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# Oiling low temperature separation process for dehydration and de-hydrocarbon of natural gas and practical application

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**Abstract.** In order to reach the specification of water and hydrocarbon dew point, most of the water and $C_3$ and heavier hydrocarbons need to be removed from natural gas. Low Temperature Separation (LTS) process by throttling expansion or refrigeration is commonly employed for natural gas dehydration and de-hydrocarbon as the most economical method. In case of some special natural gas which contains small quantity of the wax (mainly $C_{16+}$ components), wax is easily deposited in the throttling facilities that leads the blockage of internals of low temperature separator and affects safety operation. Oiling LTS process, a novel process for this kind of special natural gas, is investigated in this work. The oil is injected into the natural gas to reduce the wax precipitation temperature and thus to avoid wax blockage. It is a kind of improvement and innovation process based on conventional treatment ones. According to the simulation analysis and practical application for a natural gas processing plant in Aksu, China, the new process presents good performance. Also, it is found that the best injection point is at the inlet pipe of raw gas pre-cooler and the optimal injecting stream is from the three-phase separator. Also, the injection oil flowrate is determined by refrigerating temperature and the concentration of wax. Meanwhile, the detailed economic analysis of oiling LTS process was involved. Comparing the conventional process, its adding investment is around 423,583$ accounting on 5% total capital cost of this unit. However, from the comprehensive aspects, this optimized process presents a good economic benefit.

**Keywords:** Refrigerating temperature, Low Temperature Separation (LTS), Wax precipitation temperature, Wax blockage, Oiling.

## 1 Introduction

Natural gas is a non-renewable source which is used in a wide range of applications because of its clean-combustible and chemical characteristics, for example the generation of electricity and the obtainment of petrochemical products [1, 2]. Natural gas is composed of hydrocarbons, mainly including methane, ethane and a small percentage of propane, butane, and also, it includes some impurities such as water vapor, sulfides, carbon dioxide, nitrogen, or heavier hydrocarbons [3, 4]. Its specific composition varies according to the site where it has been obtained [5].

The composition of sales gas is restricted to accomplish the adequate conditions to be transported and used [6, 7]. These restrictions are ruled by the local entities in each country or the customers' requirements [8]. Different processes are necessary for cleaning of the raw natural gas to meet the specification of the pipeline (Pipeline-Quality Natural Gas) and for environmentally clean-burning gas. Totally, gas processing is applied to achieve the following goals [9, 10]:

1. The purified natural gas meets the utilization requirement as fuel (residential or industrial).
2. The valuable components separated from the raw gas can be used as petrochemical feed stocks, fuels (*e.g.*, propane), or industrial gases (*e.g.*, ethane, helium).
3. Liquefaction of the natural gas to be transported or stored.

In the natural Gas Processing Plant (GPP), the unit operations will be dependent on the gas composition, the type of facility, and the product specifications [11]. With the aim of satisfying these bound values, the natural gas should be treated in different process sectors. One of the sectors is the dehydration and de-hydrocarbon stage, where the water and heavy hydrocarbons are removed to meet the requirement of hydrocarbon and water dew point [12, 13].

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**Table 1.** The specification of hydrocarbon dew point of international organization and countries.

| No. | Organization or country | Specification of hydrocarbon |
|---|---|---|
| 1 | ISO | No liquid hydrocarbon at the handing temperature and pressure |
| 2 | EASEE-Gas | −2 °C at 1–70 bar |
| 3 | Austria | −5 °C at 40 bar |
| 4 | Belgium | −3 °C at 69 bar |
| 5 | Canada | −10 °C at 54 bar |
| 6 | Italy | −10 °C at 60 bar |
| 7 | Holland | −3 °C at 70 bar |
| 8 | UK | Summer: 10 °C at 70 bar |
| | | Winter: −1 °C at 69 bar |
| 9 | Russia | Warm zone: 0 °C |
| | | Cold zone: −5 °C in summer; −10 °C in winter |
| 10 | China | No liquid hydrocarbon at the transportation temperature and pressure |

Note: *EASEE-Gas – European Association for the Streamlining of Energy Exchange-Gas.* 1 bar = $10^5$ Pa = 0.1 MPa.

Normally, the GPP considers the dew point control unit to treat the natural gas from the upstream [14] for implement of this goal of remove the water, heavy hydrocarbons and recovery the Natural Gas Liquid (NGL).

The hydrocarbon dew point is varied in different countries and regions [15, 16]. The specification of hydrocarbon dew point of international organization and countries is shown in Table 1. The water dew point of sales natural gas is that the water dew point shall be 5 °C lower than the lowest environment temperature during transportation of the sales gas.

Generally, there are two conventional processes for natural gas dehydration and de-hydrocarbon, namely low temperature separation and molecular sieve adsorption respectively. They both can remove the water and heavy hydrocarbons simultaneously, of which the Low Temperature Separation (LTS) is most frequently used due to its good process performance and economic benefit [17, 18]. Figure 1 shows the conventional LTS process, which mainly includes pre-cooler, high pressure separator, after-cooler, refrigeration device (such as Joule-Thomson (J-T) valve, expander, and propane refrigeration system etc.), and low temperature separator.

The refrigerating temperature depends on the properties of natural gas, especially the components [19, 20]. The condensing temperature of light components is quite low, while the heavier components can be condensed successively with the refrigerating temperature decrease. Obviously, the refrigerating temperature is determined by the specification of dew point and components of raw natural gas [21–23].

The raw gas enters the pre-cooler to exchange heat with cold product gas. After cooling, the raw gas is sent to the high-pressure separator to separate preliminarily the gas/liquid. The gas from the top of separator is cooled through the after-cooler where the glycol as the gas hydrate inhibitor is injected into the raw gas for reducing the gas hydrate forming temperature. After that, the gas flows into the refrigeration device to decrease the temperature further, where most of the heavier components and saturated water are liquefied from the gas phase. And then, the cold raw gas enters into the low temperature separator where the most of the liquid drops with the diameter no less than 5 μm, are removed by high efficient internals. The product gas from the top of low temperature separator returns to the after-cooler and pre-cooler successively for recovering the cold energy. Afterwards, the product gas is transported to gas grid as the sales gas [24]. The liquid stream from the bottom of the low temperature separator and high-pressure separator is mixed together and flows into the downstream unit [25–27].

The key points of this process are the J-T throttling expansion valve for decreasing the temperature and the low temperature separator with high-efficiency internals, which are easily blocked during actual operation. Thus the anti-blocking operation shall be considered in this unit. The hydrate inhibitor is employed to reduce the hydrate forming temperature for preventing the hydrate blockage. However, for some special gas containing a small quantity of wax, the heavy hydrocarbons are apt to deposit as solid from gas phase directly with the temperature decreasing sharply, which can cause the blockage in the low temperature separation system. This will pose a threat not only to the disqualification of the product gas, but also for security operation and emergency shutdown of the whole plant.

