$\xi/\%$	Curves	Fraction of total exergy loss/%
Total exergy loss	5.355	—	—	100
Compressor	0.878	19.5	COMP	16.4
Heat exchanger and distillation subsystem	4.477	—	—	83.6
Heat exchanger 1	0.622	7.9	HX1	11.6
Heat exchanger 2	0.283	2.3	HX2	5.3
Heat exchanger 3	0.603	5.9	HX3	11.3
Heat exchanger 4	0.953	6.0	HX4	17.8
Heat exchanger 5	0.069	1.1	HX5	1.4
Heat exchanger 6	0.216	3.3	HX6	4.0
Heat exchanger 7	0.126	70.7	HX7	2.3
De-methane distillation column	0.249	13.8	T1	4.6
De-ethane distillation column	1.039	14.5	T2	19.4
De-butane distillation column	0.317	46.9	T3	5.9

For Heat exchanger 4, the exergy loss is due to the large temperature difference between the cool and the hot streams. According to Fig. 5, the energy level of the energy donor in Heat exchanger 4 is -0.43 , which is 0.6 higher than the energy level of the corresponding acceptor. Difference between energy levels of the donor and the acceptor causes the large exergy loss in

Heat exchanger 4.

In the case of the de-ethane distillation column, its exergy loss is large due to the large temperature difference between the cooling water and the cooled material in the condenser, and between the hot steam and the heated material in the reboiler. According to Fig. 5, the energy levels of the energy donors in the

condenser and reboiler are -0.27 and 0.23 , respectively, which are 0.3 and 0.2 higher than the energy level of the corresponding acceptors, respectively. Difference between energy levels of the donors and the acceptors causes the large exergy loss in the de-ethane distillation column.

Exergy loss rate of units is calculated from Eq. (6), as shown in Table 11. While exergy loss rate of Heat exchanger 4 and the de-ethane distillation column are lower, 6.0% and 14.5% , respectively, their exergy losses are 17.8% and 19.4% of the total exergy losses of the process, respectively, much larger than the others, as shown in Table 11. Large input exergy in these two units is the main reason to cause large exergy loss. These two units should be optimized to reduce their exergy losses.

Also shown in Table 11, although exergy loss rates of the de-butane distillation column and Heat exchanger 11 are high, their exergy losses are lower than most of the others. It is due to small input exergy is one of the main reasons to cause low exergy losses with high exergy loss rates. These two units also should be optimized to improve the individual and the total exergy efficiencies. The analysis results show that Heat exchangers 4 and 7, the de-ethane distillation column, and the de-butane distillation column are the key units, leaving rooms for further improvement.

5 TECHNO-ECONOMIC ANALYSIS

In the process, OFG is raw material. The cryogenic energy of LNG and electricity are energy consumption; the products are the natural gas, ethane, LPG and heavy hydrocarbons, respectively. The entire price is determined with the cost of electricity and water in specific location of the industrial implementation. Price of the cryogenic energy is set as $100 \text{ CNY} \cdot \text{t}^{-1}$ LNG, the prices of the materials and products in the process are showed in Table 12.

Table 12 Prices of feeds and products in the new process

	Item	Price	Unit
feed	oil field associated gas	750	$\text{CNY} \cdot \text{km}^{-3}$
	LNG	100	$\text{CNY} \cdot \text{t}^{-1}$
	power	0.50	$\text{CNY} \cdot \text{kW}^{-1} \cdot \text{h}^{-1}$
	water	3.1	$\text{CNY} \cdot \text{t}^{-1}$
product	natural gas	980	$\text{CNY} \cdot \text{km}^{-3}$
	ethane	3760	$\text{CNY} \cdot \text{t}^{-1}$
	LPG	3560	$\text{CNY} \cdot \text{t}^{-1}$
	light hydrocarbon	3010	$\text{CNY} \cdot \text{t}^{-1}$

Total investment of the process, in the case of $1.5 \times 10^6 \text{ m}^3 \cdot \text{d}^{-1}$ OFG capacity, is 497 millions CNY.

The cost of the equipments purchase and installation is 360 millions CNY, which includes 36 km pipeline investment of OFG. The life time of the process is 16 years, and the construction time is one year. After initialization, the production load is 80% of the full production load in the first year, and reaches the full production load in the second year afterward. In this paper, it is assumed that the benchmark rate of return is 14% , loan interest rate is 6.723% , and the depreciable life of the fixed assets is 14 years with the straight-line depreciation. After calculation, the internal rate of return of the new process is 93.3% , and its payback period is 2.25 years.