## 2 Practical operation of GPP

### 2.1 The GPP process description

Dabei GPP located in Aksu, Xinjiang province, China [28], consists of one gas gathering unit with maximum total capacity of 1 BCM (billion cubic meters) per year, three identical dew point control units with 0.5 BCM per year matching with the gas gathering unit, two condensate stabilization units with 252 tons per day and two glycol regeneration units. The overall process flow diagram is shown in Figure 2 and the parameters of the raw gas are shown in Table 2. The GPP employed the LTS process to treat raw gas which is called sweet gas as it contains

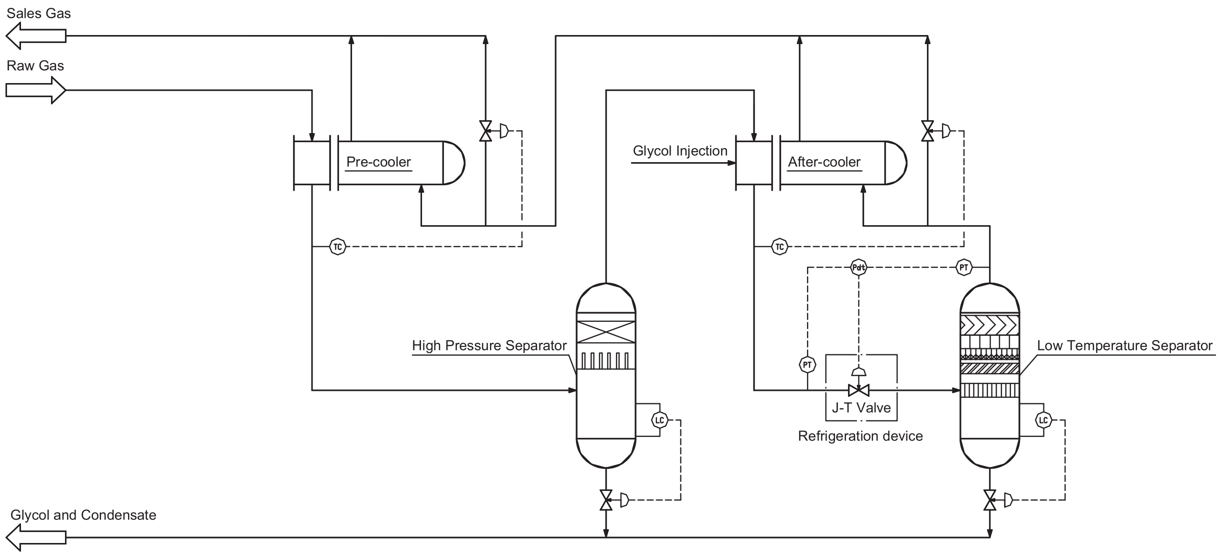


**Fig. 1.** Conventional Low Temperature Separation (LTS) process flow diagram. Note: J-T valve, expander and propane refrigeration system are used as refrigeration device for decrease the natural gas temperature.

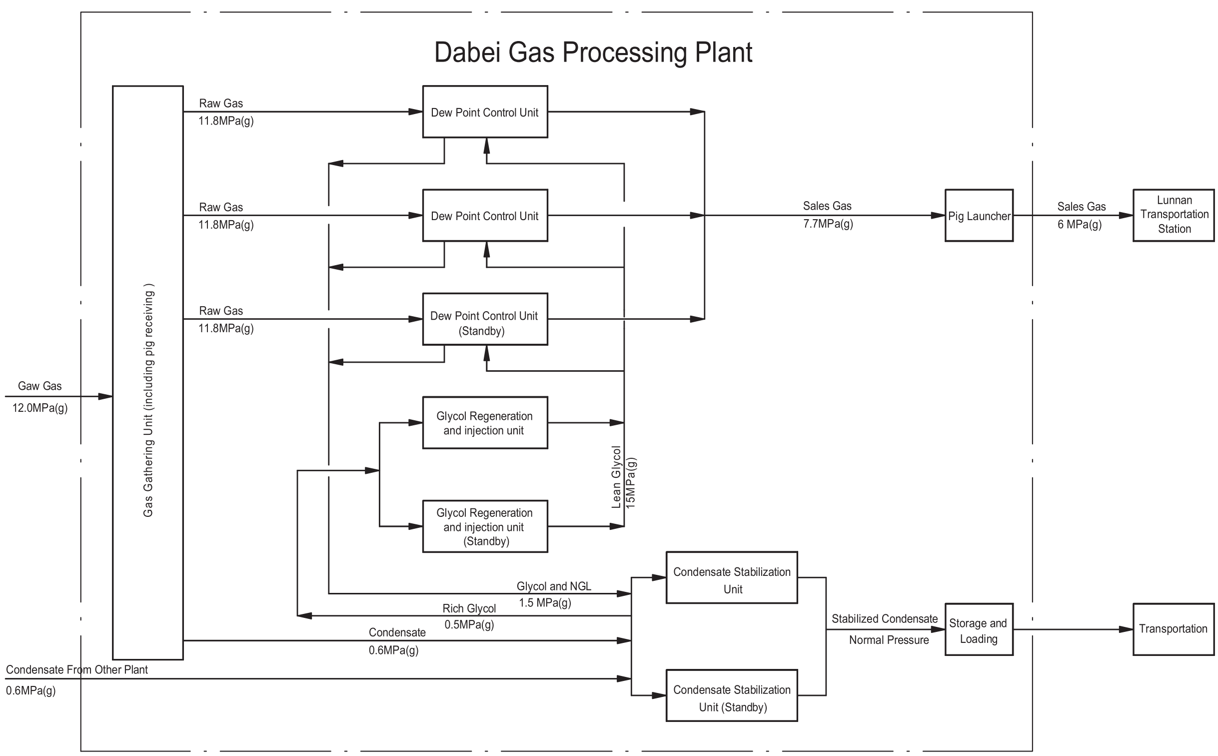


**Fig. 2.** The overall process flow diagram.

no $H_2S$. The water and hydrocarbon dew points of product gas shall meet the transportation requirement that is no more than −5 °C at 6 MPa (gage pressure) which is the operating pressure of Lunnan gas transmission initial station of China's West-East natural gas transmission.

In the overall process scheme of Figure 2, the raw gas from the gathering system which transports the raw gas from the well pad to the GPP is initially separated in the gas gathering unit, and then the raw gas enters the dew point control unit to remove the water and heavy

**Table 2.** Raw gas (inlet of dew point control unit) conditions and composition.

| Property | Value | Property | Value |
|---|---|---|---|
| Molar Flow, kgmol/h | 17,618 | $nC_{14}$ | 0.003393 |
| Pressure, MPa.g | 12 | $nC_{15}$ | 0.000622 |
| Temperature, °C | 40 | $nC_{16}$ | 0.000067 |
| Composition, v% | | $nC_{17}$ | 0.000376 |
| $C_1$ | 99.204298 | $nC_{18}$ | 0.000160 |
| $C_2$ | 0.001260 | $nC_{19}$ | 0.000090 |
| $C_3$ | 0.001865 | $nC_{20}$ | 0.000015 |
| $iC_4$ | 0.001804 | $nC_{21}$ | 0.000003 |
| $nC_4$ | 0.001317 | $nC_{22}$ | 0.000008 |
| $iC_5$ | 0.039149 | $nC_{23}$ | 0.000003 |
| $nC_5$ | 0.030363 | $nC_{24}$ | 0.000001 |
| $nC_6$ | 0.047091 | $nC_{25}$ | 0.000002 |
| $nC_7$ | 0.101977 | $nC_{26}$ | 0.000002 |
| $nC_8$ | 0.039671 | $nC_{27}$ | 0.000000 |
| $nC_9$ | 0.005070 | $nC_{28}$ | 0.000002 |
| $nC_{10}$ | 0.001077 | $nC_{28}$ | 0.000000 |
| $nC_{11}$ | 0.001789 | $CO_2$ | 0.0978 |
| $nC_{12}$ | 0.001455 | $N_2$ | 0.0068 |
| $nC_{13}$ | 0.001478 | $H_2O$ | 0.090218 |

hydrocarbon components, after dehydration and de-hydrocarbon, the natural gas is transported as sales gas. The gas condensate from the gas gathering unit and the glycol–NGL mixture from the dew point units are sent to condensate the stabilization unit to be treated, respectively. The glycol liquid separated from the stabilization unit flows to the glycol regeneration units to be regenerated and finally injected into the raw gas, and the stabilized condensate is delivered to the storage system for storage and transportation.