As OFG is less expensive than ethane, LPG, and heavy hydrocarbon, the new process is economically promising.

6 CONCLUSIONS

A novel LNG cryogenic energy utilization process for natural gas liquids recovery from oil field associated gas was proposed, in which ethane, LPG (propane and butane) and heavier hydrocarbons are recovered, respectively. The recovery rate of NGL is 96.8% . Compared to the electric refrigeration process, the proposed process saved 62.6% electricity.

The exergy analysis of the proposed process was carried out. In addition, a method of the energy utilization diagram was used to identify the operation units with large exergy loss. The analysis results showed that the exergy efficiency of the process is 44.3% ; Compared to the electric refrigeration process, exergy efficiency of the new process is improved by 16% . Exergy efficiencies of the de-butane distillation column and heat exchanger 7 were low, and the exergy losses were large in heat exchanger 4 and the de-ethane distillation column. The exergy analysis and the EUD analysis had identified the units with low exergy efficiency and the units with large exergy loss, the results could then be used to further optimize the entire process.

Techno-economic analysis of the proposed process was carried out. The results showed that total investment of the process, in the case of $1.5 \times 10^6 \text{ m}^3 \cdot \text{d}^{-1}$ OFG capacity, was 497 millions CNY. Its internal rate of return was 93.3% , and its payback period was 2.25 years.

NOMENCLATURE

A	energy level
E_x	exergy, $\text{kJ} \cdot \text{h}^{-1}$
H	enthalpy, $\text{kJ} \cdot \text{kg}^{-1}$
T	temperature, K
S	entropy, $\text{kJ} \cdot \text{kg}^{-1} \cdot \text{K}^{-1}$
W	power input, kW
η	exergy efficiency
η_e	electricity-saving efficiency
ξ	exergy loss rate

REFERENCES

- 1 Yudong, W., Principles and Processes of Natural Gas Treatment, China Petrochemical Press, Beijing (2007). (in Chinese)
- 2 Yamanouchi, N., Nagasawa, H., "Using LNG cold for air separation", *Chem. Eng. Progr.*, **75**, 78–82 (1979).
- 3 Nakaiwa, M., Akiya, T., Owa, M., Tanakay, Y. "Evaluation of an energy supply system with air separation", *Energy Convers. Manage.*, **37**, 295–301 (1996).
- 4 Chen, Z.S., Chen, W.L., "A new air separation system by using cold energy of LNG", *J. Eng. Thermophys.*, **25**, 913–916 (2004). (in Chinese)
- 5 Dispenza, C., Dispenza, G., La Rocca, V., Panno, G., "Exergy recovery during LNG regasification: Electric energy production—Part one", *App. Therm. Eng.*, **29**, 380–387 (2009).
- 6 Zhang, N., Lior, N., "A novel near-zero CO₂ emission thermal cycle with LNG cryogenic exergy utilization", *Energy*, **31**, 1666–1679 (2006).
- 7 Xiong, Y.Q., Hua, B., Jia, D.M., "Process of cold energy utilization of LNG satellite station in waste rubber cryogenic comminution", *Moden Chemistry Industry*, **29**, 59–62 (2009). (in Chinese)
- 8 Wang, Q., Li, Y.Z., Chen, X., Lv, X.G., "Analysis on cold energy of liquefied natural gas and its recovering and utilization", *Fluid Machinery*, **31**, 56–58 (2003). (in Chinese)
- 9 Wang, X.K., Zhang, H., Yan, S.R., Wang, D.F., "Big NGL recovering plant with advanced cryogenic technology", *Natural Gas Industry*, **23**, 133–35 (2003). (in Chinese)
- 10 Aspen Technology, Inc., Aspen plus version 11.1, 2009, <http://www.aspentech.com>.
- 11 Kotas, T., The Exergy Method of Thermal Plant Analysis, Butterworths, London (1985).
- 12 Hinderink, A., Kerkhof, F., "Exergy analysis with a flow sheeting simulator-I. Theory calculating exergies of material streams", *Chem. Eng. Sci.*, **51**, 4693–4700 (1996).
- 13 Jin, H.G., Ishida, M., "Graphical exergy analysis of complex cycle", *Energy*, **18**, 615–25 (1993).
- 14 Srinophakun, T., Laowithayangkul, S., Ishida, M., "Simulation of power cycle with energy utilization diagram", *Energy Convers. Manage.*, **42**, 1437–1456 (2001).