To better investigate the thermodynamic behavior of the new proposed process, phase equilibria calculations have been performed by means of Aspen HYSY V11, relying on the Peng-Robinson (PR) Equation of State. PR Equation is widely used for simulation of hydrocarbons due to that it has been calibrated by many practical data in oil and gas engineering field by *Aspen Company*. The correction factors are injected in HYSYS system. Also, the reliability of the selected simulation framework has been already validated through a comparison with the operation data. According to the parameters of inlet gas, the dew point of hydrocarbon is about 78 °C and the water dew point is 40 °C at 6 MPa.g, both of which fail to meet the product specifications and thus the raw gas should be treated by dehydration and de-hydrocarbon process. Also, it is in the state of retrograde condensation according to its properties because the heavy hydrocarbon liquid flow does not change with the temperature decrease at 12 Mpa.g and the variation trend is shown in Figure 3.

Considering the high pressure of raw gas (12 MPa.g, 40 °C), the J-T valve throttling is employed to refrigerate (shown in Fig. 1). The separating temperature is determined by the delivery pressure and dew point requirement of sales gas. For sake of the specification, the hydrocarbon dew point shall be no more than −5 °C at 6 MPa. g, and the relation between the dew point and the refrigeration temperature is shown Figure 4. Considering the sales gas pressure and separating efficiency of low temperature separator, the separating pressure is identified as 7.8 MPa.g and the refrigerating temperature is determined as −26.5 °C. Regarding the LTS process, after the precooler the temperature of raw gas is about 20 °C, and the temperature of raw gas is cooled to −10 °C by the aftercooler, and then the raw gas gets through the J-T valve to decrease the pressure to 7.8 MPa.g and temperature to −26.5 °C. Finally, the low temperature stream enters the low temperature separator to remove the water and heavy hydrocarbons.

The GPP went into operation in 2020, and the blockage in the low temperature separator and J-T valve occurred after commissioning for 2 weeks. It resulted in high pressure

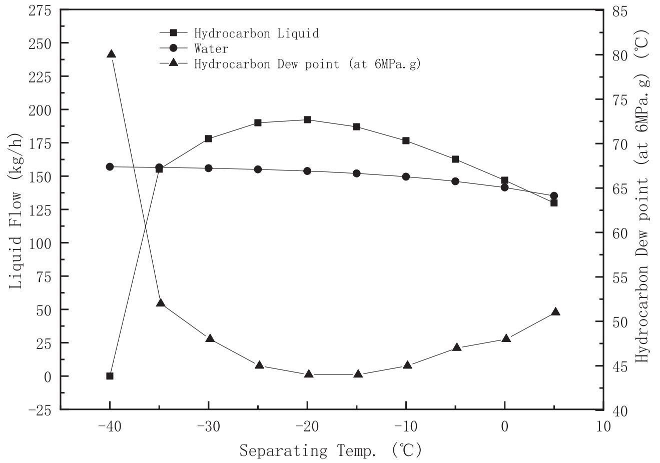


**Fig. 3.** Variation of heavy hydrocarbon liquid flow at different separating temperatures.

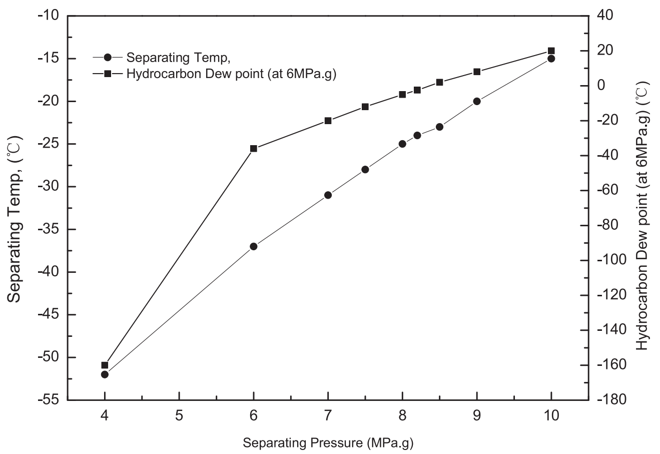


**Fig. 4.** The relation between the dew point and the separating (refrigeration) temperature.

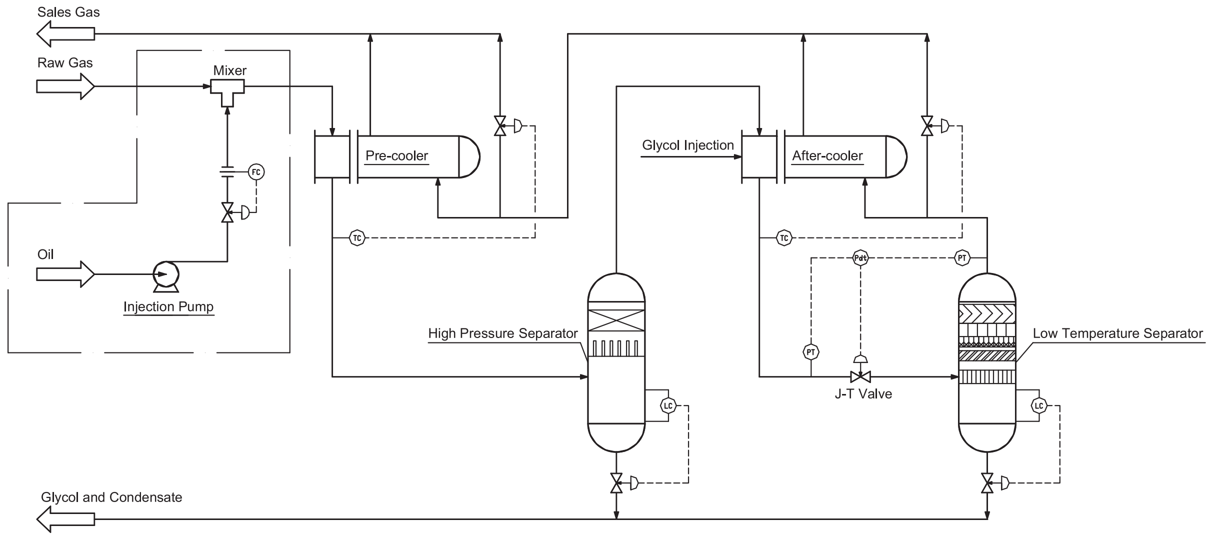


**Fig. 5.** Oiling Low Temperature Separation (LTS) process flow diagram. Pre-cooler: Operation pressure: 12 MPa.g; Operation temperature: inlet temp. 38 °C, outlet temp. 20 °C; After-cooler: Operation pressure: 11.9 MPa.g; Operation temperature: inlet temp. 20 °C, outlet temp. −10 °C; High pressure separator: Operation pressure: 12 MPa.g; Operation temperature: 20 °C; Low Temperature separator: Operation pressure: 7.9 MPa.g; Operation temperature: −26 °C.

drop in low temperature separator and affected the separation efficiency. As a result, the sales gas is disqualified and the GPP had to shut down for overhaul. According to the analysis and assay, the plugging matter was made of $C_{16}$ to $C_{30}$ which was called as wax.

### 2.2 Theoretical analysis of blockage in low temperature separator

Process simulation software packages are extensively used nowadays to estimate the product efficiency and enhance the performance of the system by optimizing operating parameters. For analyzing the reason for blockage in the LTS process, simulation is carried out by Aspen HYSYS, and the Peng-Robinson (PR) equation of state was used for modeling.

It is excluded that the gas hydrate creates this clog through the process simulation. The glycol is injected into the head bonnet of the after-cooler as hydrate inhibitor to mix with the natural gas for decreasing the hydrate formation temperature. The hydrate formation temperature drops from 17 °C to −28 °C after injecting the glycol, and it makes sure that the hydrate blockage can't take place during the operation. Another reason for blockage is the wax and the wax formation temperature of raw gas is 5.4 °C according to the simulation. Meanwhile the raw gas is in the state of retrograde condensation, the heavy hydrocarbon cannot be separated by the high pressure separator at 12 MPa.g and 20 °C, but it will deposit with the decrease of pressure and temperature. As a result, it blocks the J-T valve and the highly efficient internal of low temperature separator.

### 2.3 Improved low temperature separation process

To ensure the safety operation and economize the investment and OPerating EXpense (OPEX), the oiling LTS process is introduced to reduce the wax precipitation temperature to prevent the wax blockage during the low temperature separation. The process flow diagram of the oiling LTS process is illustrated in Figure 5.

This process has a good performance for this kind of natural gas which is composed with methane ($CH_4$) and a small quantity of heavy hydrocarbons. Compared with the conventional LTS process, oiling low temperature separation process has the following advantages:

1. Oil is employed as the wax blockage inhibitor to reduce the wax formation temperature for avoiding blockage in LTS process.
2. It takes low investment and simple operation.

## 3. Theoretical simulation analysis and practical application

The objective of this study is to find the most optimal injection stream and position to validate the applicability of the improved process, as well as to carry out the deviation between the simulation and operation.

### 3.1 Impact of injection stream composition

In order to work out the best injecting stream, the injection point is considered at the inlet pipeline of pre-cooler and

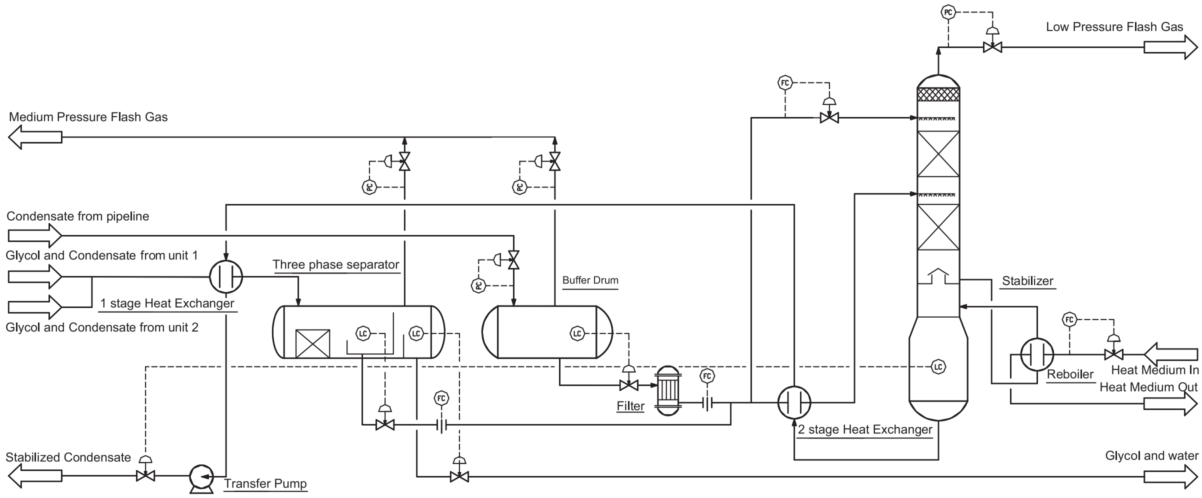


**Fig. 6.** The condensate stabilization unit flow diagram. Three-phase separator: Operation pressure: 1.5 MPa.g; Operation temperature: 50 °C; Buffer drum: Operation pressure: 1.0 MPa.g; Operation temperature: 50 °C; Stabilizer: Operation pressure: 0.4 MPa.g; Operation temperature: 140 °C (bottom of stabilizer), 83 °C (top of stabilizer).

three streams are selected as the injection oil, which are the stabilized condensate from the bottom of stabilizer which is in the condensate stabilization unit, the hydrocarbon liquid from the three-phase separator, the low temperature glycol and hydrocarbon liquid mixture from low temperature separator, based on the Dabei GPP raw gas condition and the overall process scheme. The three oil streams are separately injected into the raw gas as wax blockage inhibitor for investigation.

Within this context, according to the results of simulation, the wax precipitation temperature of the inlet/outlet gas of high-pressure separator, the wax content of the inlet/outlet gas of high pressure separator, the condensate flow from high pressure separator and low temperature separator were considered as the main performance index to compare for working out the optimal wax blockage inhibitor. For better illustrating the compositions of the injection stream, the condensate stabilization unit process flow diagram is shown in Figure 6.

In Figure 6, the raw condensate from the two dew point control units arrives at the 1st heat exchanger to be preheated by exchanging heat with stabilized condensate; subsequently the condensate enters the three-phase separator to remove the flash gas, glycol and water. Another raw condensate stream from the pipeline flows into the buffer drum and the filter in sequence. The two condensate streams are firstly mixed together and then part of them are heated in the 2nd stage heat exchanger before entering into the middle of stabilizer which is operated at approximately 150–500 kPa.g. Stabilized condensate from the bottom of column sequentially arrives at the 2nd and 1st heat exchanger to preheat the raw condensate, prior to its storage or transportation. The flash gas flows (medium pressure and low pressure) into the fuel gas system.

For better illustrating the different impacts, the three kinds of stream are named as: stream 1, from the bottom of the low temperature separator with more methane and glycol; stream 2, from the three-phase separator with more light hydrocarbons; stream 3, from the bottom of stabilizer with more heavy hydrocarbons. The compositions of three streams are shown in Table 3. The variation of wax precipitation temperatures of the inlet/outlet gas of high-pressure separator with the injected volume are depicted in Figure 7.

From Figure 7, it is evident that the wax precipitation temperature decreases with the increase of injected flow. The stream 2 has the strongest effect in the inlet/outlet gas of the high-pressure separator, indicating that it is likely the most effective ingredient for reducing the wax precipitation temperature. Comparing the three streams' compositions, with the same injected volume, the stream 2 has the highest glycol concentration and presents the worst performance, which means that glycol has a negligible effect on reducing the wax precipitation temperature. When comparing the stream 2 and 3, the former has better performance than the latter, because the content of light hydrocarbon components, such as $C_3$ to $C_8$, of stream 2 is higher than that of stream 3. From this point, the light hydrocarbon components have better performance on enriching the natural gas, because the partial pressure of light hydrocarbon in natural gas is higher than that of heavy ones when it reaches the gas-liquid equilibrium state. Considering the injecting flow of 1150 kg/h, the variation of the dew point and hydrate formation temperature when the different streams are injected into the raw gas is shown in Figure 8. The gas-liquid equilibrium changes with the heavy components increase, and the dew point and the hydration formation temperature of raw gas increase with the concentration of heavy components increase. As the stream 2 contains glycol, the hydration formation temperature obviously reduces from −18 °C to 0 °C, and the others have a slight increase. In theory, the stream 1 contains more light hydrocarbons than stream 2, but it also contains glycol and water

**Table 3.** The conditions and composition of three streams.

| Property | Stream 1 | Stream 2 | Stream 3 |
|---|---|---|---|
| Injection Pressure, MPa.g | 14 | 14 | 14 |
| Temperature, °C | −26 | 50 | 40 |
| Composition, v% | | | |
| $C_1$ | 32.256256 | 6.717451 | 0.203337 |
| $C_2$ | 0.001997 | 0.001355 | 0.000277 |
| $C_3$ | 0.009154 | 0.011521 | 0.004614 |
| $iC_4$ | 0.019130 | 0.031552 | 0.016685 |
| $nC_4$ | 0.019550 | 0.034414 | 0.019395 |
| $iC_5$ | 1.195913 | 2.370400 | 2.310176 |
| $nC_5$ | 1.230689 | 2.494358 | 2.491781 |
| $nC_6$ | 4.505280 | 9.830542 | 10.512421 |
| $nC_7$ | 18.547487 | 43.435303 | 47.325529 |
| $nC_8$ | 10.684464 | 26.948665 | 29.329368 |
| $nC_9$ | 1.576218 | 4.157094 | 4.602200 |
| $nC_{10}$ | 0.325147 | 0.865807 | 0.432213 |
| $nC_{11}$ | 0.452022 | 1.177213 | 0.622972 |
| $nC_{12}$ | 0.286314 | 0.723253 | 0.409374 |
| $nC_{13}$ | 0.179356 | 0.434324 | 0.267621 |
| $nC_{14}$ | 0.217736 | 0.512803 | 1.270121 |
| $nC_{15}$ | 0.026852 | 0.062272 | 0.041619 |
| $nC_{16}$ | 0.001553 | 0.003627 | 0.012891 |
| $nC_{17}$ | 0.010105 | 0.023568 | 0.080771 |
| $nC_{18}$ | 0.001889 | 0.004323 | 0.017460 |
| $nC_{19}$ | 0.000820 | 0.001879 | 0.007708 |
| $nC_{20}$ | 0.000074 | 0.000169 | 0.000721 |
| $nC_{21}$ | 0.000007 | 0.000016 | 0.000073 |
| $nC_{22}$ | 0.000014 | 0.000032 | 0.000139 |
| $nC_{23}$ | 0.000004 | 0.000008 | 0.000036 |
| $nC_{24}$ | 0.000002 | 0.000004 | 0.000016 |
| $nC_{25}$ | 0.000002 | 0.000004 | 0.000017 |
| $nC_{26}$ | 0.000002 | 0.000005 | 0.000020 |
| $nC_{27}$ | 0.000000 | 0.000000 | 0.000000 |
| $nC_{28}$ | 0.000004 | 0.000009 | 0.000032 |
| $nC_{29}$ | 0.000001 | 0.000003 | 0.000011 |
| $nC_{30}$ | 0.000000 | 0.000001 | 0.000000 |
| Glycol | 13.024883 | 0.003077 | 0.001627 |
| $H_2O$ | 15.215747 | 0.078411 | 0.012078 |
| $N_2$ | 0.026792 | 0.002145 | 0.000001 |
| $CO_2$ | 0.184535 | 0.074389 | 0.006695 |

which impair the performance of decreasing the wax precipitation temperature. In other word, the stream 2 needs more injected flow to reduce the wax precipitation temperature under the same conditions. For avoiding the wax deposit during the LTS process, the wax precipitation temperature is determined to be −31.5 °C for ensuring that the wax precipitation temperature is 5 °C lower than the refrigerating temperature −26.5 °C, so the injected flow of stream 2 is determined as 1150 kg/h according to the Figure 7.

After mixing with the oil, the compositions of raw gas change as shown in the Figure 9, which illustrate the variation of the heavy compositions ($C_{10}$–$C_{30}$). The wax content of inlet gas of high-pressure separator increases linearly with the injection volume increase, which means that the three streams can make the raw gas richer than before. It presents that the wax content of outlet gas of high-pressure separator adds linearly with the injection volume increase by mixing the stream 1 and 2, nevertheless

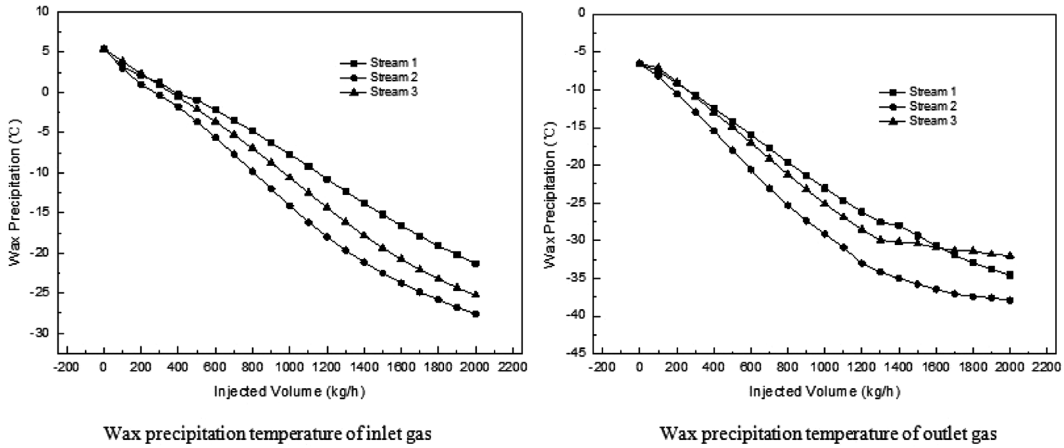


**Fig. 7.** The variation of wax precipitation temperature of the inlet/outlet gas of high-pressure separator with the injected volume.

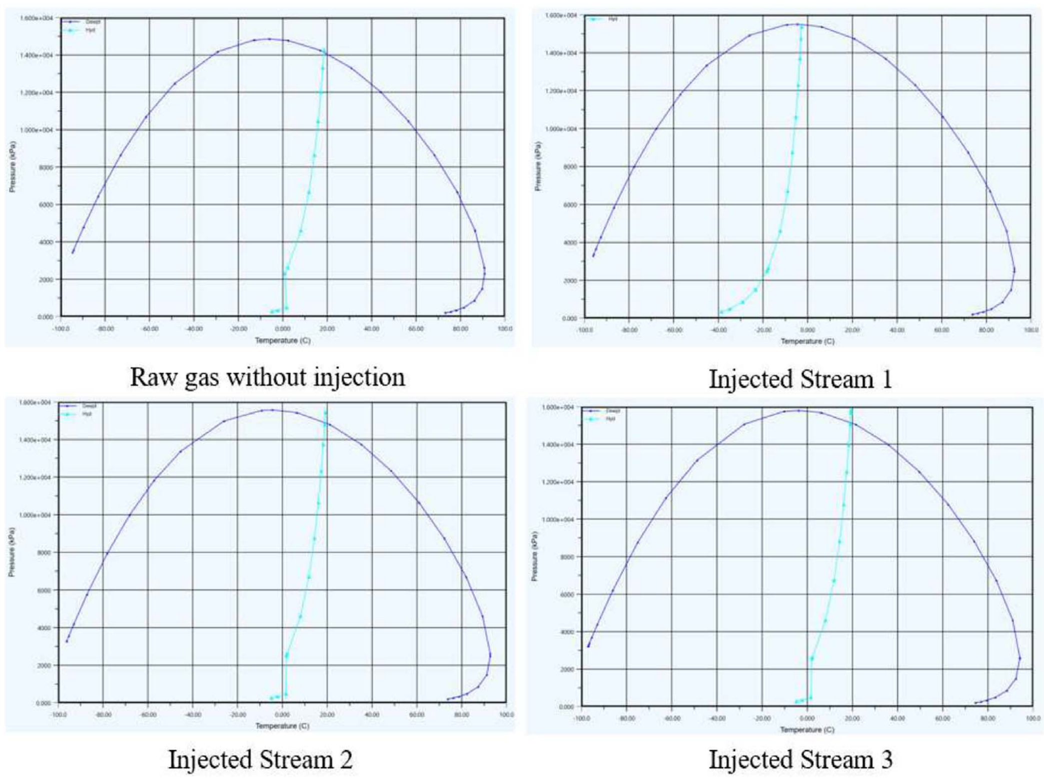


**Fig. 8.** The variation of gas-liquid equilibrium of raw gas with injecting stream.

the wax content drops slowly with the injection increase by mixing the stream 3. It illustrates that the heavy hydrocarbon components are separated by high pressure separator after mixing stream 3. The condensate flow which is separated from high pressure separator increases with the injection volume increase as a linear relation (see Fig. 10). The condensate flow of the high-pressure separator with mixing the stream 1 is largest than others, and it means that most of glycol liquid is separated by high pressure separator. So the condensate separated from low temperature separator is less than others (see Fig. 10). Analyzing these figures, the effective components are hydrocarbons, special light hydrocarbons which can reduce the wax precipitation temperature of natural gas for effectively opposing blockage, while the glycol is only applied to drop the hydrate formation temperature for avoiding the hydrate formation.

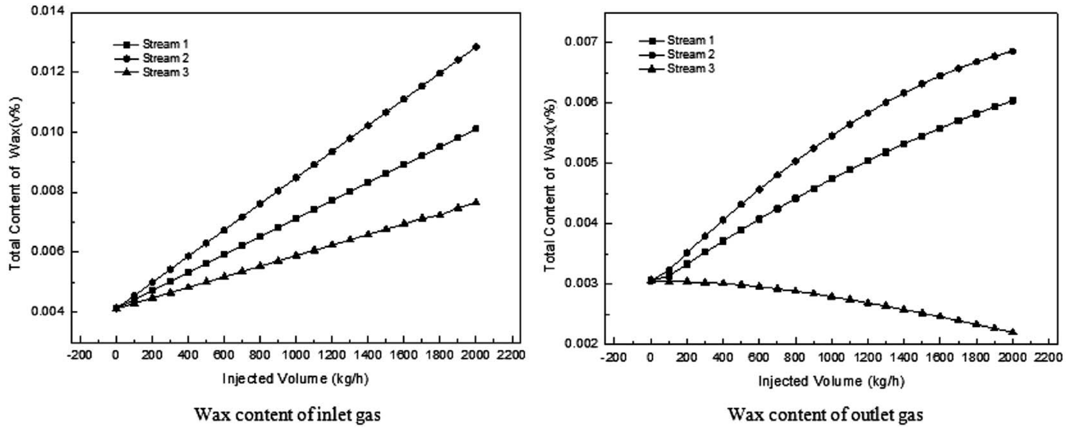


**Fig. 9.** The variation of wax content of inlet/outlet gas of high-pressure separator with the injected volume.

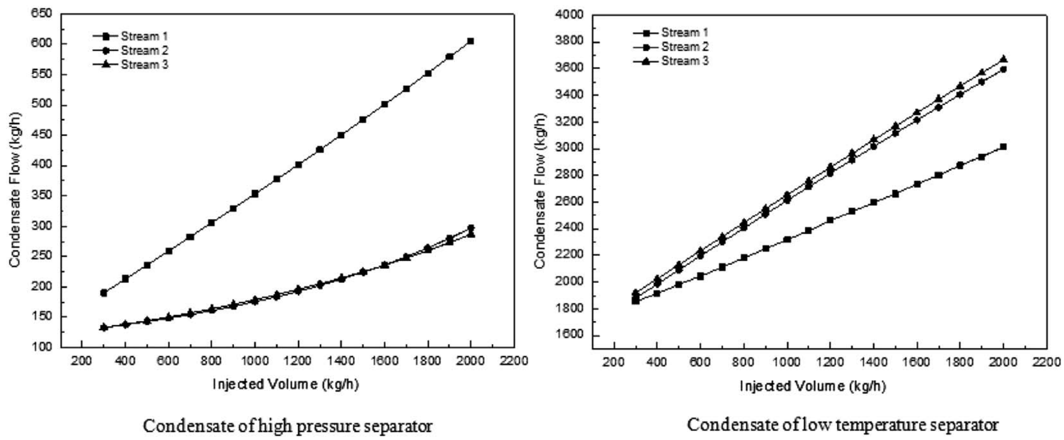


**Fig. 10.** The variation of condensate flow of high-pressure separator/low temperature separator with the injected volume.

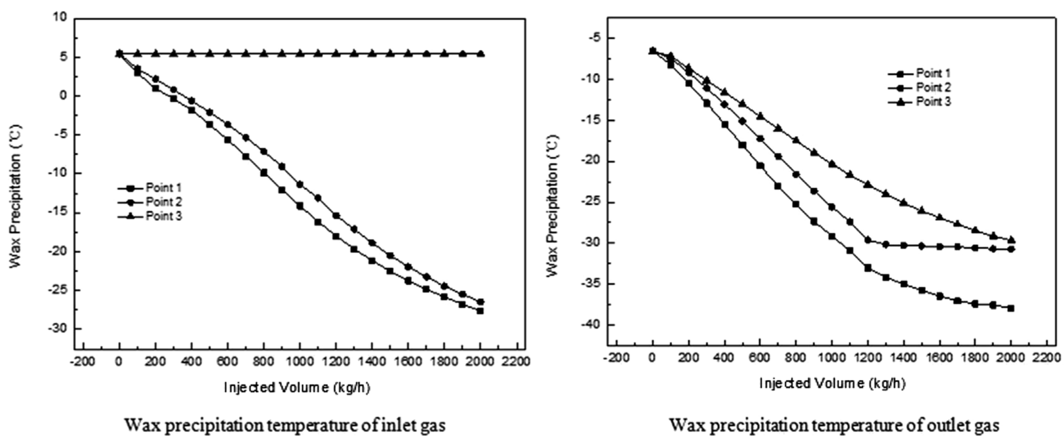


**Fig. 11.** The variation of wax precipitation temperature of the inlet/outlet gas of high-pressure separator with the injected volume.

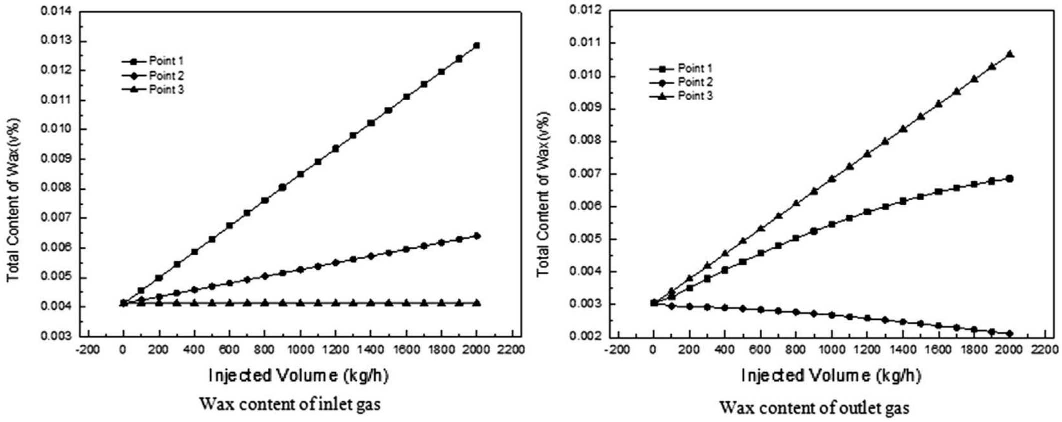


**Fig. 12.** The variation of wax content of inlet/outlet gas of high-pressure separator with the injected volume.

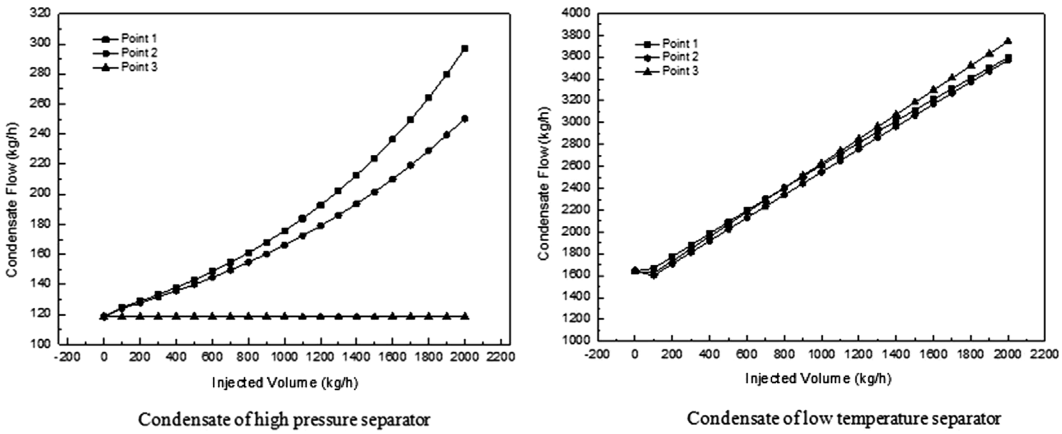


**Fig. 13.** The variation of condensate flow of high-pressure separator/low temperature separator with the injected volume.

Comprehensive view, the stream 2 hydrocarbon from the three-phase separator is the best choice as the wax blockage inhibitor.

## 3.2 Impact of injection point position

The injection point is another important key of this new process. Similarly, based on the Dabei GPP raw gas parameters, the stream 2 is considered as the injection stream and three points are simulated and compared, which are the inlet pipeline of pre-cooler, the inlet pipeline of high-pressure separator and outlet pipeline of high-pressure separator.

In this section, the wax precipitation temperature of the inlet gas of pre-cooler and the outlet gas of high-pressure separator, the wax content of the inlet/outlet gas of high pressure separator, the condensate flow from high pressure separator and low temperature separator are considered as the main performance index to work out the best injection point and the comparisons are shown in Figures 11–13.

All of the indexes change with the injection flow increase, including the wax precipitation temperature decreases with injection flow increase, including the wax contents and condensate from the bottom of the high separator increase with adding of the injection flow excepting the point 3. Because the injection point 3 is after the high pressure separator, the wax precipitation temperature, the wax content of inlet gas of high pressure separator and the condensate from the bottom of high pressure separator have no change with the injection flow increase. Figures 11–13 present that the mixing temperature (the raw gas temperature) affects the oiling performance for LTS process. The high mixing temperature is beneficial to enrich the raw gas, for instance, the point 1 is at the inlet pipeline of pre-cooler for which the operating temperature is about 40 °C and the temperature of point 2 and 3 is about 20 °C. Hence the wax precipitation temperature of point 1 is lower than those of others (see Fig. 11) and the heavy components of point are more than others (see Fig. 12). The point 3 is at the outlet pipeline of high-pressure separator and the heavy components are not separated

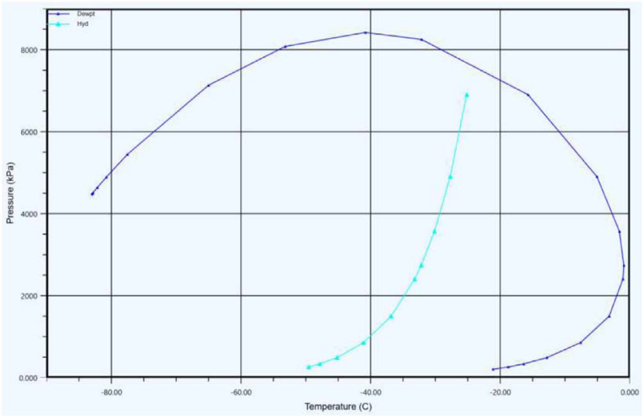


**Fig. 14.** The gas-liquid equilibrium of sales gas.

after mixing into raw gas, so its wax contents is highest at the three points (see Fig. 13). Meanwhile, it does not make much of difference in the condensate flow from the bottom of high-pressure separator. Overall, it can be concluded that the point 2 is considered as the best injected position for oiling LTS process.

### 3.3 Comparison between simulation and practical operation

According to the simulation results, the oiling process was utilized for dew point control unit in Dabei GPP. The hydrocarbon liquid from three phase separator as the wax blockage inhibitor with 1150 kg/h was injected into the raw gas at the inlet pipeline of pre-cooler. After 4 weeks, the blockage cannot be found in the J-T valve and low temperature separator and the pressure drop of low temperature separator was no more than 50 kPa. Both of the hydrocarbon dew point (−7 °C at 6 MPa.g) and water dew point (−25 °C) of sales gas meet the requirement of national standard and transportation. The gas-liquid equilibrium of sales gas is presented in Figure 14.

For carrying out the deviation between the simulation and operation, five groups of operating data are obtained from the DCS and daily analysis of the dew point control unit. The comparisons between simulation and operation in aspects of the total content of wax of the inlet/outlet gas of the high-pressure separator, condensate from bottom of high-pressure separator and low temperature separator were conducted, as shown in Figures 15 and 16.

The results show that the oiling injection process has good performance for avoiding wax blockage. The parameters of simulation are remarkably close to those from actual operation, which means the simulation results are reliable and the Aspen HYSYS software is suitable for simulation of phase equilibrium.

## 4 Economic assessment

The preliminary economic evaluation of oiling LTS process is presented in the following section. Firstly, the size

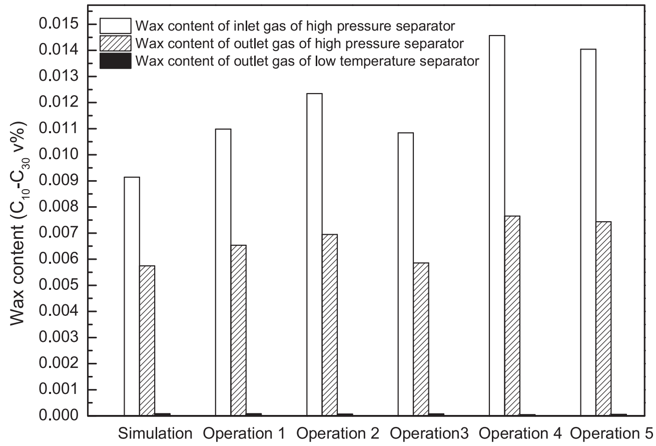


**Fig. 15.** The comparison of wax content between simulation and operation.

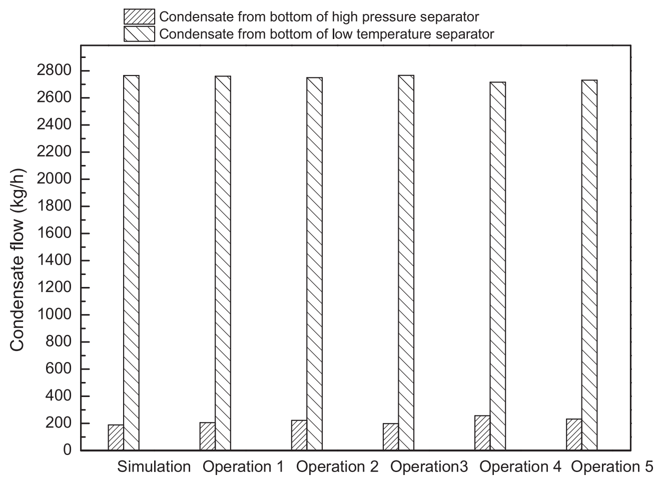


**Fig. 16.** The comparison of condensate between simulation and operation.

calculations of the vessels are performed in Table 4; and then the process equipment costs, and other capital investments are listed in Tables 5 and 6 while the OPEX is listed in Table 7; finally, an economic analysis is presented.

Tables 5 and 6 show the size and cost of equipment of the dew point control unit based on the design parameters introduced in the previous section. As the $Cl^-$ concentration of producing water is about 12 × $10^4$ mg/L, 316 stainless steel is selected as material for pre-cooler and high pressure separator. Considering the low separating temperature, 316 L stainless steel is used for after-cooler and low temperature separator. Electricity is from the state grid cooperation of China, and water is from water station, also the instrument gas and other utilities are from GPP. The price of electricity and water are computed from actual operation and the sales gas price refers international NYME natural gas price in 2021. As it can be seen, the dew point control unit requires around 8,471,671$ investment, however this value is preliminary, and it has some differences from actual costs.

**Table 4.** Process equipment size.

| Name | Diameter (m) | High (m) | Volume ($m^3$) |
|---|---|---|---|
| Pre-cooler (fixed tube sheet type) | 1 | 12 | 9.42 |
| High pressure separator (vertical type) | 1.4 | 4.4 | 6.77 |
| After-cooler (fixed tube sheet type) | 1.2 | 16 | 18.09 |
| J-T valve | / | / | / |
| Low temperature separator (vertical type) | 1.6 | 4.4 | 8.84 |
| Injection pump (metering pump) | / | / | / |

Note: The pre-cooler is formed of two identical fixed tube sheet type heat exchangers and the size shown in the table is only for single one. The after-cooler is same as pre-cooler.

**Table 5.** Process equipment (fixed) costs for each dew point control unit

| Type | Equipment | No. | Unit price $ (2021) | Total Price $ (2021) |
|---|---|---|---|---|
| Vessel | High pressure | 1 | 245,048 | 245,048 |
| | Low temperature separator | 1 | 259,140 | 259,140 |
| Heat exchanger | Pre-cooler | 1 | 1,015,227 | 1,015,227 |
| | After-cooler | 1 | 1,327,605 | 1,327,605 |
| Pump | Injection pump | 2 | 68,065 | 136,130 |
| Special valve | J-T valve | 2 | 237,605 | 475,210 |
| | Sum | | | 3,458,360 |

Note: Injection pump and J-T valve is one use and one standby.

**Table 6.** Capital investments for each dew point control unit

| Items | Cost $ (2021) |
|---|---|
| Process Equipment and installation | 3,458,360 |
| Valves, Piping and installation | 2,565,307 |
| Instrumentation and installation | 2,037,326 |
| Electricity and installation | 91,476 |
| Civil | 319,202 |
| Total cost | 8,471,671 |

**Table 7.** Operating costs and product profit for each dew point control unit

| Items | Quantity | Unit price | Total cost/profit ($ per annum 350 days) |
|---|---|---|---|
| Water (treated) | 100 ($m^3$/per annum) | 0.8 ($/$m^3$) | 80 |
| Electricity | 264 (kWh/day) | 0.12 ($/kWh) | 11,088 |
| Manpower | Unification consideration of the GPP | – | – |
| Depreciation | 10% (Based on 10 years) | – | 847,167 |
| Maintenance | 2% of capital cost | – | 169,433 |
| Sum | | | 1,027,768 |
| Sales gas | 498.69 ($10^4$ $m^3$/day)<br>LHV: 34.26 MJ/$m^3$<br>HHV: 38 MJ/$m^3$ | 5.085 ($/MMBtu)<br>NYMEX natural gas (Nov. 2021) | 288,205,036 |

## 5 Conclusion

Oiling LTS process is a kind of improvement and innovation process based on conventional dehydration and de-hydrocarbon ones. It can be used to remove hydrocarbon and water without any blockage. The composition of the injected oil and injection point are optimized. The advanced process is employed in practical GPP and the comparisons between simulation and practical operation, as well as the economic analysis are conducted.

The conclusions can be drawn out and summarized below:

1. The oiling LTS process presents a good performance in respect of reducing the wax precipitation temperature for avoiding the blockage during the operation.
2. The new process can change the composition of raw natural gas that makes more hydrocarbon liquid is separated in the high-pressure separator when the gas is in the state of retrograde condensation.
3. The hydrocarbon liquid from the three-phase separator has a good effect as wax precipitation inhibitor for the new process when it is injected into the natural gas at the inlet pipe of the pre-cooler.
4. The hydrate inhibitor such as glycol has a negligible effect on the wax precipitation temperature.
5. From the comparison between the simulation and practical data, it can be seen that the simulation results are reliable.
6. The dew point control unit with 0.5 BCM per year of capacity requires around 8,471,671$ investment. The adding investment of the oiling LTS process is around 423,583$ accounting on 5% total capital cost of this unit.

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